Simon Judd

THE MBR BOOK

Principles and Applications of Membrane Bioreactors in Water and Wastewater Treatment

SECOND EDITION

Dedication

Once again, for Oliver and Samuel. And also for our family – Ivor and Margaret, Lorna, Ciss, Robert and Jane, Daisy and Heyes, John and Patricia, Lucy, Cameron and Dynamite.

The MBR Book

Principles and Applications of Membrane Bioreactors for Water and Wastewater Treatment

Second edition

Edited by
Simon Judd
Claire Judd

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Preface

This is only the second edition of *The MBR Book*, the first edition having been published in 2006, but it's the fourth on membrane technology from the Centre for Water Science at Cranfield University in the United Kingdom. The first of these was the original book on membrane bioreactors: *Membrane Bioreactors for Wastewater Treatment* by Tom Stephenson, Simon Judd, Bruce Jefferson and Keith Brindle, which came out in 2000 (IWA Publishing). This was followed in 2003 by *Membranes for Industrial Wastewater Recycling and Reuse*, by Simon Judd and Bruce Jefferson (Elsevier). Since then there have been a few books dedicated to membrane technology for wastewater treatment, three of which were all published in 2006: *Membrane Systems for Wastewater Treatment* (WEFPress, 2006), *Membrane Technology for Waste Water Treatment* (Johannes Pinnekamp and Harald Friedrich, FiW-Verlag, 2006) and *The MBR Book* (Elsevier, 2006). As a poignant demonstration of history repeating itself, the publication year of the second edition is the same as that of two other wastewater membrane reference texts: *MBR Practice Report: Operating Large Scale Membrane Bioreactors for Municipal Wastewater Treatment*, by Christoph Brepols (IWA Publishing) and *The Guidebook to Membrane Technology for Wastewater Reclamation*, led by Mark Wilf (Balaban Publishers). Membrane wastewater books are, it seems, like London buses.

There have also been many more books, both on biological treatment and membrane technology, which have included sections on MBRs. A comprehensive listing of these would be challenging. Two of the most recent, both from 2008, are *Biological Wastewater Treatment, Principles, Modelling and Design*, by Mogens Henze, Mark van Loosdrecht, George Ekama and Damir Brdjanovic from IWA Publishing, and *Advanced Membrane Technology and Applications*, by Norman Li, Tony Fane, Winston Ho and Takeshi Matsuura from Wiley (2008). However, there are several books which similarly aim to cover either membrane technology or biological treatment in a rather more comprehensive manner than provided in *The MBR Book*. Biological treatment texts include the biotreatment 'bible' of Metcalf and Eddy: *Wastewater Engineering – Treatment and Reuse* by George Tchobanoglous, Franklin Burton and David Stensel (McGraw Hill, 2003) and also the commendable *Biological Wastewater Treatment* by Leslie Grady, Glen Daigger and Nancy Love, the third edition of which is also due out in 2010 (IWA Publishing).

Writing the second edition of *The MBR Book* was initially viewed as being a simple enough task, with the format used for the first edition being serviceable enough, only requiring updates from the past four to five years. However, there has been an explosion of activity over this period; assurances to the publisher that this edition would not exceed 30% of the first have proven woefully under-conservative. In the intervening period the number of discernible MBR membrane products has more than doubled that of the first edition, and it is acknowledged that the 44-or-so membrane products identified and described cannot be considered comprehensive. The past five years have also seen some important landmark plants installed – up to 110 megalitres/day in capacity. Scientific studies of MBRs have continued to be published at much the same rate as ever – about 20% exponential growth each year since the mid-1990s. It is these developments that have contributed to a 45% expansion of the original text to produce the second edition.

As with first edition, the second edition of *The MBR Book* is set out in such a way as to segregate the science from the engineering, in an attempt to avoid confusing, irritating or offending anyone of either persuasion. The book is meant to include as much practical information as possible, whilst still covering the science and technology. There are five chapters, with the membrane and biological fundamentals covered in Chapter 2 along with most of the scientific studies. The commercial MBR membrane products are summarized in Chapter 4 and their application to wastewater treatment is described in Chapter 5; the information from Chapter 5 is compiled and used for the design section in Chapter 3. New to the second edition are, in Chapter 1, summaries of the status of the technology across 13 countries and a brief précis of research trends. Also, Chapter 3 has been completely redrafted to provide a cost modelling and cost benefit analysis method, as well as a section on operation and maintenance. The latter is considerably more extensive than in the first edition, and has been informed by an expert panel of practitioners. Extensive cross-referencing between sections and chapters, including figures or tables in other chapters, is employed to try to ensure a degree of coherence throughout the tome.

A list of symbols and a glossary of terms and abbreviations are included at the end of the book, and those relating specifically to the membrane technology are outlined in Appendix C as a preface to the commercial MBR membrane module specifications. However, since a few terms and abbreviations are more extensively used than others, and possibly not universally recognized, it is probably prudent to list these to avoid confounding some readers (see following table). It is acknowledged that resolution of the inconsistencies in the use of terms to describe the membrane component of MBR technologies has not been possible, specifically the use of the terms 'module' (see Appendix C) and 'fouling'. This is something which is to be addressed by the Water Environment Federation (and the best of luck with that one).

Term	**Meaning**
Common units	
MLD	Megalitres/day (thousands of cubic metres per day)
LMH	L/(m^2 h) (litres per square metre per hour)
Billion	1000 Million
Process configurations	
iMBR	Immersed (internal) MBR
sMBR	Sidestream (external) MBR
a-lsMBR	Air-lift sidestream MBR
anMBR	Anaerobic MBR
Membrane configurations	
FS	Flat sheet (plate-and-frame, planar)
HF	Hollow fibre
MT	Multitube
Fouling	
Reversible	Removed by physical cleaning, such as backflushing or relaxation
Irreversible	Not removed by physical cleaning but removed by chemical cleaning
Irrecoverable	Cannot be removed
Aeration	
SAD	Specific aeration demand, either with respect to the membrane area (SAD_m) or permeate flow (SAD_p)

Given the broad range of stakeholders encompassed, it is inevitable that inconsistencies in terminology, symbols and abbreviations have arisen. It is also certain that, despite the best efforts, the text includes a number of inaccuracies and omissions, for which the authors cannot be held liable. We have, naturally, done everything we could to ensure that the information presented is as accurate and complete as possible, but, notwithstanding this and because of the complex nature of the subject, interested parties are strongly advised to check facts and figures with the relevant organisations before acting on any information provided.

It would be remiss to preface this book without offering the most grateful and sincere thanks to the many contributors – more than 150 in total. These include product suppliers, technology providers, consultants, contractors, end users and academics. Almost all the practical operational data provided have been supplied by the technology providers, although corroboration of some information from end users has been possible in some cases. All information providers are listed in the following section and on the title page of each chapter, and their assistance, kindness and, at times, superhuman patience in responding to a plethora of detailed queries by the authors are gratefully acknowledged. Contributions have also come from academic staff and students – predominantly from Cranfield University in the United Kingdom. With regard to the latter, specifically most grateful thanks is offered to current students of, and recent graduates from, the Centre for Water Science and, in

particular (in alphabetical order), Harriet Fletcher, Wenjing Ma, Ignacio Martin, Ewan McAdam, Ana Santos and Bart Verrecht. Gratitude is similarly expressed to the incomparable Pierre Le-Clech from the University of New South Wales, who took on the unenviable task of updating the sections on membrane fouling behaviour in Chapter 2, and to the various members of the Department of Applied Mathematics, Biometrics and Process Control at Ghent University, who contributed to the modelling sections in Chapter 3.

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Finally, we would encourage readers to participate in one (or more) of the now several on-line forums dedicated to the discussion of membrane bioreactor technology, especially ours (The MBR Group – Membrane Bioreactors at www.linkedin.com).

As with any piece of work, the editors would welcome any comments from readers, critical or otherwise, and our contact details are included in the following section.

SJ and CJ

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Simon Judd is Professor in Membrane Technology at the Centre for Water Science at Cranfield University, United Kingdom, where he has been on the academic staff since August 1992. Since abandoning a chequered career in hairdressing, Simon has co-managed most of the biomass separation MBR programmes conducted within the School, comprising 15 individual research project programmes and encompassing 13 doctorate students dating back to the mid-1990s. He has been principal or co-investigator on three major UK Research Council-sponsored programmes dedicated to MBRs with respect to in-building water recycling, sewage treatment and contaminated groundwaters/landfill leachate, and is also Chairman of the Project Steering Committee on the multi-centred EU-sponsored EUROMBRA project. As well as publishing extensively in the research literature, Simon has co-authored three textbooks in membrane and MBR technology, and delivered a number of keynote presentations at international membrane conferences on these topics. He is the manager of *The MBR Group*, an online discussion forum on LinkedIn (www.linkedin.com).

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A number of individuals and organizations have contributed to this book, in particular to the product descriptions in Chapter 4 and the case studies referenced in Chapter 5. The editors would like to thank all contributors for their co-operation, sometimes at short notice, and acknowledge the particular contribution of the following (listed in alphabetical order, by organization and last name; websites, where listed, were accessed in July 2010):

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Chapter 1

Introduction

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1.1. DEFINITION

The term 'membrane bioreactor' (MBR) applies to all water and wastewater treatment processes integrating a permselective membrane with a biological process. All currently available commercial MBR processes employ the membrane ostensibly as a filter, rejecting the solid materials developed by the biological process to provide a clarified and disinfected product. It is this type of MBR, the biomass rejection MBR (Section 1.1), which forms the primary focus of this book. The progress of technological development and market penetration of MBRs can be viewed in the context of their historical development (Section 1.2), current market penetration (Section 1.3), key drivers

(Section 1.4) and the status of MBR research (Section 1.5), all impacting to some degree on the future prospects of the technology (Section 1.6).

1.2. HISTORICAL PERSPECTIVE

1.2.1. Membranes and Membrane Technology

The membrane industry did not exist until the early twentieth century; the main research on membrane separation phenomena was aimed at elucidating the physico-chemical principles of the process, and the mechanism of diffusion. However, some of these early-stage achievements still impact on the academic research and industrial applications today. These include Fick's (1855) phenomenological laws of diffusion, van't Hoff's (1887, 1888) osmotic pressure equation, for which he was awarded the first Noble Prize in Chemistry in 1901, and Thomas Graham's pioneering work in gas separation using both porous membranes and dense membranes is still relevant today. Graham discovered that rubber exhibits selective permeability to different gases, and also found low-molecular weight substances to be concentrated in the permeated gas when the membrane pore size is close to the mean free path of gas molecules (Graham, 1861, 1866). Graham's work was inspired by Schmidt's (1856) earlier study, where he had used bovine heart membranes (the pore dimension being 1−50 nm) to separate soluble *Acacia* − arguably the first documented ultrafiltration (UF) experiment.

The first synthetic UF membranes were prepared by Bechhold from collodion (nitrocellulose). Bechhold was also the first to measure membrane bubble points, and to propose the term 'ultrafilter' (Bechhold, 1907). Other important early researchers, Elford, Zsigmondy, Bachmann, and Ferry, etc., further developed Bechhold's membrane preparation method. Commercial application of collodion porous membranes can be attributed to Zsigmondy's laboratory at the University of Goettingen, Germany; Zsigmondy and Bachmann were the first to propose a method to produce porous collodion membrane in an industrial scale (Zsigmondy & Bachmann, 1918, 1922). Based on this technology, the world's first commercial microporous membrane supplier, Sartorius Werke GmbH, was established in Goettingen in 1925, although its products were mostly sold to research laboratories. The early porous collodion membrane formation method was named 'dry inversion', which is still in use today.

During World War II, damage to German distribution networks by bombing raids led to the development of techniques for rapid analysis for bacteria in water supplies. Using Sartorius membranes, Müller and others at Hamburg University developed an effective method to cultivate micro-organisms in drinking water. This was the first large-scale application of microfiltration (MF) membranes. Following on from this work and in recognition of the strategic importance of MF membranes, Alexander Goetz, a professor in the California

Institute of Technology, was sponsored by the US military to duplicate the Sartorius membrane technology. Goetz developed an improved membrane formation method, now called 'vapour-induced phase separation'. The main innovation of his method included using a copolymer of cellulose acetate and cellulose nitrate as the membrane material, and preparing the membrane in a high moisture environment. This technology was later transferred to Lowell Inc., and in 1954 Lowell established the Millipore Corporation to commercialise the membrane. This represents the incipient stages of the US microporous membrane industry.

The period between the 1960s and the 1980s is often regarded as being the golden age of membrane science. The crucial breakthrough was the development of the asymmetric cellulose acetate membrane by Loeb and Sourirajan in 1963 (Loeb & Sourirajan, 1964). Loeb and Sourirajan's membrane preparation method is often referred to as 'wet phase inversion' or 'non-solvent-induced phase separation' (NIPS). Microporous membranes prepared by this method have an asymmetric porous structure: a very thin surface microporous layer (~0.2 μm) supported by a substrate having larger pores. Because of its thin separation layer, the NIPS membrane demonstrates significantly improved fluxes.

The Loeb and Sourirajan membrane preparation method had a great influence on the development of reverse osmosis (RO), UF, MF and gas separation. Loeb and Sourirajan's goal was focused on producing high-flux RO membranes, but other researchers, particularly Alan S. Michaels, realized the general applicability of the technique. Michaels was the founder of Amicon Inc. In the 1960s, Amicon Inc. collaborated with Dorr-Oliver Inc. to develop new kinds of UF membranes prepared by using various polymers such as polyacrylonitrile (PAN), polysulfone (PS), poly(vinylidene difluoride) (PVDF) and others (Michaels, 1963), applying the new products on an industrial scale.

Thermally induced phase separation (TIPS) represents another important improvement in the development of membrane technologies. In TIPS, polymer and its diluents are mixed under high temperature to form a uniform solution. Gradually reducing the temperature of the casting solution causes phase separation and consequently a porous structure. The first commercial TIPS membrane may be attributed to Castro (1981). In the following two decades, TIPS membranes have been used in a variety of applications, such as blood plasma filtration, membrane distillation, fuel cells and medical dressings. Advantages of TIPS membranes include high porosity, high permeation rate, high physical strength, narrow pore size distribution and greater water fluxes than those of NIPS membranes: the pure water flux of typical TIPS MF membranes commonly exceeds 1000 L per m^2 membrane per hour per bar pressure (LMH/bar), compared with 200–300 LMH/bar for NIPS UF and MF materials. TIPS membranes typically used for MF are of 0.1–0.4 μm pore size.

Two other commercially important membrane production methods are the radiation track etched and melt extrusion and cold-stretching methods.

Radiation track etching was developed in the 1960s (Fleischer, Price, & Walker, 1969) with limited application in the manufacture of flat membrane due to its poor permeability and high cost. The melt extrusion and cold-stretching method, on the other hand, is much lower in cost. The method was first developed by Celanese Corp. in 1974 (Druin, Loft, & Plovan, 1974). In 1977, Mitsubishi Rayon Corp. produced a hollow-fibre (HF) polyethylene (PE) MF membrane by this membrane formation method. As an immersed membrane module, the HF PE MF membrane of Mitsubishi Rayon has found many applications in the field of wastewater treatment.

1.2.2. Membrane Bioreactor Technology

1.2.2.1. The Early Years: 1970s—1990s

The first membrane bioreactors (MBRs) were developed commercially by Dorr-Oliver in the late 1960s (Bemberis, Hubbard, & Leonardet, 1971), combining UF with a conventional activated sludge process (CASP), for application to ship-board sewage treatment (Bailey, Bemberis, & Presti, 1971). Other bench-scale membrane separation systems linked with a CASP were reported at around the same time (Hardt, Clesceri, Nemerow, & Washington, 1970; Smith, Gregorio, & Talcott, 1969). These systems were all based on what have come to be known as 'sidestream' configurations (sMBR, Fig. 1.1a), as opposed to the now more commercially significant 'immersed' configuration (iMBR, Fig. 1.1b). The Dorr-Oliver membrane sewage treatment (MST) process was based on flat-sheet (FS) UF membranes operated at what would now be considered excessive pressures (3.5 bar inlet pressure) and low fluxes (17 L/(m^2 h), or LMH), yielding mean permeabilities of less than 10 LMH/bar. Nonetheless, the Dorr-Oliver system succeeded in establishing the principle of coupling a CASP with a membrane to simultaneously concentrate the biomass whilst generating a clarified, disinfected product. The system was marketed in Japan under license to Sanki Engineering, with some success up until the early 1990s. Developments were also underway in South Africa which led to the

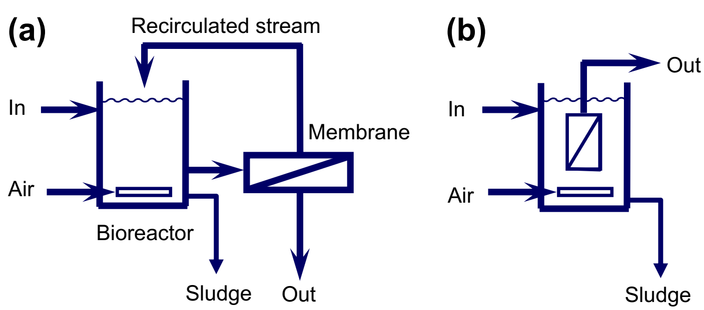

FIG. 1.1 Configurations of a membrane bioreactor: (a) sidestream and (b) immersed.

commercialization of an anaerobic digester UF (ADUF) MBR by Weir Envig (Botha, Sanderson, & Buckley, 1992), for use on high-strength industrial wastewaters.

At around this time, from the late 1980s to early 1990s, other important commercial developments were taking place. In Japan, the government-instigated water recycling programme prompted pioneering work by Yamamoto, Hiasa, Mahmood, & Matsuo (1989) to develop an immersed HF UF MBR process, as well as the development of an FS-microfiltration iMBR by the agricultural machinery company, Kubota (Section 4.2.1). This subsequently underwent demonstration at pilot scale, first at Hiroshima in 1990 (25 m^3/day, or 0.025 megalitres per day or MLD) and then at the company's own site at Sakai-Rinkai in 1992 (0.110 MLD). By the end of 1996, there were already 60 Kubota plants installed in Japan for domestic wastewater and, later on, industrial effluent treatment, providing a total installed capacity of 5.5 MLD. Also in Japan, Mitsubishi Rayon introduced its *SUR* MBR membrane module, based on its *Sterapore* product, in 1993.

Both these products to some extent displaced some of the older sidestream systems which had been established in Japan, though side-stream MBRs continue to be used in Japan and elsewhere. The installation of in-building wastewater recycling plants in Japan based on the Orelis Environment (formerly Rhodia Orelis and before this Rhône Poulenc) *PLEIADE*® FS sMBR system, actually pre-dates that of the Kubota plants for this duty. The *PLEIADE*® system was originally trialled in France in the 1970s and by 1999 there were 125 small-scale systems (all below 0.2 MLD) worldwide, the majority of these being in Japan and around a dozen in France. The Dorr-Oliver MST system was similarly rather more successful in Japan than in North America in the 1970s and 1980s (Sutton, Mishra, Bratby, & Enegess, 2002). Wehrle Environmental, part of the very well-established Wehrle Werk AG (formed in 1860) of Germany, has been applying its multitube (MT) sMBRs (predominantly employing Norit X-Flow polymeric MT membrane modules) to landfill leachate treatment since 1990. A sidestream MBR Degremont system based on ceramic membranes was introduced in the mid-1990s, and other ceramic membrane products have also been employed in a few sMBR applications. These pumped sidestream systems all tend to be used for industrial effluent treatment applications involving relatively low flows, such that their market penetration compared with the immersed systems, particularly in the municipal water sector, has been limited.

At around the same time as Kubota were developing their product, in the USA Thetford Systems were developing their *Cycle-Let*® process, another sidestream process, for wastewater recycling duties. Zenon Environmental, a company formed in 1980 and who subsequently acquired Thetford Systems, were developing an MBR system. By the early 1990s, the *ZenoGem*® immersed HF UF MBR process had been patented (Tonelli & Canning, 1993; Tonelli & Behmann, 1996), and the first immersed HF *ZeeWeed*® module, the

ZW145 which provided 145 square feet of membrane area, was introduced to the market in 1993 (Section 4.3.1). By the end of the Millennium the total installed capacity of Zenon plants had reached 150 MLD.

1.2.2.2. The Late 1990s Onwards: the Development of Other MBR Products

The first Kubota municipal wastewater treatment works installed outside Japan was at Porlock in the United Kingdom in 1997 (Section 5.3.1.1), following successful trials at Kingston Seymour by Wessex Water in the mid-1990s. The first Zenon membrane-based plant of similar size installed outside of the USA was the Veolia (then Vivendi) *Biosep*® plant at Perthes en Gatinais in France in 1999 (Section 5.3.1.1). Both these plants have a peak flow capacity just below 2 MLD, and represent landmark plants in the development and implementation of immersed MBR technology.

By the late 1990s, however, other MBR membrane products and systems were under development, leading to an explosion of commercial activity from the turn of the Millennium to the present day. Whereas the first half of the 1990s saw the launch of only three major immersed MBR membrane products, originating from just two countries (USA and Japan), the first five years of the following decade saw the launch of at least 10 products originating from seven countries, coupled with three significant acquisitions in the mid-noughties (Section 1.3). For 12 major suppliers (Table 1.1) as at 2010, there were either existing or planned MBR installations of more than 10 MLD capacity. In addition to those products listed for which there are 'flagship' large plants, there are currently at least another 33 MBR membrane products (Chapter 4), all of which have come to the market since around 2000, in addition to a number of proprietary MBR technologies based on a few of the membrane products.

1.3. MARKET

1.3.1. General

MBR systems have been implemented in more than 200 countries (Icon, 2008); growth rates and the extent of implementation vary regionally according to the state of economic development and infrastructure. Common to all regions, however, is the fact that sales of the technology have generally grown faster than the GDPs of countries installing them, significantly so in China, as well as more rapidly than the industries that use them (Srinivasan, 2007; BCC, 2008). Global growth rates between 9.5 and 12% are routinely quoted in reports produced by market analysis, and the market value of the MBR industry is predicted to approach \$0.5 billion (\$500 million) by 2013. Data taken from two sources for the period between 2000 and 2013 indicate a mean growth rate of 11.6–12.7% (Fig. 1.2). These data

TABLE 1.1 MBR Membrane Module Products, Bulk Municipal Market

Supplier	Country	Date launched	Acquired	Date, first >10 MLD plant
Asahi Kasei	Japan	2004	–	2007
GE- *ZeeWeed®*	USA	1993	Jun-06	2002
Korea Membrane Separation-*KSMBR®*	Korea	2000	–	2008
Koch Membrane Systems – *PURON®*	USA	2001	Nov-04	2010*
Kubota *EK*	Japan	1990	–	1999
Kubota *RW*	Japan	2009		–
Memstar	Singapore	2005	–	2010*
MICRODYN-NADIR	Germany	2005	–	2010*
Mitsubishi Rayon (*SADF*)	Japan	2005	–	2006
Mitsubishi Rayon (*SUR*)	Japan	1993		–
Motimo	China	2000	–	2007
Norit	Netherlands	2002	–	2010*
Siemens Water Tech. –*MEMCOR®*	Germany	2002	Jul-04	2008
Toray	Japan	2004	–	2010*

**Projected 2010 or 2011.*

also suggest that growth may be slow marginally in the period between 2010 and 2015 due to the global economic downturn. The difference in absolute values between the two studies reflects differences in assumptions made regarding eligible costs and income. It has been suggested in another report, for example, that the global membrane bioreactors (MBRs) market will reach $1.3 billion by 2015 (GIA, 2009).

1.3.2. Suppliers

A review of the share of the municipal market across the MBR membrane product suppliers reveals it to be still dominated by the original three suppliers (Fig. 1.3), with Kubota providing around 20–25% of the total number of MBR installations for the top 11 MBR membrane providers (with

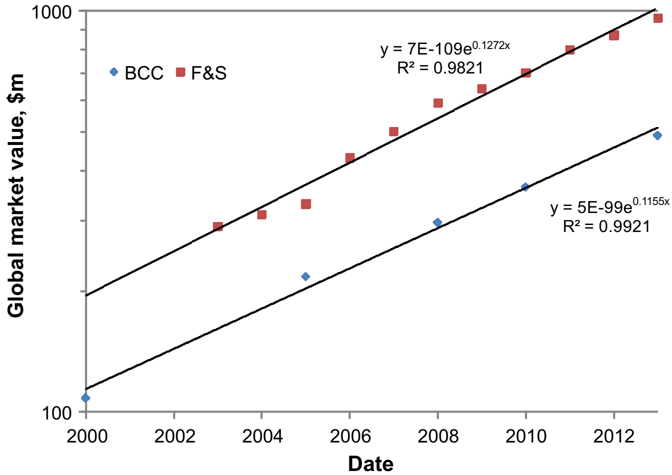

FIG. 1.2 MBR global market value in $bn; data taken from Frost and Sullivan and BCC reports. (*Srinivasan, 2007; BCC, 2008*).

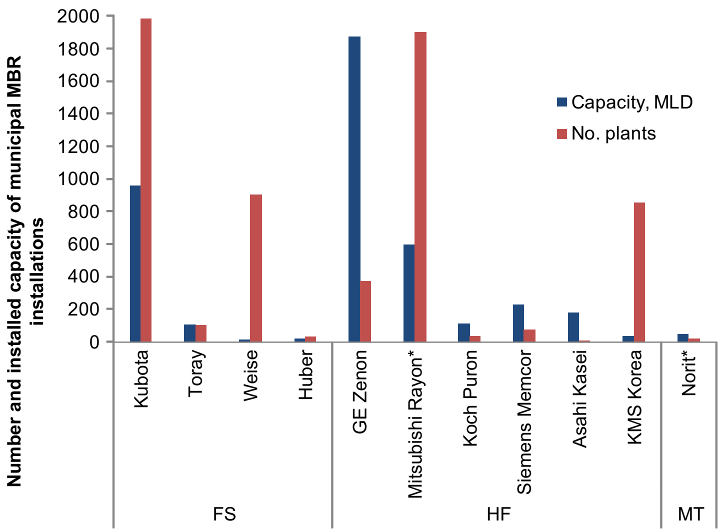

FIG. 1.3 MBR municipal market; *estimated figures from available information (*from Santos and Judd, 2010*).

respect to installed capacity) and GE Zenon more than 40% of the total global installed capacity for MBR treatment. Mitsubishi Rayon Engineering (MRE) have an estimated similar number of municipal installations to Kubota, with their activities largely limited to the Far East. However, newer MBR membrane products are increasing in number and market share. As

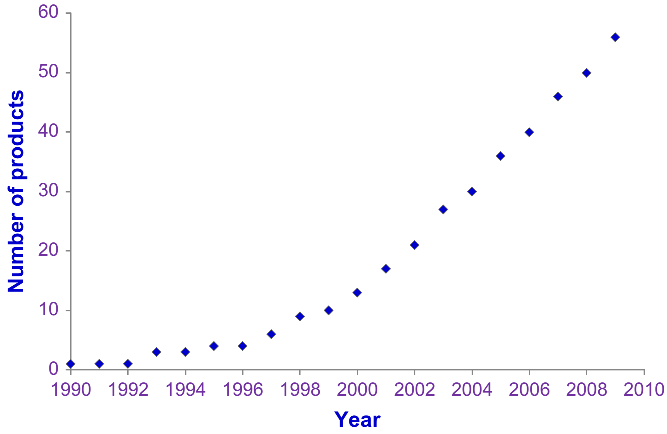

FIG. 1.4 Number of MBR membrane module products.

recently as 2003, the three most established players of Kubota, Mitsubishi Rayon and Zenon held 85−90% of the municipal MBR market, with around 800 installations between them (Pearce, 2008). By the end of 2009 the total number of installations provided by these three suppliers had risen to around 4400, with at least 32 other membrane suppliers with wastewater treatment MBR reference sites and a total number of MBR membrane module products approaching 60 (Fig. 1.4).

A review of the geographical location of the MBR membrane module suppliers (Fig. 1.5) reveals them to derive primarily from East Asia, with China, Korea and Japan accounting for more than half of the 45 MBR membrane product suppliers identified by May 2010, and the EU nations, and principally Germany, providing much of the remainder. Moreover, there are more such products either currently close to being commercialized or else already commercially available but not visible through the usual routes of internet

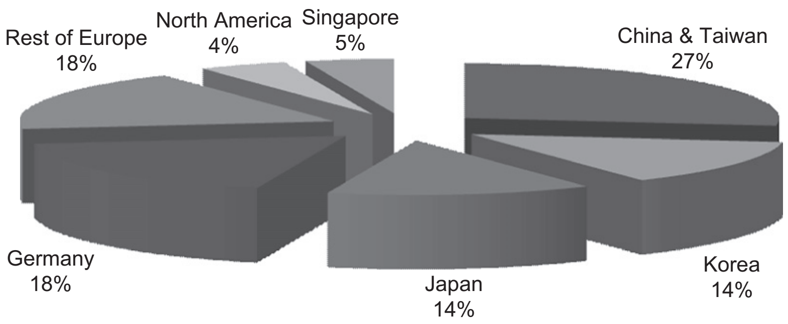

FIG. 1.5 MBR membrane product suppliers by geographical location.

search engines, international trade shows or articles/advertisements in trade magazines. Whilst it is not always possible to distinguish between original membrane or membrane module manufacturers (i.e. original equipment manufacturers, or OEMs) and those which acquire these products and rebrand them for sale, it is apparent that many of these products are discrete and establishing a market either by geographical region or industrial sector. It is also the case that the majority of the commercially available iMBR membrane

TABLE 1.2 The 20 Largest MBR Plant (May 2010)

Project	Technology	Date	PDF MLD
Shending River, China	BOW	2010	120
Wenyu River, China	Asahi K/BOW	2007	100
Johns Creek, GA	GE Zenon	2009	94
Beixiaohe, China	Siemens	2008	78
Al Ansab, Muscat, Oman	Kubota	2010	78
Peoria, AZ	GE Zenon	2008	76
Cleveland Bay, Australia	GE Zenon	2007	75
Sabadell, Spain	Kubota	2009	55
San Pedro del Pinatar, Spain	GE Zenon	2007	48
Syndial, Italy	GE Zenon	2005	47
Broad Run WRF, VA	GE Zenon	2008	47
Beijing Miyun, China	MRE	2006	45
NordKanal, Germany	GE Zenon	2004	45
Tempe Kyrene, AZ	GE Zenon	2006	44
Brescia, Italy	GE Zenon	2002	42
Traverse City, MI	GE Zenon	2004	39
Linwood, GA	GE Zenon	2007	38
North Kent Sewer Authority, MI	GE Zenon	2008	35
Jinqiao Power, China	GE Zenon	2006	31
Dubai Sports City, UAE	GE Zenon	2009	30

PDF, Peak daily flow; MLD, Megalitres per day; BOW, Beijing Origin Water; and MRE, Mitsubishi Rayon Engineering.

module products are based on either flat sheet (FS) or HF configuration, normally formed as rectangular panels or, in the case of a few of the HF products, cylindrical bundles. MBR technologies are distinguished as much by the engineering of the process as the design and configuration of the membrane itself.

Globally, there is also a pronounced upward trend in plant size, reflecting observations reported for the EU by Lesjean and Huisjes (2008) and Lesjean, Ferre, Vonghia and Moeslang (2009), as well as in diversity of technology providers – although the largest MBRs are predominantly fitted with GE Zenon technology (Table 1.2). A review of the largest installations, including those in planning or construction and due before 2011, reveals that some suppliers who have launched products post-2000 have been able to secure contracts for very large projects – particularly in China and the Middle East. This would seem to reflect a more general trend in increasing acceptability of comparatively new technologies. Of the 14 products listed in Table 1.1, only three pre-date 2000 and many have less than 50 reference sites. Notwithstanding this, some very large installations are planned based on these technologies despite some being no more than a few years old. This provides further evidence of the change in the perception of MBRs. Whilst still viewed by many practitioners as a 'new' or 'high-risk' technology, it appears that fewer reference sites are now required for a technology to be considered commercially acceptable at a large scale. Indeed, a correlation of the time taken for a technology to achieve the first 10 MLD capacity plant provides a stark illustration of this, with the gestation time sharply decreasing since the turn of the Millennium (Fig. 1.6).

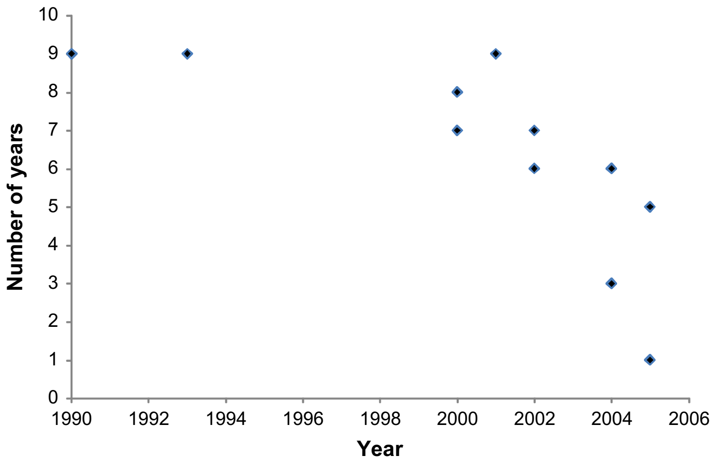

FIG. 1.6 Time taken between product launch and installation of first plant of more than 10 MLD capacity for 12 MBR membrane products (including two coincidental data).

1.4. DRIVERS

1.4.1. Global Key Drivers

As already noted, the increased number of large MBR plants would seem to reflect a growing confidence in the technology, and this has been accompanied by a significant increase in the number of product and technology suppliers. Whereas in the past MBRs may have been disregarded in favour of conventional treatment plants, it is now the case that for applications where footprint is limited and a high product water quality is demanded – and for reuse in particular – the MBR is the technology of choice. However, notwithstanding generally high global growth rates (Fig. 1.2), implementation varies significantly from country to country. Countries such as Japan, Singapore, USA, China and many parts of Europe have embraced MBR technology since the advent of the immersed configuration in the 1990s, whereas the particular challenges presented in other countries such as, for example, India and Malaysia have resulted in little or no take-up to date. These differing rates of development aside, a number of global key drivers emerge which influence each of the respective regional MBR markets to an extent depending on political, economic and environmental circumstances, the key drivers being:

- Legislation
- Local water scarcity
- Return on investment
- Environmental impact
- Public and political acceptance.

1.4.1.1. Legislation

Legislation and associated regulatory functions exert the greatest influence on the global MBR market, and particularly so in the municipal sector. Legislation often drives the specification of both potable and discharge water quality as well as the extent of freshwater resource preservation, through demand management or reuse, and so influences the choice of water and wastewater treatment technologies. Of critical importance, therefore, is the extent to which existing assets are able to deliver treated water to the quality demanded by newly promulgated legislation, as well as the capacity of the regulators to enforce it.

In the European Union, pertinent legislation is manifested as a series of acts principally relating to environmental protection and water and wastewater management. Whilst these pieces of legislation (typically in the form of Directives) serve to provide Europe-wide standards, individual countries are able to interpret the Directives nationally and determine their implementation plans within the framework provided. As a result, some countries appear to have implemented the laws more fully than others. Thus, in those countries fully embracing MBRs as the best available technology, rigorous water quality contracts with the suppliers are instigated with accountability and punitive financial measures imposed in the event of product water quality breaches.

European legislation of most importance with respect to MBRs includes the following:

- The EC Bathing Water Directive (2006): this directive should be adopted by member states by 2015 and is designed to improve bathing water quality with respect to pathogenic micro-organism levels. Introduced to replace the original Directive of 1976, it moves from a simple sampling and monitoring of bathing waters approach to focus on bathing water quality management.
- The Water Framework Directive (2000): this is the most substantial piece of EC water legislation to date, and demands that water throughout the EU member states be managed on a catchment level and lays down environmental quality standards (EQSs) in the field of water policy. This very comprehensive Directive integrates many other Directives concerning water resources and discharges, and has produced a number of daughter Directives since its promulgation. Possibly the most notable of these is the Priority Substance Directive of 2008, in which limits on concentrations in surface waters of 33 priority substances and eight other pollutants have been proposed. This Directive replaced five other previous ones.
- The Urban Wastewater Treatment Directive (1995): the purpose of this Directive, which was agreed in 1991, is to protect the environment from the negative effects of sewage discharges. Treatment levels are set taking into account the size of sewage discharges and the sensitivity of the waters into which the discharges are to be released.

In the USA, much of the legislative framework is centred around the following:

- The Pollution Prevention Act (1990): the purpose of this legislation is to focus industry, government and public attention on reducing the amount of pollution through cost-effective changes in production, operation and raw materials use. Pollution prevention also includes other practices that increase efficiency in the use of energy, water or other natural resources, and protect water resources through conservation. Such practices include recycling, source reduction and sustainable agriculture.
- The Safe Drinking Water Act (1974): this focuses on all waters actually or potentially intended for drinking, whether from aboveground or underground sources. The Act authorizes the Environmental Protection Agency (EPA) to establish safe standards of purity and requires all owners or operators of public water systems to comply with primary (health-related) standards. Whilst numerous amendments and regulations have been introduced since 1974, many of these relating to the control of disinfection by-products and other organic and inorganic contaminants, none appears to have been directed specifically towards wastewater reuse.

- The Clean Water Act (CWA) (1972): this established the basic framework for regulating discharges of pollutants into US waters and authorized the setting of wastewater standards for industry. The Act was revised in 1977, 1981 and 1987, and was originally intended to ensure receiving waters became 'fishable' or 'swimmable', although studies suggest that there is still room for improvement in meeting this goal.

In an attempt to reach the 'fishable' and 'swimmable' goals in the USA, the total maximum daily load (TMDL) programme has been established. Section 303(d) of the CWA requires the establishment of a TMDL for all impaired waters. A TMDL specifies the maximum amount of a pollutant that a water body can receive and still meet water quality standards considering both point and non-point sources of pollution. The TMDL addresses each pollutant or pollutant class and control techniques based on both point and non-point sources, although most of the emphasis seems to be on non-point controls. MBRs offer the opportunity of a reduction in volume of point source discharges through recycling and improving the quality of point discharges to receiving waters. It is this that has formed part of the rationale for some very large MBRs, such as the Broad Run Water Reclamation Facility plant at Loudoun County in Virginia.

In the USA, individual states, and particularly those with significant water scarcity such as California and Florida, may adopt additional policies and guidelines within the federal legislative framework. The state of Georgia, for example, has implemented a water reuse initiative entitled 'Guidelines for Water Reclamation and Urban Water Reuse'. The guidelines include wastewater treatment facilities, process control and treatment criteria, as well as system design, operation and monitoring requirements. California has introduced a series of State laws since the promulgation of the Federal Water Pollution Control Act, as amended in 1972.

This is a small selection of pertinent legislation since a full review of legislation, regulations and guidelines from across the globe is beyond the scope of this book, though the legislative and regulatory position in a few individual countries is discussed in Section 1.4.2. However, with both social (e.g. population growth) and environmental (e.g. climate change) trends putting ever more stress on water resources, there is every reason to suppose that legislation will continue to be used to improve the efficiency and security of water services.

1.4.1.2. Local Water Scarcity

Even without legislation, local water resourcing problems alone can provide sufficient motivation for recycling. Water scarcity is determined by the ratio of total freshwater abstraction to total resources, indicating the availability of water and the pressure on water resources. Water stress occurs when the demand for water exceeds the amount available during a certain period, or when poor quality restricts the use of available water. Areas with low rainfall and high population density or those where agricultural or industrial activities are intense are particularly prone to water stress. Changing global weather patterns aggravate the situation, in particular for those countries which are prone to drought conditions. Water stress induces deterioration of freshwater resources in terms of quantity (aquifer over-abstraction, dry rivers, etc.) and quality (eutrophication, organic matter pollution, saline intrusion, etc.). A widely used measure of water stress is the water exploitation index (WEI), representing the annual mean total demand for freshwater divided by the long-term average freshwater resource. It provides an indication of how the total water demand puts pressure on the water resource.

Data from the year 2009 indicate that nine European countries (Belgium, Bulgaria, Cyprus, Germany, Italy, the former Yugoslav Republic of Macedonia, Malta, Spain and the United Kingdom), representing 18% of Europe's population, were considered to be water stressed; this compared with only four countries so classified in 1999. It is estimated that, in 1990, around 1.9 billion people lived in countries which used more than 20% of their potential water resources. By 2025, the total population living in such water-stressed countries is expected to increase to 5.1 billion, this figure rising further to 6.5 billion by 2085. On the other hand, climate-related water stress is expected to decrease in some countries, for example, the USA and China, while in Central America, the Middle East, Southern Africa, North Africa, large areas of Europe and the Indian subcontinent, climate change is expected to increase adversely water stress by the 2020s. It is also predicted that 2.4 billion people will live in areas of extreme water stress (defined as using more than 40% of their available water resources) by 2025, 3.1 billion by 2050 and 3.6 billion by 2085.

1.4.1.3. Return On Investment

MBRs tend to be more costly and energy intensive than conventional processes, despite the significant decrease in membrane costs since the initial commercialization of the immersed configuration in 1990 (Kennedy & Churchouse, 2005). Because of this and the perceived novelty of the technology, reflected in a paucity of extensive reference data needed to support investment decisions, there has in the past been some reluctance to invest in the process in some areas. However, the maturing of the technology and the much wider knowledge of the process, in particular the key aspects of energy optimization and process failure risk, have promoted greater confidence in the technology generally and subsequently greater willingness to invest in ever larger plant (Table 1.2).

Membrane costs and, in particular, membrane life remain of key concern. Membrane purchase costs decreased almost exponentially over the course of the 1990s (Kennedy & Churchouse, 2005) as a simple consequence of supply and demand, contributing to a decrease in the treated water cost of more than an order of magnitude. Given the generally lower production costs achievable in the highly industrialized Far Eastern countries of China and Korea, it seems likely that membrane costs will continue to decrease — though not as dramatically as during the 1990s. Membrane life, on the other hand, remains a challenging parameter to define. There is increasing evidence from some plants that membrane life can exceed a decade, and is more determined by the extent of manual intervention than any other factor relating to routine operation. Provided a long membrane life can be assumed, then the costs of installing and running MBRs can be comparable with those of conventional treatment plants on a whole-life basis, with the added benefit of improved effluent quality. MBRs are also becoming more energy efficient, as new products materialize and means of operating existing plant at lower aeration demands are devised.

An additional consideration in some countries is the availability of state incentives. An example is the Enhanced Capital Allowance scheme introduced in the United Kingdom in 2001, whereby tax incentives are offered for water-efficient technologies as part of the Green Technology Challenge. Other countries, such as the USA, Australia, Canada, Finland, France, the Netherlands, Switzerland, Japan and Denmark, have all offered incentives in various forms to promote innovative water-efficient technologies and reduction in freshwater demand. The number of countries and governmental organizations offering such incentives is growing, essentially making more affordable advanced technologies such as MBRs and other membrane-based processes generally required to attain reusable water. Lastly, the small footprint generally incurred by MBRs compared with conventional processes provides a further financial incentive relating to the cost of land.

1.4.1.4. Environmental Impact

Many of the environmental impact aspects of the MBR technology relate either to cost (Section 1.4.1.3) or plant size. The most significant components of the operating costs are the energy demand, membrane replacement and waste (primarily sludge) management. The nature of biological processes generally is that a reduction in the sludge generated demands an increase in the energy input. For an MBR the reduction in sludge generated can be accompanied by a reduction in the plant size. There is thus a trade-off between embedded and generated carbon which is greatly affected by the sludge management component. However, notwithstanding these energy-related issues, a key facet of MBRs providing a favourable environmental impact is the consistently high product water quality.

MBRs are capable of the quantitative removal of suspended solids and pathogenic micro-organisms from municipal effluents, very significant removal of ammonia and, if appropriately configured, nutrient removal. The capability for disinfection has led to the wide-scale implementation of the technology at coastal sites around Europe to achieve compliance with the Bathing Water Directive. The ability of the process to produce a product water which can be fed into a reverse osmosis (RO) plant with no further processing required has

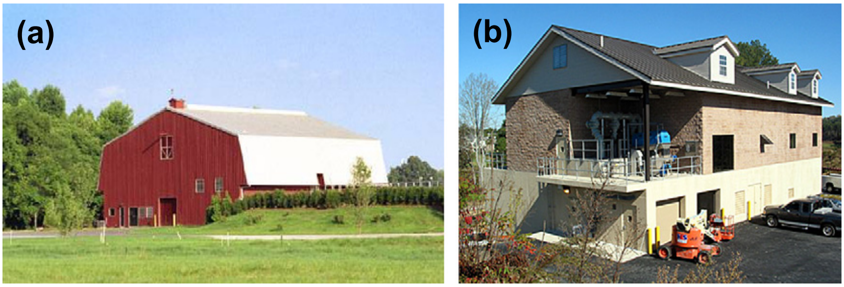

FIG. 1.7 MBR buildings: (a) Cauley Creek, GA and (b) Hamptons, GA, *(with kind permission from GE Zenon and Ovivo).*

also influenced its uptake; indirect potable reuse (IPR) by, for example, aquifer recharge or direct reuse by industrial processes and/or utilities generally demands RO treatment. On the other hand, MBRs do not generally offer significant (i.e. an order of magnitude or more) improvements in process efficacy over that of conventional processes for the removal of low-concentration priority substances (Section 2.3.10).

The physically smaller size of MBR plants compared with those conventionally employed for effluent treatment becomes important in areas where: (a) unit land costs are high and increasing at a rate greater than that of the general price index, (b) space on site is limited and (c) legal restrictions have been imposed on the permitted visual impact of the plant. The latter has led to the housing of MBRs in buildings quite unlike those normally associated with municipal wastewater treatment (Figs 1.7, 5.2, 5.4 and 5.35). The option of being able to limit the obtrusiveness of the plant has directly influenced the decision to implement the technology at a number of sites worldwide, as well as in the retrofitting to existing plants.

1.4.1.5. Public and Political Acceptance of MBR Technology

A key theme governing the take-up of MBRs in any country is the acceptance of the technology by the various stakeholders, which can include the public, the politicians and, of course, the decision-makers within the procuring organization. The various individual aspects of the technology itself likely to positively or negatively influence stakeholders have already been outlined, specifically the cost, footprint and energy demand (Section 1.4.1.3), the product water quality and plant (Section 1.4.1.4) and the security of the water supply (Section 1.4.1.1). The most contentious perception issue directly impacting on the uptake of MBR technology is that of wastewater reuse, of which a plethora of literature is available.

Despite the very persuasive technical aspects endorsing the direct reuse of municipal wastewater for potable water supply in water scarce or water-stressed regions, only one such 'toilet to tap' plant currently exists in the

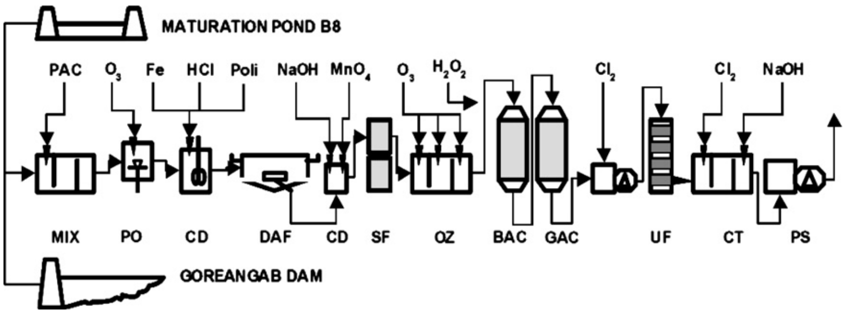

FIG. 1.8 The treatment scheme at the New Goreangab Reclamation Plant at Windhoek *(by kind permission from Jürgen Menge, City of Windhoek).*

world – the New Goreangab Reclamation Plant at Windhoek in Namibia commissioned in 2002. This plant, which does not employ MBR technology but instead uses UF, has 12–13 individual process steps all designed to provide fail-safe drinking water quality from a treated sewage feed (Fig. 1.8). Whilst technically the process may be considered somewhat over-engineered, the human dimension demands this multi-barrier approach to wastewater recovery for potable use, and that extremely rigorous operation and maintenance protocols are in place to ensure appropriate final product water quality.

1.4.2. National Key Drivers

1.4.2.1. South Africa

The South African MBR market is in its incipient stages, and although there are a few small plants, there are only two MBR plants greater than 1 MLD in capacity. Kubota established a presence in South Africa in 2004 and spearheaded the introduction of the concept of the MBR process in Southern Africa, undertaking several pilot trials on different industrial processes and on domestic wastewater effluent. Research on the MBR process was initiated at the University of Cape Town in 2004, using Kubota bench-scale panels to undertake a comparison with the CASP. In the past 2–3 years several of the international MBR vendors have established a presence in South Africa, and there has been a flurry of pilot plant trials driven by a looming water shortage crisis, and the need to meet discharge standards, footprint limitations and waste minimization.

The longest running MBR plant in Southern Africa was implemented at the Illovo Sugar plant at Sezela using Kubota flat plate membranes. This MBR plant was commissioned in 2005 and operates at a 1.2 MLD flow capacity (Section 5.2.1.7) using 4800 Kubota *EK 400* membrane units. A driver for implementing MBR technology was the introduction of stricter discharge controls on industrial effluent discharges by the South African national regulator, the Department of Water Affairs and Forestry. The ability of MBR systems to operate at a higher biomass concentration also makes the technology more resistant to any toxins that may enter the process, a particular aspect of the Sezela plant. This plant has been running successfully since commissioning with its original membranes, even though the plant has for a significant proportion of its history been operating at temperatures above 55 °C. The largest MBR plant as of June 2010 is the Zandvliet plant in the Western Cape, which treats municipal wastewater and supplements the capacity of a conventional wastewater treatment works. The drivers here were threefold: to increase the capacity of the wastewater treatment works; to meet stringent discharge standards; and the eventual necessity for water reuse. The plant, commissioned in early 2009, has a mean capacity of 18 MLD and uses ZeeWeed *ZW500D* HF modules. Initial problems were experienced with the pre-treatment train but these appear to have been resolved.

Various water authorities in the Western Cape, Eastern Cape and KwaZulu Natal appear to be rapidly embracing MBRs, mainly driven by the impending water shortage, but also because of discharge into environmentally sensitive areas. A 20-MLD municipal wastewater plant planned for Malmesbury, 100 km north of Cape Town on the Western Cape, is as at July 2010 out to tender, and construction should commence towards the end of 2011 for commissioning in 2012. The main drivers here are discharge into an ecologically sensitive area, and eventual water reuse. A 40-MLD municipal wastewater plant, primarily for water reuse, is also as at July 2010 out to tender for Belville in the Western Cape. A further 100 MLD plant to treat municipal wastewater and provide reuse water for the Coega Industrial Zone in Port Elizabeth, Eastern Cape, is also planned to go out to tender in 2011. It appears that most of the major international MBR vendors are tendering for these plants.

Umgeni Water, in KwaZulu Natal, is currently conducting pilot-scale MBR trials at the Darville Wastewater Treatment Works. This plant is aimed to supplement the capacity of the Darville Works, as well as to provide recovered water. Three MBR technologies are being evaluated in parallel: Norit (external air-lift), Toray (flat sheet) and Pall (hollow fibre), ranging in capacity from 2 m^3/h to 5 m^3/h. The main objective of the trials is to determine the stability and operability of the various technologies under 'developing economy' conditions, such as operational failures, electricity downtime and surges in feed quality. All three units should be commissioned by the end of 2010. This demonstration trial has the potential of opening up the municipal MBR market by overcoming the reservations held by many water authorities that MBRs are a 'first-world' technology unsustainable in developing economy conditions.

Various industrial MBR pilot plant trials have been performed at textile, distillery and other sites, but have yet to be manifested at full scale. A possible barrier to implementation is the driver being limited to meeting discharge standards, which can be obtained by existing chemical treatment processes, rather than water reuse. However, some of the larger industries are actively involved in MBR pilot trials that are likely to lead to full-scale applications in the very near future. The major drivers for these are water reuse, to meet water balances, and waste minimization. Two to three years ago it was expected that wine estates and other farms in the Western Cape could see a major swing towards small-scale MBRs. However, this has not occurred, partially due to the international economic downturn. There are, however, some indications that private developers of housing estates are contemplating MBRs to facilitate reuse for irrigation and utilities, such as fire extinguishing.

There are also a few MBR units in neighbouring countries to South Africa. A unit based on Kubota flat-sheet (FS) membranes has been operating at the Grand Palm Casino in Botswana since 2006, treating domestic effluent and producing irrigation water, at a capacity of 0.5 MLD. A unit, also using Kubota FS panels, has been set up by the oil company Sasol in Mozambique to treat domestic effluent for fire control water at a capacity of 0.5 MLD. A small unit based on Microdyn-Nadir FS units has been installed at the British Embassy in Harare, treating 0.24 MLD of domestic effluent.

On the research side, a group at the University of the Western Cape is looking into modification of polymeric membranes using specific nano-structures to produce low fouling membranes for MBRs. The Department of Chemical Engineering at Durban University of Technology is currently evaluating FS membranes fabricated from a woven fabric; it is claimed that these membranes are substantially more robust than current commercial membranes, and are thus better suited to small-scale MBR applications in developing economies. The Pollution Research Group, University of Natal, is investigating the integration of membranes into the Decentralized Wastewater Treatment Systems (DEWATS) anaerobic baffled reactor (ABR), to polish the effluent for possible agricultural use.

1.4.2.2. Australia

Investigation of MBR technology in Australia commenced in the late 1990s, following its emergence elsewhere in the world. The first full-scale Australian MBR was built at Picnic Bay on Magnetic Island near Townsville (north Queensland), and has been operational since 2002.

Australia is the world's second driest continent (second only to Antarctica). Most of its population lives in a relatively narrow coastal band where rainfall is typically highest. However, recent droughts and a range of local factors (including water supply and demand, size and yield of dam storages) have resulted in increasingly strained freshwater supplies for many Australian towns and cities. Although the main drivers for MBR technology are similar to those worldwide, water scarcity is a key factor in Australia. The main specific drivers may be summarized as follows:

- *Water recycling initiatives*, partly driven by constraints on discharge to receiving waters and partly by scarcity of freshwater or efficiency improvements within industry. The high quality of treated water from MBR systems is an obvious advantage for water recycling, and this typically includes low concentrations of solids and pathogens in the MBR permeate (Section 1.4.1.4). Furthermore, chemically assisted MBRs can also achieve very low phosphate concentrations in the permeate, potentially eliminating pre-treatment for water reclamation processes where inorganic scaling of reverse osmosis membranes is a concern.

- *Financial considerations*, particularly driven by escalating civil construction costs for sewage treatment plants. This, together with decreasing membrane costs over the past two decades, has meant that an MBR plant today can typically be constructed for approximately the same capital cost as a conventional wastewater treatment plant, especially when comparing alternative processes for achieving the same treated water quality. Operating costs for MBRs (dominated by power and membrane replacement, Section 3.5.3) have also decreased in recent years. Given historically relatively low bulk electricity costs in Australia, near parity on whole-life cost has also been demonstrated compared with conventional plants. However, a rapid escalation in electrical power costs (2009–2010) and the prospect of further increases (including carbon permit costs) may change such financial outcomes in the future.
- *Space constraints*, driven by relatively high population densities in coastal areas and legislative or other barriers to approvals for new wastewater treatment sites (e.g. pumping costs to more remote locations in relatively flat coastal zones; complex local planning regulations; or community opposition). The low odour emission rate typical of MBRs is also a significant driver in terms of overall plant footprint considerations in this context.

MBR implementation is influenced by a combination of federal and state-based legislation that drives wastewater treatment and water quality in Australia. The most important federal law in this respect is the Environment Protection and Biodiversity Conservation Act of 1999, which allows a Commonwealth Minister to decide whether a potential project threatens endangered species in proximity to designated lands such as World Heritage Listed or Ramsar sites, both of these relating to international environmental protection legislation.

In this regard, the large MBR projects in North Queensland (e.g. Townsville, Cairns) have been driven partly by environmental concerns over the Great Barrier Reef, which is a World Heritage Area. A regulatory authority set up by the Australian Commonwealth (Federal) Government is tasked with managing the Great Barrier Reef Marine Park in accordance with the principles of ecologically sustainable development, aiming to protect its natural qualities, while providing for reasonable use. One of the main threats to reefs is increased nutrient load to the marine ecosystem due to human activity, including wastewater and agricultural run-off. The design criteria for the expansion of wastewater treatment plants in areas adjacent to or within the jurisdiction of the

Great Barrier Reef Marine Park Authority were therefore aimed at capping or reducing nutrient loads discharged to the marine environment. Through a combination of biological and physico-chemical nutrient removal processes, MBRs were found to be best suited to producing high-quality low nutrient effluent suitable for water recycling (e.g. land irrigation of golf courses, sports fields, public open spaces and toilet flushing), and hence either zero or limited marine discharges.

State-based legislation is generally administered regionally under the relevant Environmental Protection Acts (or similar) that require licenses (often named Development Approvals) for establishing, operating and expanding wastewater treatment plants. An expansion of a wastewater treatment plant (WwTP) to a capacity that exceeds an existing license stipulation would typically be considered an environmentally relevant activity, although nomenclature and thresholds differ in the respective states. Depending on the scope of the project, an Environmental Impact Statement (EIS) might be required, based on a detailed environmental study of likely effects of the treatment plant, including discharges to water, air and land, noise and other nuisances. A receiving water study will typically be included, tested against guidelines such as those of the Australian and New Zealand Environment Conservation Council (ANZECC) or other regional objectives (such as the Healthy Waterways Partnership in South East Queensland).

By way of example, the Queensland Environmental Protection Agency from 2004–2005 onwards has adopted an operating policy which requires all new Development Approval applications for wastewater treatment plants to include an assessment of options for water recycling (effluent reclamation and reuse) to the maximum potential (targeting >90% reuse). In terms of this operating policy, nutrient removal (effluent targets 5 mg N/L Total N and 1 mg P/L Total P as 50th percentiles) is a default requirement in the absence of effluent reuse, with relaxations from annual loads calculated on this basis permitted for sites where effluent reuse can be demonstrated. In practice, high levels of effluent reuse have not been possible in all cases due to climatic and other factors. However, a high treated water quality improves the potential to maximize reuse. Due to their suitability for water recycling applications (e.g. agricultural irrigation), MBRs have thus provided advantages with regard to meeting such regulatory requirements.

Between 2006 and 2008, revised legislation and guidelines were published in Australia covering water recycling. This followed major droughts, particularly in the southern and eastern states (South Australia, Victoria, New South Wales and Queensland). Examples include *National Guidelines for Water Recycling: Managing Health and Environmental Risks* (published by the Natural Resource Management Ministerial Council, 2006), *Water Quality Guidelines for Recycled Water Schemes* in Queensland (Department of Natural Resources & Water, 2008) and *Public Health Regulations (Amended 2008)* in Queensland (Public Health Act, 2005). These guidelines recommend a risk assessment-based approach to water recycling, or in some cases legislate the recycled water quality requirements according to class (e.g. Class A+ or Class A). Validation of performance of a given technology or process step (e.g. an MBR) is based on a requirement either to perform actual challenge tests (e.g. for indicator bacteria and viruses), or to cite published literature sources (e.g. US EPA or scientific papers).

A significant challenge to the application of MBRs in Australia is handling of wet weather flows in municipal applications. Australian sewer and stormwater collection systems are designed to be separate. Despite this, significant infiltration of stormwater to the sewer systems does occur in most areas. Whilst wet weather events are infrequent, it is common for design peak wet weather flow rates to be >3−11 (typically 5) times average dry weather flow rates, due to high local rainfall intensities. This either requires installation of additional membrane modules to serve during peak weather, with attendant issues of 'idling' these modules during dry weather without excessive power consumption, or some alternative strategy for handling wet weather flow. The latter comprises off-line storage; partial by-pass of the plant; or sidestream treatment in a parallel conventional continuous-flow process maintained for this purpose.

Industrial MBR applications have largely resulted from water handling efficiency and recycling initiatives, driven by a combination of the increasing potable water supply and trade waste costs, as well as mandatory water restrictions in some cases (usually drought-related). Space constraints are typically also more significant at industrial sites, making MBRs more attractive than conventional biological processes. Challenges in industrial applications include accelerated biofouling and/or inorganic scaling, particularly in combination with RO for brewery applications.

An interesting application is the so-called Gippsland Water Factory in the state of Victoria where a combination of domestic sewage and wastes from a pulp and paper mill is treated. In terms of average flow (35 MLD), it is the largest MBR plant in Australia, being commissioned in 2010. The project was driven by a number of different factors, including the need to produce a very high treated water quality for reuse and concerns over activated sludge settleability when treating pulp and paper effluent in conventional biological treatment processes that depend on sedimentation for secondary clarification (Fig. 2.16).

As of mid 2010 there were at least 44 full-scale applications of MBR in Australia, either operating or under construction, excluding smaller 'on-site' systems at household or cluster housing scale and mine sites. In terms of average flow rate, these range from approximately 0.04 (small systems located in buildings or sewer mining applications) to 29 MLD average daily flow (ADF) for medium to large sewage treatment plants (the Cleveland Bay Wastewater Treatment Facility, Townsville, Fig. 1.9). The majority (approximately 90%) are municipal plants designed to treat predominantly domestic sewage.

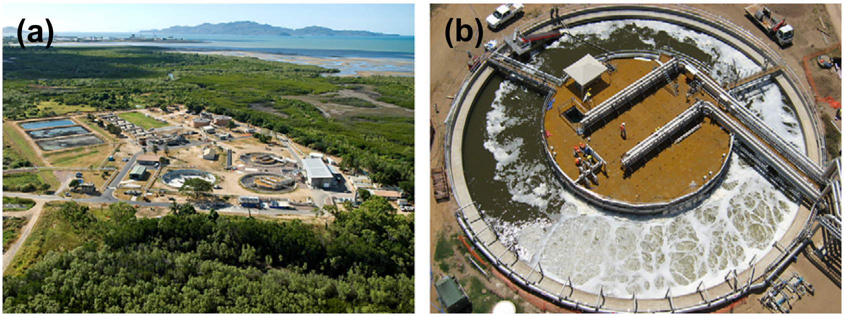

FIG. 1.9 Cleveland Bay (29 MLD ADF) Wastewater Treatment Facility, Townsville. Photos taken in 2007 during MBR commissioning of (a) the site, and (b) the MBR plant, fitted into an existing old sedimentation tank. Magnetic Island (the location of Picnic Bay, the first full-scale MBR plant in Australia), is in the far distance of (a).

1.4.2.3. China

Research into MBRs started at the beginning of the 1990s in China, and progress from that time in the application of MBR technology can be roughly divided into the following stages:

1990–2000:	Laboratory experiments, pilot-scale tests and a few demonstration projects
2000–2003:	Practical application on a scale of hundreds of tonnes per day, mainly serving small residential areas and/or industrial sectors
2003–2006:	Practical application on a scale of thousands of tonnes per day for municipal and industrial wastewater treatment
2006 onwards	Practical application of large-scale MBRs of over 10,000 tonnes per day.

The earliest practical applications were generally <100 m^3/day capacity in the early 2000s. However, since 2006, the number of large-scale MBRs has been substantially increasing with the average annual growth rate more than 50% – considerably higher than the average annual growth rate of 11.5–12.5% (Fig. 1.2) of the global MBR markets. Today there are numerous large-scale MBR plants in China (Table 1.3). Of these plants, particularly noteworthy are the MBRs of Beijing Miyun wastewater treatment plant (Section 5.3.2.1) and Wenyu River water treatment plant (Section 5.3.2.1), landmarks of MBR application in China. The Beijing Miyun MBR plant was the first >10 MLD MBR plant in China, and the Beijing Wenyu River MBR plant the first 100 MLD plant worldwide. Between 2003 and the end of 2009, there were more than 100 MBRs installed in China providing total wastewater treatment capacity of close to 1200 MLD, based on MBRs having an installed capacity greater than 0.1 MLD.

Among these installations, there are nearly 30 large-scale MBR plants with a design capacity greater than 10 MLD. The largest so far is Shiyan Shending River WwTP (Section 5.3.3.1) located in Hubei Province, which has a capacity of 110,000 m^3/d and was commissioned in October 2009 for municipal

TABLE 1.3 MBR Plants for Wastewater Treatment in China (>10 MLD)

MBR installation	Location	Wastewater Origin	Membrane Supplier	Capacity MLD	Engineering Contractor	Commissioned
Miyun WwTP	Beijing	Municipal	Mitsubishi Rayon	45	Origin Water	2006
Jingqiao power plant WwTP	Inner Mongolia	Municipal	GE	31	Lucency	2006
Huizhou Dayawan Petrochemical Engineering Corporation	Guangdong	Petrochemical	Asahi Kasei	25	NOVO	2006
Xiaohu Island Petrochemical Industrial Park	Guangdong	Petrochemical	Asahi Kasei	10	NOVO	2006
Hainan Petrochemical Engineering Corporation	Hainan	Petrochemical	Asahi Kasei	10	NOVO	2006
Luoyang Petrochemical Engineering Corporation	Henan	Petrochemical	Memstar	18	NOVO	2007
Harbin Petrochemical Engineering Corporation	Heilongjiang	Petrochemical	Memstar	10	NOVO	2007
Huizhou Tianxin Petrochemical Engineering Corporation	Guangdong	Petrochemical	Asahi Kasei	15	NOVO	2007
Tianjin airport wastewater treatment system	Tianjin	Industrial	Tianjin Motimo	30		2007
Beixiaohe WwTP (Phase I)	Beijing	Municipal	Siemens Memcor	60	Siemens	2007

(*Continued*)

TABLE 1.3 MBR Plants for Wastewater Treatment in China (>10 MLD)—cont'd

MBR installation	Location	Wastewater Origin	Membrane Supplier	Capacity MLD	Engineering Contractor	Commissioned
Huairou WwTP	Beijing	Municipal	Asahi Kasei	35	Origin Water	2007
Wenyu River water treatment plant	Beijing	Polluted river	Asahi Kasei	100	Origin Water	2007
Pinggu WwTP	Beijing	Municipal	Asahi Kasei	40	Origin Water	2008
Chengdu banknote printing complex wastewater system	Sichuan	Banknote printing	Mitsubishi Rayon	10		2008
Mentougou WwTP	Beijing	Municipal	Mitsubishi Rayon	40	Origin Water	2009
Yanqing WwTP	Beijing	Municipal	Mitsubishi Rayon	30	Origin Water	2009
Shiyan Shending River WwTP	Hubei	Municipal	Origin Water	110	Origin Water	2009
Wuxi Xincheng WwTP	Jiangsu	Municipal	Siemens Memcor	20	Siemens	2009
Wuxi Meicun WwTP	Jiangsu	Municipal	GE	30	BMEDI	2009
Wuxi Shuofang WwTP	Jiangsu	Municipal	Mitsubishi Rayon	20	Origin Water	2009
Wuxi Chengbei WwTP	Jiangsu	Municipal	Origin Water	50	Origin Water	2009
Jiujiang Petrochemical Engineering Corporation	Jiangxi	Municipal	Asahi Kasei	12	CSEP	2009

Liulin WwTP	Shanxi	Municipal	Asahi Kasei	30	Beijing E&E	2009
Jiangsu Taixing Binjiang WwTP (Phase II)	Jiangsu	Municipal + Chemical industry	Memstar	30	NOVO	2009
Jiangsu Dafenggang WwTP	Jiangsu	Pharmacy industry	Memstar	10	NOVO	2009
Sichuan Wenchuang WwTP	Sichuan	Municipal	Memstar	10	NOVO	2009
Pengwei Petrochemical Engineering Corporation	Sichuan	Petrochemical	Tianjin Motimo	10		2010
Wenyu River water treatment plant (Phase II)	Beijing	Polluted river	Mitsubishi Rayon	100	Origin Water	2010
Kunming No. 4 WwTP	Yunnan	Municipal	Origin Water	60	Origin Water	2010
Gucheng WwTP	Yunnan	Municipal	Origin Water	25	Origin Water	2010
Wuxi Hudai WwTP	Jiangsu	Municipal	Origin Water	21	Origin Water	2010
Kunshan WwTP	Jiangsu	Municipal	GE	15	BCEED	2010
Guangzhou Jingxi WwTP	Guangdong	Municipal	Memstar	100	NOVO	2010

wastewater treatment. About seven large-scale MBR plants with a total treatment volume of 331 MLD have been contracted and commissioned in 2010 (see Table 1.3). In 2008, the MBR market in China exceeded 1.6 billion CNY (over $US230 million). Today, China has become one of the most MBR-active countries; in the next five years the MBR market is expected to continue to grow at an annual rate of around 50%.

MBR applications include treatment/reuse of municipal wastewater, industrial wastewater, landfill leachate, bathing wastewater, hospital wastewater and polluted river water. Municipal wastewater applications account for about 60% of installed capacity and industrial wastewater plants about 30%, and the rest are for polluted river water treatment, of which the Wenyu River plant is an example, and other applications. The first stage of the Wenyu River plant was commissioned in 2007 with a designed capacity of 100 MLD and the second stage, which has the same design capacity, is expected to be commissioned in 2010. In the industrial sector, most MBR plants have been used for the treatment of wastewater from petrochemical installations, followed, in order of installed capacity, by those for treating effluent from the chemical, food processing and dyeing industrial sectors.

The main membrane unit suppliers in China are Asahi Kasei (Japan), Mitsubishi Rayon (Japan), GE Zenon, Siemens Memcor, Origin Water (China), Memstar (Singapore), Tianjin Motimo (China) and Norit (Netherlands). All but one of these are HF suppliers, reflecting the prevalence of this configuration in China. Professional companies handling the engineering design, equipment manufacture and operation management of MBR plants include some global international companies such as GE, Siemens and NOVO Environmental Technology (Singapore), as well as many domestic companies which have emerged such as Origin Water and Motimo Membrane Technology. Taking into account large and medium-sized plants built by the end of 2009, Origin Water, GE, NOVO and Siemens are currently the top four market leaders in China.

Of the many factors influencing the MBR markets in China, water scarcity is the most important. Water shortage is a significant problem in China, particularly in the north-eastern and north-western areas. This problem is further exacerbated by water pollution. In China, *The Water Law of the People's Republic of China*, revised in 2002, was drawn up to manage the water resources of the country. The 52nd item of this law encourages wastewater reclamation and reuse. The Government issued further national standards for reclaimed water to promote wastewater reuse (GB/T 18919-2002, GB/T 18920-2002, GB/T 18921-2002, GB/T 19772-2005, GB/T 19923-2005 and GB 20922-2007 for the classification of wastewater reuse, urban miscellaneous uses, scenic environment uses, groundwater recharge, industrial use and farmland irrigation, respectively). MBR effluent has been extensively demonstrated as meeting these national reclaimed water standards.

In addition, in some sensitive drainage basins such as at Tai Lake and Dian Lake, eutrophication is a serious problem. The local Government has provided more stringent discharge regulations to prevent the further deterioration of water quality, and the opportunity arose to upgrade existing municipal wastewater treatment plants with MBR technology, particularly where conventional processes could not reliably meet the new discharge standards. Four MBRs installed in 2009 in Wuxi city were driven by this requirement.

Although China covers a large area, some large cities still do not have sufficient available land for the construction of municipal wastewater treatment plants. The small footprint incurred by MBR technology is especially attractive for these areas. In addition, a significant decrease in MBR investment costs as well as increased maturing and acceptance of MBR technology, especially relating to domestic companies, has continued to sustain the high level of growth of the Chinese MBR market. It is highly likely that the use of MBR technology will continue to expand in China in the future. However, economic considerations, including higher investment and running costs compared with conventional processes, will play a substantial part in its acceptance. From Table 1.3, it is clear that, to date, most large-scale MBRs treating municipal wastewater have been centred in Beijing and Jiangsu Provinces, both of which are more developed than most of the other provinces. Standardized guidance for engineering design, equipment manufacture and operation management of MBRs needs to be formulated to regulate the application of MBRs in China.

1.4.2.4. India

Clean drinking water and proper sanitation have historically been major problems in India. As India's economy was opened to foreign investors and companies in the early 1990s, it brought with it unprecedented growth and prosperity and a population migration to cities and metropolitan areas (metros). This precipitated to a surge in housing demand and, in turn, a real estate and construction boom. The pace of change was so rapid that both central and state legislators were not able to react sufficiently rapidly to the demand, such that as of 2010 there is a challenge in the provision of clean drinking water and proper sanitation since infrastructure was built (in most cases) outside any framework of legislative governance.

The centralized sewage treatment systems and sewerage lines of most metros and cities were constructed several decades ago and have not undergone modernization or expansion to meet the challenges of a rapidly growing population. This problem has now reached a scale whereby it is extremely challenging to incorporate centralized facilities in these cities. The state and Central Pollution Control Boards (CPCBs) have acknowledged that the most reasonable way of managing this situation is to enforce standards, or 'norms' (Table 1.4) for point-of-use discharge. In essence, this means that all medium and large establishments — including housing societies, hospitals, hotels, educational institutions, factories, shopping malls — are expected to have their own wastewater treatment systems in-house. An added complication is that the State Pollution Control Boards (SPCBs) have powers to adapt the regulations at will,

TABLE 1.4 Example of Enforced Norms in India for Different Applications

	Inland/ Percolation	Public Sewer with Secondary Treatment	Public Sewer without Secondary Treatment	Irrigation	Marine/ Coastal	Domestic Toilet Flushing/ Gardening
pH	5.5–9	5.5–9	5.5–9	5.5–9	5.5–9	6.5–8.5
TSS	100	600	100	200	100	20
BOD	30	350	30	100	100	3
FOG	10	20	10	10	20	1

TSS, Total suspended solids; BOD, biochemical oxygen demand; and FOG, fats, oils and grease.

with the end result that some have recommended a specific scheme viewed as most suitable to achieve the prescribed standards. Unlike many other regions of the world, legislation is therefore not the key driver for the growth of MBRs in India since general discharged effluent standards prescribed by the CPCBs do not necessitate the use of MBR technology.

The most significant driver for MBRs in India is probably the shortage of clean water. Most new real estate projects are not necessarily supplied with adequate fresh water, and thus depend to a large extent on groundwater and rainwater. However, in most areas the groundwater is brackish and RO treatment costs are considered too high, while rainwater is not a guaranteed source. Water reuse has therefore become increasingly important and MBR technology more attractive, since it is the only system that can provide consistently good quality effluent for reuse.

Historically, India has been late in adopting the latest water treatment technologies. Though RO plants were marketed and sold from the late 1980s, it was only a decade after this that the technology was truly appreciated by the end user. Similarly, UF- and MF-based treatment plants have been sold since the late 1990s, but it is only since the middle of the last decade that they have been adopted on a large scale in India. In keeping with this trend, whilst the water industry in India has been aware of MBRs for a number of years they have only been effectively marketed, promoted and sold since around 2007 and the end user is yet to fully recognize their benefits. The next five years are therefore likely to be critical to the growth of the MBR market in India.

Through a series of partnerships, all the major global players in the MBR industry now have a presence in India. GE signed a Memorandum of Understanding with Thermax Ltd, while India's other major water company Ion Exchange have partnered with Toray. Kubota and Norit provide membranes and modules on a demand basis to OEMs generally purveying MBR technology.

One of the first notable MBRs in India was installed by GE Zenon to treat municipal sewage at Cobbon Park, Bangalore in 2005, with a capacity of 1.5 MLD. Since then GE Zenon have successfully commissioned six more MBRs for varied applications such as sugar and food processing effluent recycling for boiler feed. In 2010, GE Zenon commissioned a further nine MBRs, the largest having a capacity of 10 MLD. Toray also have a strong presence in India, with as of mid 2010 six MBRs in operation, the largest being a 11.2-MLD textile effluent recycling plant, and 14 more plants are in construction. The company appears to have a strong foothold in the small-to-medium-scale plant sewage treatment sector. Norit are making significant in-roads into the MBR market as the only sMBR supplier in India. Due to escalating land prices, many end users prefer underground sewage treatment systems and thus some MBR technologies, including the Norit system, offer spatial advantages where height constraints exist.

1.4.2.5. Japan

Japan has played an important role in the field of MBR development, with pioneering trials on iMBR systems as well as the development of a variety of commercial membrane systems. In the domestic market, the first full-scale commercial MBR plants were installed in the early 1980s as an external cross-flow system. At that time, the principal target was in-building wastewater recycling system, motivated by local governmental legislation. In the mid- to late 1980s, MBRs were used in other small-scale domestic and household wastewater treatment (WwT) systems, including an on-site wastewater treatment system *johkasou* (1985–) and a night soil treatment system (1988–), as well as in industrial wastewater treatment works (WwTWs). In contrast, the application of MBRs to municipal WwTPs lagged behind these trends, the first installations being in 1999 and 2005 for small-scale rural sewerage systems and larger-scale urban sewerage systems, respectively. In March 2009, the total number of MBR installations in Japan was at least 3870.

Domestic wastewater management in Japan is rather complicated; several different systems co-exist controlled by different legislation and national ministries. The status of MBR installations in respect of individual systems is summarized below (as at March 2009, unless otherwise stated).

- Urban sewerage systems. This is the central sewage treatment system, typically for urban areas, managed by the Ministry of Land, Infrastructure and Transport (MLIT). Around 2000 WwTPs are operated in this category covering 73% of the domestic population, although MBR technology has only been implemented relatively recently (for example in 2005 at Fukusaki City). There are currently 10 MBR plants in operation, followed by a further 10 plants at the construction or planning/design stage. The capacity of existing plants ranges from 0.24 to 4.2 MLD (12.5 MLD in total). All of these plants are newly constructed, motivated typically by a small footprint and a high effluent quality requirement. An important feature of these plants is a standardized process configuration and design approach in accordance with the *JS MBR Design Recommendations* prepared by the Japan Sewage Works Agency (JS) in 2003. This document shows universal design material for a small-scale municipal MBR (less than 3 MLD), including dimensioning of bioreactors and system arrangement plans. The proposed system configuration (Fig. 1.10) and design parameters can be used for any of the five membrane technologies, for which JS has evaluated performance at pilot scale.

- Rural sewerage systems. These are small-scale sewerage systems for rural areas (typically agricultural), supervised by the Ministry of Agriculture, Forestry and Fisheries (MAFF). Among more than 5000 plants covering 3% of the domestic population, MBR technology has been installed in 50 plants, the first one being in 1999. In most of these plants, any one of three types of submerged MBR configurations developed by The Japan Association of Rural Resource Recycling Solutions (JARUS) is used.
- *Johkasou* systems. These are on-site WwT systems treating domestic wastewater from individual houses, as well as from other sources including public facilities (e.g. schools), commercial buildings, restaurants and offices. Nine percent of the domestic population use this system for their household wastewater. Although no comprehensive statistics are available, according to a survey of eight membrane suppliers and engineering companies, at least 1930 MBRs have been installed in total. The key drivers behind the selection of MBRs for these purposes are the requirements for a low maintenance and small footprint technology.
- Night soil treatment systems. In some rural areas, where none of the above-mentioned WwT systems is in place, toilet wastewater (night soil) is collected by trucks from individual houses and treated at a night soil treatment plant, typically together with excess sludge collected from *Johkasous*.

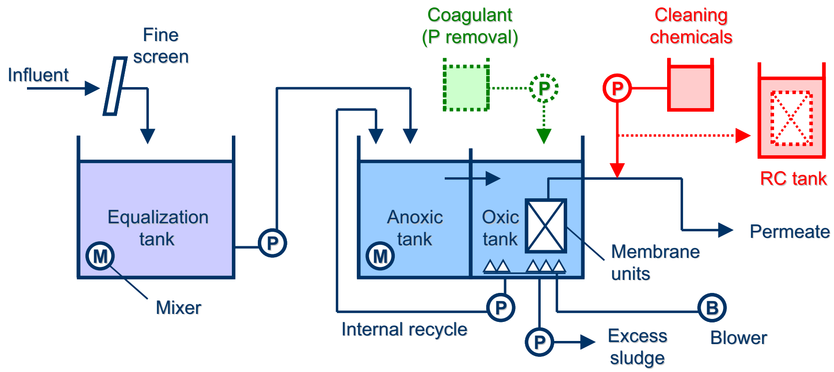

FIG. 1.10 Universal MBR system configuration provided in the *JS MBR Design Recommendations* handbook. Coagulant addition is incorporated only when phosphorus removal is required.

There are around 1000 night soil treatment plants, among which MBRs are used in 206 plants with a total capacity of 18.7 MLD. The introduction of MBRs made it possible to treat the high-strength wastewater without dilution, which had previously been demanded by conventional anaerobic or aerobic treatment.

- In-building wastewater recycling systems. Some large buildings have an on-site wastewater treatment and recycling system, in which treated water is reused for non-potable use such as toilet flushing. In some cities, the incorporation of this kind of system is obligatory for buildings of a certain size. Principally due to a requirement for small plant footprint, external MBR systems have been used since the early 1980s for this duty, with increased use of submerged MBR systems from the late 1990s. According to the above-mentioned survey, the present number of these MBR installations is no fewer than 74, whose capacity amounts to 11 MLD in total.

MBRs have also been used for many types of industrial process wastewater treatment, with around 1610 installations as at March 2009. Of the 1270 plants for which relevant information was available (Fig. 1.11), the food industry, stock

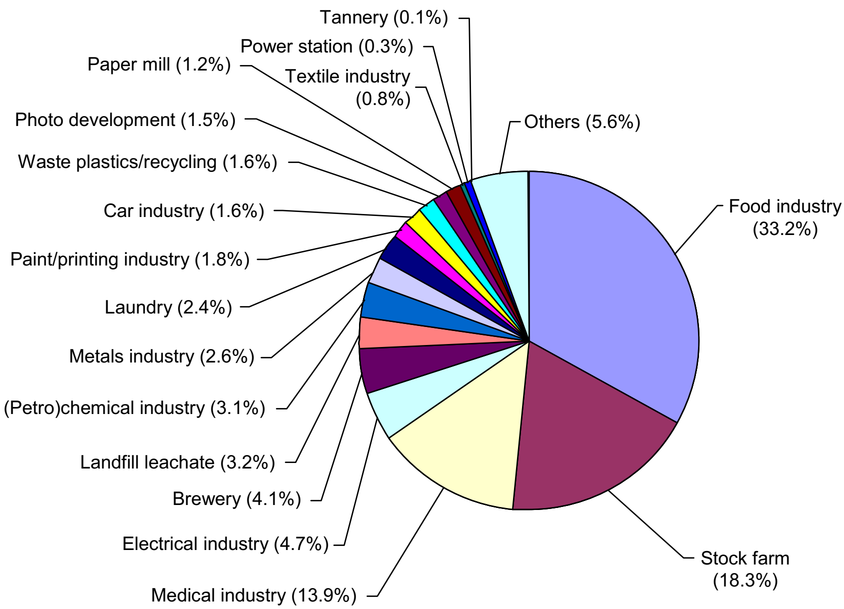

FIG. 1.11 Industrial process water treatment in Japan: percentage of MBR installations by industry, prepared from survey responses from eight membrane manufacturers and engineering companies (Asahi Kasei Chemicals, Daicen Membrane Systems, Hitachi Plant Technologies, Kubota, METAWATER, Mitsubishi Rayon Engineering, Sumitomo Electric Fine Polymer and Toray Industries), incorporating data from 1270 MBRs.

farms and medical industry were the three largest contributors, accounting for around 33%, 18% and 14% of the total number of installations, respectively.

Although there are a large number of commercial MBR installations, they are nonetheless relatively small in scale, especially where municipal WwTPs are concerned: all the existing plants have a capacity of less than 5 MLD. Possible reasons for this include relatively lax effluent regulations (e.g. nutrient removal is required only for limited areas) and the fact that there is little requirement for water reuse, since for most of the area water scarcity is not an issue. In addition, as with most other parts of the world, the somewhat conservative nature of policy-makers and engineers can make it challenging for a new technology to be accepted.

In 2009, following extensive canvassing of opinion, the MLIT published the *Guideline for Membrane Technology*. In the same year, the Ministry launched a national project – *Advance of Japan Ultimate Membranebioreactor Technology Project (A-JUMP)*. In this project, two demonstration MBR plants were constructed, one 5 MLD plant being retrofitted from an existing conventional activated sludge process (CASP) and the other 0.24 MLD plant constructed in a pumping station as a satellite treatment system. Data and experience from these plants are to be fed into a revision of the Guideline in 2010. In addition, a further project is underway, in which the train of an existing municipal WwTP will be retrofitted in 2011 with a 60-MLD MBR. The New Energy and Industrial Technology Development Organization (NEDO) also commenced several membrane-related national R&D projects in 2009, including *Developments of low-energy MBRs*. This indicates that energy consumption and costs are still regarded as an issue constraining the widespread acceptance of MBR technology in Japan.

1.4.2.6. Malaysia

The 1974 Environmental Quality Act (EQA) provides the framework for environmental regulation in Malaysia. Under its recent revision, the Environmental Quality (Industrial Effluent) Regulations (2009) Act regulates the discharge or release of industrial effluent or mixed effluent onto or into any soil, or into inland waters or Malaysian waters. 'Industrial effluent' refers to any waste in the form of liquid or wastewater generated from manufacturing processes (Table 1.5) including the treatment of water for water supply or any activity occurring at any industrial premises. 'Mixed effluent' refers to any waste in the form of liquid or wastewater containing both industrial effluent and sewage.

With respect to MBR implementation, the regulations specifically require that industrial effluent and mixed effluent follow the stipulated 'Standard A' and 'Standard B' (Table 1.6). Standard A is generally applicable to activities and industries that are sited within, or in the near vicinity of, catchment areas and is more stringent. Standard B is generally applicable to both industrial and development activities throughout the country. Table 1.6 lists the acceptable

TABLE 1.5 Main industries in Malaysia and Issues Pertaining to Industrial Pollution (Phang, Foo, & Lee, 2001)

Industry	Industrial Pollution
Rubber products processing	Issues include contaminants such as hydrogen sulphide in wastewater, odour control and stack emissions.
Palm oil industries	Anaerobic ponds are most common means of wastewater treatment. Enhanced wastewater treatment required near municipalities and catchment areas, as well as noise control, sludge treatment and air pollution (stack) control systems/technologies, and waste recycling technologies to convert fibre and trunk wastes into value-added products.
Oil, gas and petrochemicals	Main pollution problems include radioactive sludge disposal, recovery of used oils and ship-based sludge treatment and disposal.
Electronics/ electroplating	Majority of electroplating and metal-finishing industries are small to medium-sized enterprises. Wastewater effluent with heavy metal contaminants is routinely disposed of in domestic sewage systems without prior treatment. Cost-effective wastewater systems, technology to recover heavy metals from wastewater effluent, and toxic sludge treatment and recycling technologies are needed.
Food and beverage processing	A large percentage of the country's total wastewater effluent is released by food processing companies. Non-compliance is a direct result of the lack of appropriate treatment technology, over-utilized capacity and poor treatment system maintenance. Wastewater stream has high levels of BOD, COD, FOG, and suspended solids.

conditions for discharge of industrial effluent or mixed effluent of Standards A and B. In Malaysia, this regulation, together with cost, is the main driver for any decision to implement wastewater treatment technology.

Despite various research projects, as yet no MBR technology has been installed in Malaysia for either municipal or industrial treatment. This is due to a number of factors:

- the perception that newer technologies such as MBRs are very expensive to install and operate;
- the significant capital investment required by state-of-the-art MBRs, which may be beyond the capability of privately operated wastewater treatment operators;
- the possible lack of perceived importance of environmental issues, together with regulatory challenges, such that the drive to invest in proven and better quality technology is also impeded;
- the general lack of awareness of the technology among government policy-makers; and

TABLE 1.6 Parameters for Standards A and B

		Standard	
Parameter	**Unit**	**A**	**B**
Temperature	°C	40	40
pH	–	6–9.0	5.5–9.0
BOD	mg/L	20	50
COD	mg/L	50	100
Ammoniacal nitrogen	mg/L	10	20
Suspended solids	mg/L	50	100
Oil and grease	mg/L	1.0	10
Cadmium	mg/L	1.01	0.02
Mercury or chromium	mg/L	0.005	0.05
Other heavy metals	mg/L	Between 0.05 and 2	Between 0.1 and 5

- the reduced economy of scale for MBR installations compared to conventional technologies in various parts and different industries in Malaysia, making the technology difficult to justify.

For the future, however, the national sewerage company Indah Water Konsortium has been entrusted with the task of developing and maintaining a modern and efficient sewerage system for all Malaysians, and this may well help to raise the profile and drive installations of MBRs in Malaysia. Indeed, subject as always to favourable cost–benefit analysis, it can be said there are potentially many opportunities for installing new treatment processes, including MBRs.

1.4.2.7. Singapore

Over the course of four decades, Singapore has overcome water shortages despite its lack of natural water resources, and flooding and pollution in its rivers in the 1960s and 1970s. Significant strategic investment in water technology and management has meant that, today, the nation has a robust, diversified and sustainable water supply from four different sources known as the Four National Taps (water from local catchment areas, imported water from Malaysia, reclaimed water known as *NEWater* and desalinated water). It is in the latter two that membrane technology has played a vital role, and particularly in the development of *NEWater*. By the end of 2010, about 30% of total water consumption in Singapore will be provided by *NEWater* supply.

The national water agency, Public Utilities Board (PUB), is responsible for the collection, production, distribution and reclamation of water in Singapore. PUB's work focuses on increasing water resources, keeping costs competitive, and managing water quality and security. The organization invests heavily in research and development – from fundamental research work ('first level'), to pilot-scale studies ('second level') and, potentially, demonstration-scale studies ('third-level' research). Third-level research allows working plants to be constructed and operated so that operational and process challenges, if any, can be identified and managed in a full-scale plant.

MBR systems are made particularly attractive in Singapore, where land is scarce, by their small footprint, and PUB embarked on an MBR technology study programme in 2002. Three pilot systems (GE Zenon, Mitsubishi Rayon and Kubota) were set up in early 2003 and evaluated for over two years (Section 3.2.1.3). These trials generated valuable information on the design and operation of the MBR systems under a tropical environment. After more than two years of pilot testing at Bedok Water Reclamation Plant (WRP), reliability of MBR technology and its effectiveness in producing better feedwater quality for the production of *NEWater* through MBR-RO was demonstrated; operating conditions were optimized in the pilot-scale phase so that a robust design could be specified in the next stage.

PUB has constructed a 23-MLD MBR demonstration plant at Ulu Pandan WRP to establish the feasibility of municipal-scale operation for future development of water reclamation and *NEWater* plants; operational parameters applicable to Singapore's tropical environment have been established. The demonstration plant (Section 5.3.1.7), which treats domestic wastewater, has been in operation since December 2006 and the product supplied to industries for high-purity process feedwater. Most recent optimization has demonstrated sustainable operation at an overall energy consumption of less than 0.4 kWh/m^3, possibly the lowest of any reported. The MBR technology, coupled with RO (Fig. 1.12), produces better and consistent quality *NEWater* for industrial use. Since this process obviates the final sedimentation tank of a conventional process as well as provides better quality feedwater to RO in *NEWater* production, it can potentially reduce the production cost of *NEWater* by about 20%. MBR permeate is also supplied to industries as Industrial Water (IW) for their other process needs.

PUB is to set up another MBR plant with a capacity of 68 MLD at Jurong Water Reclamation Plant. The product water will be used as IW and supplied to industries. The plant is expected to be commissioned by 2011. Further investigations are underway for an MBR plant coupled with a UASB-MBR-RO system to reclaim industrial wastewater at the Jurong plant. Other MBR technologies such as those of Siemens, Ultra-Flo, Huber, Asahi Kasei, Memstar, Toray, Norit and Koch, have also been tested in PUB's various WRPs.

MBR technology is also gaining acceptance in the private industrial sector in Singapore. Sembcorp Industries, a leading centralized utilities and energy

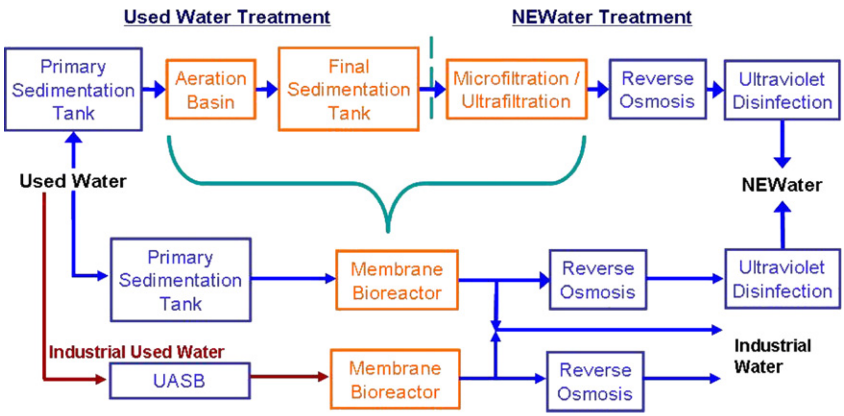

FIG. 1.12 *NEWater* and Industrial Water production process.

provider at Jurong Island, is applying MBR technology for complex industrial wastewater treatment. In 2009, Sembcorp announced the opening of its Integrated Wastewater Treatment Membrane Bioreactor Plant. The plant, an expansion of the company's integrated wastewater treatment system in the Sakra area of Jurong Island (Singapore's petrochemical cluster), was Sembcorp's second plant in Singapore employing MBR technology. The wastewater plant receives 10 streams of wastewater discharged from chemical process plants. The wastewater, with an average chemical oxygen demand (COD) of 3765 mg/L, contains a variety of chemicals such as alcohols, aldehydes, organic acids, phenol, cyanide and sulphides. The wastewater was originally treated using anaerobic expanded granular sludge blanket (EGSB) and aerobic moving bed bioreactor (MBBR) technologies. Integration of an MBR into the aerobic treatment stage expanded the plant's capacity by an additional 2.9 MLD. The MBR plant consists of an aerobic activated sludge reactor and a membrane tank, with 480 HF membrane modules submerged in the membrane tank. The MBR receives anaerobic effluent from the EGSB and aerobically reduces the COD from 1508 mg/L to 20 mg/L. The treated effluent is further processed in a water reclamation plant to produce high-quality demineralized water which is then reused as high pressure boiler feed.

Sembcorp also applies MBR technology for treating a wastewater stream containing 3–5% of salinity coupled with COD as high as 14,000 mg/L. The MBR used is a pressurized sidestream type with a design capacity of 1.3 MLD. A total of 30 vessels housing tubular ultrafiltration membranes are linked to the bioreactors. Operating under thermophilic conditions (with temperatures between 45 °C and 55 °C) and with full sludge digestion, the MBR produces almost zero sludge.

A large petrochemical company on Jurong Island completed MBR pilot trials and adopted this technology for treating the waste stream for attaining better effluent discharge standards. The full-scale plant with a capacity of 11 MLD is expected to commission in 2010/2011.

1.4.2.8. Germany

The great majority of German sewage treatment plants have been upgraded in recent years in accordance with the demands of the Wastewater Ordinance (AbwV, *Abwasserverordnung*), and in particular have reached a high purification level with regard to the elimination of the nutrients nitrogen and phosphorus (Steinmetz, 2008). The overall annual investment in wastewater treatment infrastructure is relatively stable, although it decreased from €6.9 billion in 2000 to €4.6 billion in 2006. This decrease arises because most of the required investment to reach EU legislation goals had already been made (ATT, 2008).

There are nearly 10,000 municipal WwTPs in Germany. A total of 96% of all inhabitants are connected to public sewer systems. Ninety-nine percent of the total volume of 9.4 billion m^3 of discharged domestic wastewater is treated by biological nutrient removal (BMU, 2010a), corresponding to the highest standard of the EU Directive (91/271/EWG, 1998) for the treatment of municipal sewage. The remainder of household wastewater is mostly treated in decentralized small-size or package-plant WwTPs. Industrial wastewater treatment also shows a clear trend towards higher standards. Between 1991 and 2004, the amount of wastewater receiving only mechanical treatment decreased from 31 to 9%, while at the same time the amount of biologically treated wastewater increased from 38 to 67% (BMU, 2010b).

The first full-scale municipal MBR installations in Germany of >500 population equivalent (p.e.) went into operation in 1999. Since then, their number has increased to at least 15, with capacities from 700 to 80,000 p.e. (see Table 1.7). A great many of them are found in the federal state of North Rhine-Westphalia (NRW). As with NRW, many regions in Germany are densely populated, and the water resources are under stress from industrial, agricultural and civic utilizations. Although water scarcity is uncommon, maintaining a high water quality is one of the major goals of the national environmental policy. In particular, MBRs have been established in specific geological situations where pathogens in the effluent of WwTPs might contaminate the aquifer via rivers that percolate entirely, as at Monheim or Glessen, or where the receiving rivers are used as bathing waters, as at Hutthurm.

Package-plant MBRs with sizes from 4 to 40 p.e. have been installed in larger numbers, many of them in environmentally sensitive areas. In some cases, MBR technology is the only option approved by local environmental authorities for decentralized wastewater treatment.

Industrial MBR applications are well known in Germany for the treatment of landfill leachate, and also for wastewater treatment in the pharmaceutical and food and beverage industries among others (Rosenwinkel, Wagner, & Nagy, 2000); between 50 and 60 full-scale installations were in operation in

TABLE 1.7 MBRs for Municipal Wastewater Treatment in Germany

Plant	Treatment Capacity (p.e.)	Membrane Supplier	Year of Commissioning	Federal State
Nordkanal	80,000	GE	2004	NRW
Hutthurm	22,000	Huber	2008	Bavaria
Makranstädt	12,000	GE	2000	Saxony
Eitorf	11,625	Kubota	2005	NRW
Seelscheid	10,500	Kubota	2004	NRW
Monheim	9700	GE	2003	Bavaria
Konzen	9700	Kubota	2007	NRW
Glessen	9000	GE	2008	NRW
Rurberg	6200	Kubota	2005	NRW
Rödingen	3000	GE Koch Membrane Systems	1999	NRW
Schramberg	2600	GE	2004	Baden-Württemberg
Xanten-Vynen	2000	A3	2005	NRW
Büchel	1000	Kubota	1999	NRW
Knautnaundorf	900	Martin Systems AG	2001	Saxony
Simmerath	700	Koch Membrane Systems	2003	NRW

2007 (Lesjean & Huisjes, 2008). Maritime use on board large ships is another field of application. The nature of the wastewater, the demand for additional treatment, water reuse, site restrictions and the opportunities for retrofitting to existing WwTPs can be seen as major drivers for MBR technology implementation in the industrial sector.

In general, membrane technology is publicly perceived as a future key technology for the water sector (Pinnekamp & Friedrich, 2006; UBA, 2007). The German Association for Water, Wastewater and Waste DWA has established several working groups to prepare technical guidelines on the application of MBRs for municipal and industrial wastewater treatment. Numerous initiatives and institutions have promoted membrane technology, and MBR technology in particular, e.g. DEBRANE (2010), an R&D network for promotion of membrane technology, BMBF (2009), a federal ministry for education and research, and SIMAS (2010), a training institute for membrane applications in wastewater treatment. Many German SMEs in the water sector, membrane suppliers and construction companies, are export-orientated. The successful application of new technologies at home might be seen as a means of entering overseas markets. In addition to environmental policies on water quality, capacity building and the endorsement of new technologies in the water sector can be seen as primary strategic objectives for the public funding of MBR technology (Uhlenberg, 2007). Earmarked assets from the wastewater levy imposed by wastewater discharge regulations (AbwAG, 1976) are used to fund measures to improve water quality. MBR projects received a share of the funding which initially reduced the economic risk involved to the end user in investing in a new technology. Since MBR technology has emerged from its early stage of development, the amount of funding has been commensurately reduced over the years.

In view of the high standard and degree of wastewater treatment, the domestic market for wastewater treatment in Germany can be considered as relatively mature. Therefore opportunities for the application of new MBRs, be it as greenfield or retrofitted installations, are not widespread. Although MBRs have known advantages, they also must compete against other well-established treatment technologies, such as UV disinfection or tertiary filtration. It is then often an inclination towards innovative technologies, together with the strict requirements on water quality enforced at a regional level, that encourages the application of MBR technology.

1.4.2.9. Italy

In 2005, the size of the Italian MBR market was estimated as \$12.1 million with a compound annual growth rate (CAGR) of 11.5% until 2011 (Frost & Sullivan, 2005). This, when compared to the total European market size estimated as \$90.4 million in 2006 with a CAGR of 7.9% until 2013, implies that Italy had around one-seventh of the European market at that time and, as such, must be considered one of the key regions for MBR implementation in Europe. Moreover, a report by the AMEDEUS consortium in 2009 (Lesjean, 2009) indicated that the cumulative number of MBR plants in Italy grew from 84 at the end of 2005 up to 149 at the end of 2008, equating to a sustained growth rate, in terms of the number of MBR plants over this period, of 21% per year.

The main driver in the Italian municipal market is European regulation for nutrient removal in sensitive areas, together with the need for increased capacity in ageing, confined WwTWs. For the industrial sector, implementation is driven by increasingly stringent requirements towards pollution control, especially those concerning direct discharge into water bodies. Tourist resorts in southern Italy in particular are increasingly favouring MBRs as a non-intrusive solution with reusable effluent quality and little impact on bathing waters.

Italy was one of the first European regions to embrace MBR, and some of its oldest plants have already reached a 10-year life. The market is characterized by medium- and small-sized municipal and industrial MBR systems with a few exceptions at larger scale. Examples of these are wastewater plants within urban agglomerations such as Rome, Milan, Venice, Bologna, Trento and, the largest of all, Brescia (Section 5.3.1.3). Coastal areas of special beauty such as Elba, Albarella, Capri and Santa Margherita have also adopted MBRs as the best available technology. In some mountain regions with difficult access such as Siffiano, San Martino di Castrozza or Santa Anna di Alfaedo membrane technology has been selected due to its superior effluent quality and reduced footprint.

The industrial market mainly comprises food and beverage industries, with capacities in the range of up to 0.1 MLD. Other contributions are from the oil and gas, tannery, tanker washing and laundry industries. Only rarely do some industrial effluent treatment installations exceed 5 MLD in capacity, in which cases MBR technology appears to have been the only technically possible solution. These projects are complex in nature, often accompanied by lengthy pilot trials and awarded via qualification and tendering processes.

MBR technology is nowadays acknowledged by most municipal end users and many industrial clusters in Italy as being viable for wastewater treatment. Major utilities groups such as Acea and Hera have adopted MBRs for a number of municipal projects, and their value is recognized by process consultants and contractors. Successful process contractors range from national environmental engineering firms such as Ladurner Acque, Atzwanger, Sernagiotto or Dondi, to international consortia such as Siba (Veolia), Ondeo-Degremont or Severn Trent. The main technology suppliers to date have been GE and Kubota, with Siemens, Koch Membrane Systems (KMS) and Toray also having installations in the country. The municipal sector is likely to continue embracing MBR technology as a solution for plant enlargement, tertiary treatment and reuse. There is no reason to project a slowdown in MBR acceptance in the years ahead, other than that incurred by the global economic slowdown. It is expected that most projects will be in the range of up to 3 MLD in capacity, with perhaps one project per year above 3 MLD but unlikely to exceed 10 MLD.

1.4.2.10. The Netherlands

At the turn of the Millennium, a plan was submitted for the total management of MBR implementation and commercialization in the Netherlands. This 10-year plan was supported by many Dutch water authorities and the *Stichting Toegepast Onderzoek Waterbeheer* (STOWA), the Dutch Foundation for Applied Water Research, and a special innovation fund was set up to support MBR development. DHV Water led the implementation of this plan with the pioneering Beverwijk comparative four-year research programme, followed thereafter with the execution of the demonstration plant at Varsseveld. GE

Zenon, Toray and Xflow were respectively represented at MBR Varsseveld, MBR Heenvliet (Section 5.4.1.1) and MBR Ootmarsum (Section 5.4.1.3), the latter being the most recent installation. Up until the final STOWA reports from Varsseveld in 2006/7, the Netherlands was internationally recognized for the country's commitment to MBR technology. However, the planned larger Hilversum installation did not progress beyond the demonstration stage, and the pace of implementation of the technology has possibly slowed in recent years.

The main driver for the Netherlands MBR programme was water quality improvement and footprint reduction, both of which have been verified in the last five years of operation of Varsseveld and the other demonstration plants. The effluent quality from each MBR was excellent, as expected, although maintaining stable operation has been found to be more challenging. Varsseveld has been optimized on total energy, with respect to pre-treatment and MBR operation, without consideration of sludge dewatering or other sludge management operations. The optimum energy demand for the plant has been reported as $<0.65\ kWh/m^3$ during dry weather flows (DWF) and $<0.45\ kWh/m^3$ at peak (rain weather) flows. This underlines one of the major issues with MBRs, this being the design and operating parameters most appropriate when flows fluctuate due to infiltration. Also, the low HRTs of <4 h, as incurred during storm flows in the system, tend to cause problems regarding the maximum treatment rate.

For small-scale industrial MBRs, where the drivers are based on cost savings, space requirements, water reuse, discharge quality and robustness of operation, adoption of the technology is more widespread. Compared to the three demonstration municipal MBR installations, there are approximately 30 industrial MBRs to date. As freshwater becomes more costly, and industry is restricted on the amount of groundwater that can be extracted, water reuse projects involving MBR technology are becoming more popular, particularly in the food, leachate, waste handling and tanker washing industries. Plants generally vary in capacity from 0.25 to 2.5 MLD, and the emphasis in the food industry is on water recovery for process operations.

1.4.2.11. Spain

Spain embraced MBRs later than other European regions such as the United Kingdom, Germany, France or Italy. The oldest industrial plants date from 2000, and the first municipal plant was installed in 2003. However, the Spanish market developed much faster than in neighbouring countries, boosted by membrane technology licensees and manufacturers. The public administration proved eager to issue and accept public tenders based on the MBR process, prompting process consultants and main contractors to include MBRs in their portfolios. Nowadays, Spanish MBR systems are mainly medium-sized municipal and small-sized industrial installations. However, there is also an unusual number of large-scale municipal plants, as compared with other European countries. Examples of the latter in the 20–50 MLD capacity range are San Pedro del Pinatar, Sabadell and Gava, and in the 10–20 MLD range are Alcoi, Arenales (all near the Mediterranean shore of continental Spain), and Tamaraceite (in the Canary Islands). Coastal areas of special beauty such as Agulo, Agaete, Valldemossa, El Campello, La Tinas or Cudillero, with capacities <5 MLD, adopted MBRs as the best available technology given the technology's landscaping options, small footprint and effluent quality.

The industrial market is represented to a large extent by food & beverage projects, with capacities up to 11 MLD. Wineries and fruit processing plants account for most of the reference sites. Other contributions come from the leachate, cosmetics and pharmaceutical sectors.

Spain's largest projects have been awarded to main national contractors such as Drace, Cadagua, Befesa, Aqualia or Acciona. The major MBR technology process consultants and integrators are HERA-Amasa and ITT, with the medium contractors being Aquagest, Comsa-Deisa, DAM, HERA-Amasa, MP Medioambiente, Dinotec, Intersa or Integra Water, among others. The main technology suppliers to date have been Kubota, GE Zenon and Toray although other suppliers such as Siemens, KMS, Weise Water and Huber all have a presence.

The estimated size of the Spanish MBR market in 2005 was $13 million (Frost & Sullivan, 2005), with a projected compound annual growth rate (CAGR) of 16.8% until 2011. Although similar in size to the Italian MBR market, the Spanish market was considered less well developed at that time. A report by the AMEDEUS consortium (Lesjean, 2009) indicated that the cumulative number of MBR plants in Spain grew from 47 at the end of 2005 up to 111 at the end of 2008. This represented an equivalent sustained growth in terms of number of MBR plants of 33% per year over that period. However, the impact of the global recession on the MBR market appears to have been more profound in Spain than in other EU nations, such that no new municipal projects appear to have been implemented since 2009.

A primary driver for MBR technology in the Spanish municipal market is wastewater reuse, especially in the regions around the major economic centres on the Mediterranean coast, Canary and Balearic Islands. Over the last decade, ambitious European, national and regional plans allocated generous budgets to improve the quality of the water infrastructure, including desalination plants and wastewater treatment facilities. Beyond environmental protection, new legislation in Spain aims to increase water availability by the efficient and safe use of regenerated water. Furthermore, the European regulations for nutrient removal in sensitive areas (such as bathing waters and continental water bodies) are also important drivers, together with the need for increased capacity in ageing, confined WwTPs.

For the industrial sector reuse, limited footprint and rising pollution discharge costs remain the major incentives for installing MBRs. In recent years, golf resorts and real estate in southeastern Spain have also seen a significant take-up of MBRs as a solution providing reusable effluent of a quality suitable for irrigation. However, as with other regions of the world, the Spanish wastewater treatment industry is highly cost-sensitive, both at capital investment and at operation cost levels. MBR technology suppliers thus face challenges in providing a technology solution of comparable cost effectiveness to competing technologies, given the constraints of the generally higher energy demand and the costs incurred by membrane supply and replacement. This is to some extent exacerbated by the general perception that there remain a number of unknown quantities pertaining to the MBR technology, specifically membrane life, chemical resistance and sensitivity to pre-treatment. Furthermore, operators who are used to managing peaks via overflows and bypasses do not necessarily feel confident with the absolute nature of the MBR process. In this sense, tertiary membrane filtration or hybrid systems are seen as more reliable designs.

The Spanish municipal sector is likely to continue to adopt MBRs as a solution for plant enlargement, tertiary treatment and reuse. Moreover, it is expected that decentralized systems will become more significant in years to come. The future for industrial plants and real estate is more uncertain given the current delicate financial situation of the private sector.

1.4.2.12. United Kingdom

The MBR market in the United Kingdom and Ireland is, in comparison to other countries, relatively mature, with over 10 years operational experience in full-scale plants in both the municipal and industrial sectors. Europe's first full-scale municipal MBR plant at Porlock WwTW, England, was commissioned in February 1998 (Section 5.2.1.1). Other large-scale MBR plants were commissioned over the next two years, including the 7100 m^3/d industrial plant at Ballyragget, Ireland in May 1999 and the visually impressive municipal MBR plant at Swanage WwTW, England in June 2000 (Section 5.2.1.2). Whilst the early MBR market was dominated by the suppliers Kubota and Zenon, the number of membrane suppliers has since increased as the MBR market has matured such that most major MBR membrane manufacturers now have references in either the UK municipal and/or industrial sectors.

In the municipal sector, the main legislative drivers for the application of MBRs in the United Kingdom and Ireland relate to EU legislation: the Urban Wastewater Treatment Directive, the Water Resources Act, the Bathing Water Directive and the Shellfish Directive. In the industrial sector, the Integrated Pollution Prevention and Control (IPPC) regulations, with the associated references to Best Available Technologies (BAT) and the Enhanced Capital Allowance (ECA) scheme, are also driving the market. The ECA scheme (ECA, 2010) is a key part of the British Government's programme to manage climate change. The scheme allows businesses to claim 100% first-year capital allowances on their spending on qualifying plant and machinery against their taxable profits in the period they make the investment. This can deliver a helpful cash flow boost, by reducing the tax charge, and a shortened payback period. To qualify for an ECA, membrane filtration systems must be able to produce a final effluent suitable for water reuse and meet published criteria. One of the key criteria is that the system must allow at least 40% of the treated effluent to be reused on site. The food and beverage sector in particular has embraced the ECA scheme with a number of schemes successfully meeting the required criteria.

The installed membrane surface area of Kubota-based MBR plants in the United Kingdom and Ireland has been increasing roughly linearly since the late 1990s (Fig. 1.13). The 170,000 m^2 membrane area represents a total of 54 Kubota plants, compared with 25 Zenon across the same region. For both technologies there are roughly two municipal installations for every industrial one.

It is anticipated that the future municipal MBR market will be dominated by those sites where space is limited, the effluent standards are being tightened and the existing assets are at the end of their operational life. This market is thought likely to offer a steady number of opportunities, with the increase in the installed membrane surface area dependent on the flow being treated at the most appropriate sites. In the industrial sector, it is expected that the MBR market will increase year on year, reflecting the increasing costs of potable water supply and wastewater disposal. The ECA allowances would be expected to continue to reduce the investment payback periods to acceptable timeframes.

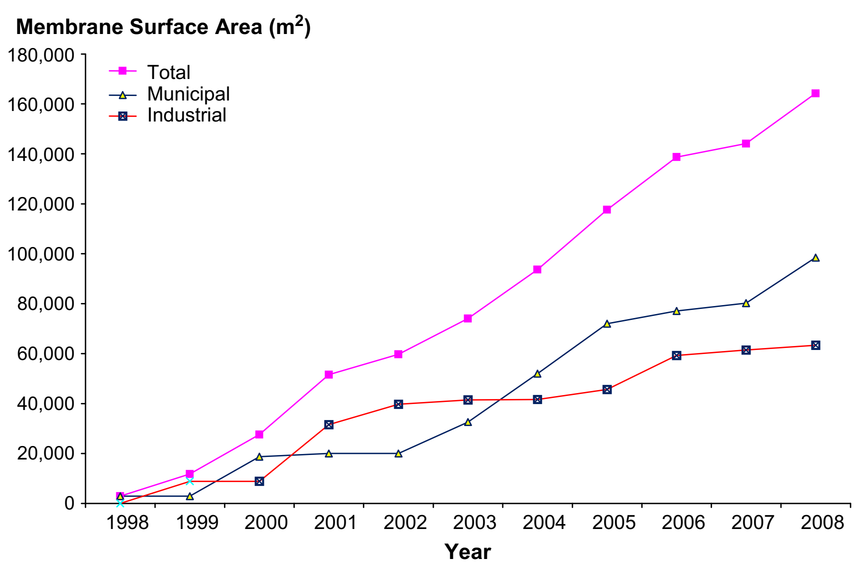

FIG. 1.13 Cumulative installed membrane surface area in the United Kingdom and Ireland, Kubota plant.

The financial benefits are being augmented by the growing acceptance of recycling treated effluents, and by generally decreasing MBR costs.

Notwithstanding improvements made in capital cost and energy demand, the main threats to the United Kingdom and Ireland MBR market are the high capital and operational costs associated with the process and, more recently, its perceived carbon footprint. However, comparison of process technologies conducted in the past have not always taken account of the complete process envelope including, for example, sludge processing and management costs and the embedded carbon associated with the more extensive infrastructure (i.e. the concrete tanks) of the larger conventional plant. MBR membrane manufacturers and systems designers continue to reduce the system costs to make the process more competitive, and the difference in energy demand between MBRs and conventional processes is steadily decreasing, such that it is likely that parity in carbon footprint between conventional and MBR wastewater treatment processes is attainable – if not already achieved. The increased emphasis on carbon accounting by UK water utilities makes this balance critically important for wider implementation in the municipal sector.

1.4.2.13. United States

Based on a 2009 survey of eight major MBR manufacturers in the USA, the total number of large municipal MBR installations (with capacity greater than or equal to 1 MGD, or 3.8 MLD) has grown more than fivefold from 13 to 68 in the five years between 2004 and 2009. In the same period, the cumulative capacity of the MBR installations has also grown more than fivefold from 40 MGD (152 MLD) to 204 MGD (773 MLD). By the end of year 2011, 85 MBR installations, with capacity of 1 MGD or greater, are expected to be in operation in the USA, with total treatment capacity of 345 MGD (1302 MLD). As shown in Fig. 1.14, the majority of the installations (66 out of 85) are under 5 MGD (19 MLD) in capacity but a few larger MBR installations are either in operation or are under contract. These installations include the 11 MGD (42 MLD) capacity MBR at the Johns Creek Environmental Campus in Georgia, as well as the 39 MGD (148 MLD) MBR in King County, Washington.

Figure 1.15 shows the breakdown of large municipal MBR installations (capacity $\geq$ 1 MGD) by US state, and indicates that California, Georgia, Washington, Arizona and Florida have a higher number of MBR installations than others. These five states have developed regulations or guidelines that strongly encourage water reuse as a strategy for conserving water resources by specifying water quality requirements, treatment processes, or both, for a full range of water reuse applications. By the end of 2011, California and Georgia will, respectively, have an installed treatment capacity of 63 and 51 MGD (240 and 190 MLD) using MBR technology. For the 85 installations in the USA, improved water quality and reliability requirements (55%) and footprint limitation (36%) have been the key drivers behind selecting the MBR process.

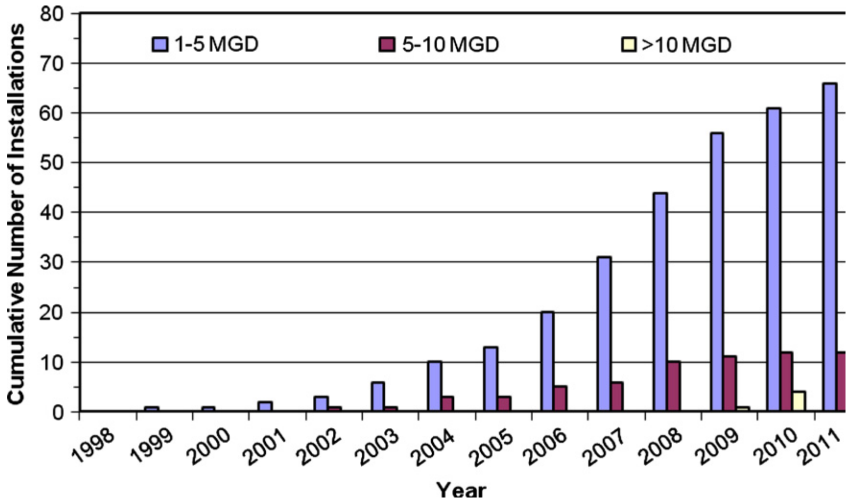

FIG. 1.14 Cumulative number of municipal MBR installations in the USA.

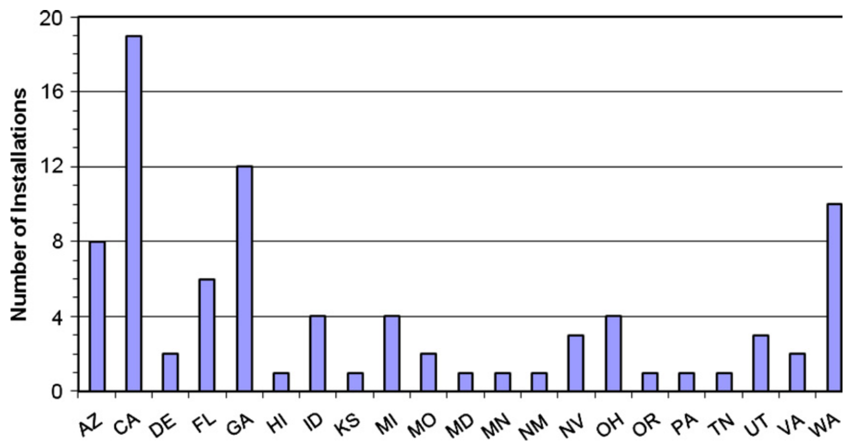

FIG. 1.15 Number of MBR installations in various states in the USA. *(Oppenheimer, Rittmann, DeCarolis, Hirani, & Kiser, 2010.)*

Two major federal policies that affect recycled water projects include the Clean Water Act (CWA) and the Safe Drinking Water Act (SDWA). Although these federal acts promote the use of recycled water, they do not directly govern water reuse practices in the USA. Since the CWA regulates the discharge of pollutants into navigable bodies, it encourages the use of recycled water to reduce pollutant discharge into rivers, lakes and wetlands. The SDWA regulates the drinking water quality in the USA so that any project using recycled water to augment surface reservoirs or groundwater aquifers for indirect potable reuse needs to comply with SDWA.

As of November 2002, 25 states had adopted water reuse regulations, 16 states had developed guidelines or design standards, and nine states had no regulations or guidelines (US EPA, 2004). The states of California, Florida, Hawaii and Washington have adopted water reuse regulations and have also developed regulations and guidelines for the use of recycled water for groundwater recharge and indirect potable reuse. Since these applications require high-quality effluent, low in organic matter and microbial contaminants, they may promote use of membrane technologies such as micro/ultrafiltration, MBR and RO.

The primary barriers to the use of recycled water in the USA include numerous federal, state and regional regulations, capital investment costs and public perception. In most cases, treated wastewater for reuse applications is subject to more stringent requirements than the treated wastewater for river discharge. This results in additional treatment requirements and related capital and O&M costs for recycled water facilities. There is also a requirement for initial investment in distribution systems to deliver recycled water to customers and end users. Public perception has also prohibited the use of recycled water in indirect potable reuse applications due to the 'toilet to tap' stigma, and such perception makes it difficult for regional authorities to approve recycled water projects. However, recent approval of projects such as the City of San Diego's advanced water treatment demonstration plant has shown increased public acceptance of indirect potable reuse (City of San Diego, 2006).

1.5. RESEARCH

A review of the research conducted in MBRs for wastewater treatment over the past two decades, based on Scopus (2010), reveals distinct trends. A search of the published journal paper titles between 1990 and 2009 based on the primary terms listed in Table 1.8 identifies 1450 publications, with a year-on-year accumulation of 20% from 1994 onwards. Within this data set, a search of papers in key subject areas – the secondary terms in Table 1.8 – reveals the subject of fouling to be the most prevalent. Along with the topic of micropollutants, papers concerning fouling are also the ones growing the most rapidly in number – around 36% growth p.a., according to this analysis (Fig. 1.16).

The research trends shown in Fig. 1.16 are to a large extent corroborated by a word cloud analysis (Wordle, 2010) on the keywords from all 1450 papers (Fig. 1.17). A word cloud expresses the analysis as a graphic where the font size of the keyword represents the frequency with which it arises; certain generic terms are necessarily excluded from the analysis. According to this graphic, it is the subject of fouling which has attracted the most attention from the academic community, particularly as it relates to extracellular polymeric substances (EPS).

TABLE 1.8 Search of Journal Paper Titles Published 1990–2009

Primary Terms	Secondary Terms*
'MBR'	1. 'Fouling' OR 'Biofouling'
OR	2. 'Clogging' OR 'Sludging'
'Membrane bioreactor'	3a. 'Hollow fibre' OR 'Hollow fiber'
OR	3b. 'Flat sheet' OR 'Flat plate' OR 'Plate and frame'
'Membrane separation bioreactor'	3c. 'Tubular' OR 'Multi-tube'
OR	4. 'Nanofiltration' OR 'MBR-NF' OR 'Membrane distillation'
'Membrane biological reactor'	OR 'MD' OR 'Forward osmosis' OR 'FO'
AND	5. 'Micropollutants' OR 'Endocrine disrupting compounds' OR
('Wastewater' OR 'Sewage')	'EDC' OR 'Pharmaceutical' OR 'Personal care products' OR 'PCP'

** Truncated terms also included.*

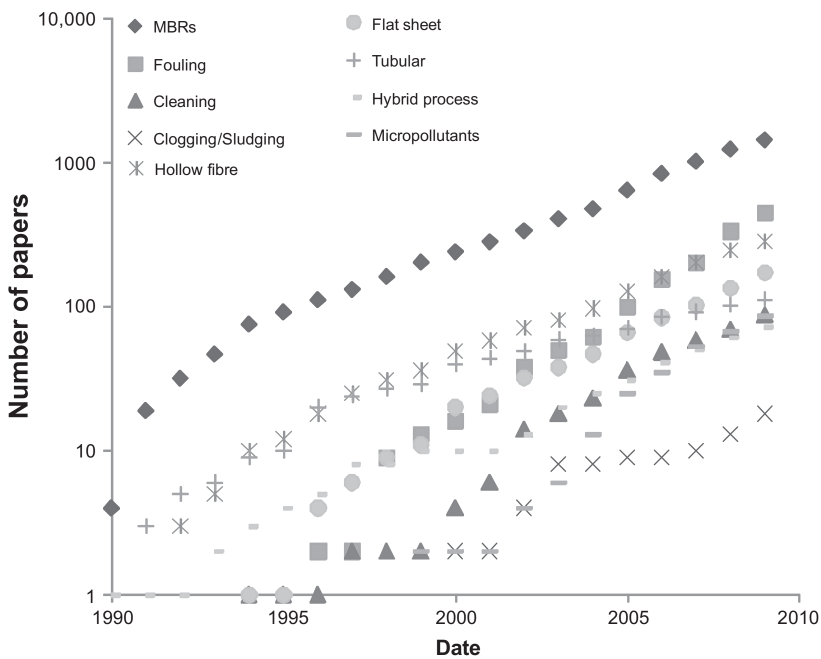

FIG. 1.16 Research trends in MBRs: number of publications in key subject areas according to Scopus (2010).

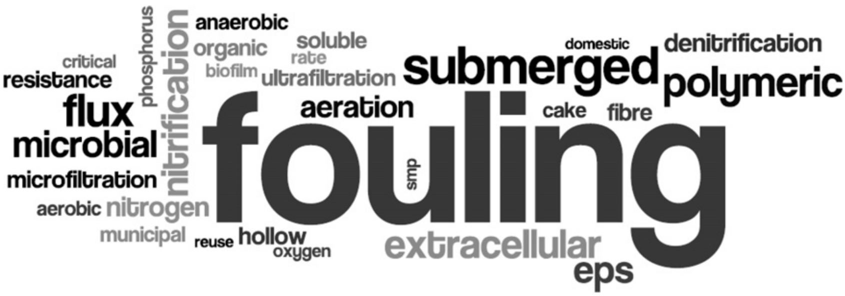

FIG. 1.17 Word cloud produced from MBR research keywords, excluding (a) generic technical terms for MBR studies and their plurals (*Activated sludge, Bioreactor, Biological, Filtration, Liquor, MBR, Membrane, Mixed, Products, Reactor, Retention, Separation, Substances, Technology, Treatment, Wastewater*), and (b) generic terms for scientific studies and their plurals (*Analysis, Application, Comparison, Effect, Evaluation, Impact, Influence, Investigation, Performance, Research, Study, Activity, Based, Characteristics, Compounds, Concentration, Conditions, Different, Factors, High, Low, New, Novel, Operation, Operational, Plants, Process, Production, Properties, Removal, Size, Specific, Structure, System, Time, Transfer, Treating, Using*).

1.6. SUMMARY

In most countries continued growth of the MBR market is predicted, although growth rates vary markedly between countries and/or regions. The most rapid growth is in China, whereas the global recession has had a more adverse effect on countries such as Spain where any significant growth may be delayed until beyond 2011. For any region, a number of different factors — legislation, water scarcity, perceived return on investment, environmental considerations and public and political perception and engagement — influence the market to a greater or lesser extent, against a landscape of legacy and economic factors.

Although three main suppliers still dominate and the membrane configurations are still limited to the three that existed in the early 1990s (FS, HF and MT), the market is now supplemented with many other MBR membrane product and MBR technology suppliers. The commercial significance of the technology is manifested in:

- a global market exponential growth of 11.5—12.7%, or possibly higher;
- an implementation growth rate of over 50% p.a. in China specifically, and over 20% in some European countries;
- a steady increase in the number of commercial MBR membrane products by 4—5 per year since the turn of the Millennium;
- a decreased gestation time between the commercialization of an MBR membrane product and its large-scale implementation;
- marginally decreasing capital costs, in part due to the increased competition;
- marginally decreasing operating costs, primarily due to improvements made in membrane aeration efficiency; and
- increasing public acceptance of water reuse.

Against this, the more widespread adoption of the technology as the preferred process over competing technologies for municipal wastewater treatment is constrained by capital and operating costs, which remain higher than those of competing processes, and relative process complexity and reliability. However, it now appears that the MBR is the process of choice for water reuse applications, particularly where space is limited. Time will tell whether a combination of technical advances and the inexorable increase in the demand for ever higher treated water quality can sustain, or increase, the growth in the MBR market to the point where it becomes the automatic first choice for wastewater treatment.

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Chapter 2

Fundamentals

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2.3.5.3	Bruce Jefferson and Nacho Martin Garcia	Cranfield University
2.3.6.1–2.3.6.5	Pierre Le-Clech	University of New South Wales
2.3.6.6	Bruce Jefferson and Nacho Martin Garcia	Cranfield University
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2.3.10	Ana Santos	Cranfield University
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2.1. MEMBRANE TECHNOLOGY

2.1.1. Membranes and Membrane Separation Processes

A membrane as applied to water and wastewater treatment is simply a material that allows some physical or chemical components to pass more readily through it than others. It is thus perm-selective, since it is more permeable to those constituents passing through it (which then become the permeate) than those which are rejected by it (which form the retentate). The degree of selectivity depends on the membrane pore size. The coarsest membrane, associated with microfiltration (MF), can reject particulate matter. The most selective membrane, associated with reverse osmosis (RO), can reject singly charged (i.e. monovalent) ions, such as sodium (Na^{+}) and chloride (Cl^{-}). Given that the hydraulic diameter of these ions is less than 1 nm, it stands to reason that the pores in an RO membrane are very small. Indeed, they are only visible using the most powerful of microscopes.

The four key membrane separation processes in which water forms the permeate product are RO, nanofiltration (NF), ultrafiltration (UF) and MF (Fig. 2.1). Membranes themselves can thus be defined according to the type of separation duty to which they can be put, which then provides an indication of the pore size. The latter can be defined either in terms of the effective equivalent pore diameter, normally in µm, or the equivalent mass of the smallest molecule in daltons (Da) the membrane is capable of rejecting, where 1 Da represents the mass of a hydrogen atom. For UF membranes specifically the selectivity is thus defined by the molecular weight cut-off (MWCO) in daltons. For the key membrane processes identified, pressure is applied to force water through the membrane. However, there are additional membrane processes in which the membrane is not necessarily used to retain the contaminants and allow the water to pass through, but can instead be used either to:

(a) selectively extract constituents (extractive); or
(b) introduce a component in the molecular form (diffusive).

The range of membrane processes available is given in Table 2.1, along with an outline of the mechanism by which each process operates. Mature commercial

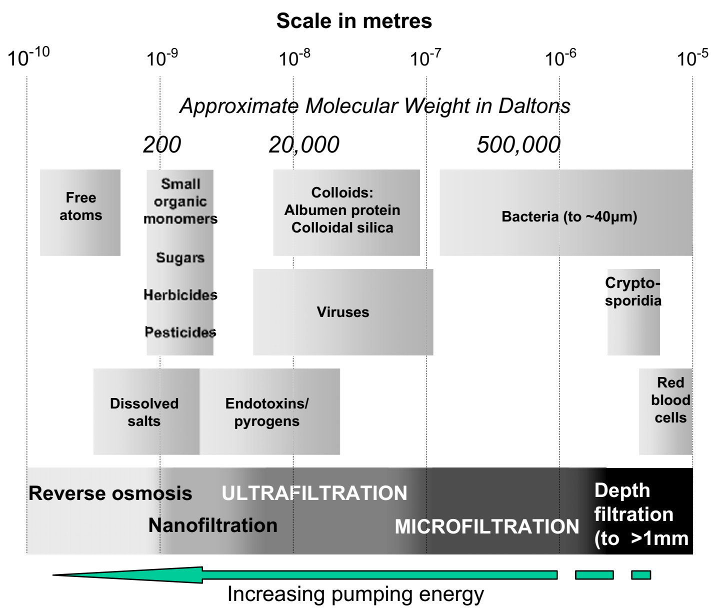

FIG. 2.1 Membrane separation processes overview *(Judd & Jefferson, 2003)*.

TABLE 2.1 Dense and Porous Membranes for Water Treatment

Pressure-driven/rejection	Extractive/diffusive
Reverse Osmosis (RO) Separation achieved by virtue of differing solubility and diffusion rates of water (solvent) and solutes in membrane.	Forward Osmosis (FO) Separation driven by difference in osmotic pressure across the membrane set up by employing an inert and recoverable 'draw' solution on the permeate side.
Nanofiltration (NF) Separation achieved through combination of charge rejection, solubility-diffusion and sieving through micropores (<2 nm*).	Electrodialysis (ED) Separation achieved by virtue of differing ionic size, charge and charge density of solute ions, using ion-exchange membranes.
Ultrafiltration (UF) Separation by sieving through mesopores (2−50 nm*)	Membrane Distillation (MD) Separation driven by employing a partial vacuum on the permeate side to provide a difference in partial pressure.
Microfiltration (MF) Separation of suspended solids from water by sieving through macropores (>50 nm*).	Membrane Extraction (ME) Constituent removed by virtue of a concentration gradient between retentate and permeate side of membrane.
	Gas Transfer (GT) Gas transferred under a partial pressure gradient into or out of water in molecular form.

**IUPAC (1985).*

membrane applications in water and wastewater treatment are limited to the pressure-driven processes and electrodialysis (ED), which can extract problem ions such as nitrate and those ions associated with hardness or salinity. Membrane technologies as applied to the municipal sector are predominantly pressure driven and, whilst the membrane perm-selectivity and separation mechanism may vary from one process to another, such processes all have the common elements of a purified permeate product and a concentrated retentate waste (Fig. 2.2).

The rejection of contaminants ultimately places a fundamental constraint on all membrane processes. The rejected constituents in the retentate tend to accumulate at the membrane surface, producing various phenomena which lead to a reduction in the flow of water through the membrane (i.e. the flux) at a given transmembrane pressure (TMP), or conversely an increase in the TMP for a given flux (reducing the permeability, which is the ratio of flux to TMP). These phenomena are collectively referred to as fouling. Given that membrane fouling represents the main limitation to membrane process operation, it is unsurprising that the majority of membrane material and process research and

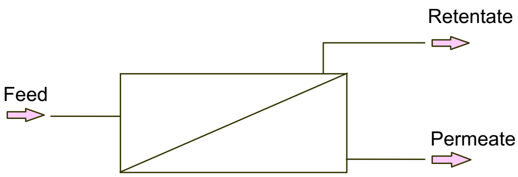

FIG. 2.2 Schematic of membrane.

development conducted is dedicated to its characterization and amelioration (Sections 1.5 and 2.3.6–2.3.9).

Fouling can take place through a number of physicochemical and biological mechanisms which all relate to increased deposition of solid material onto the membrane surface (also referred to as blinding) and within the membrane structure (pore restriction or pore plugging/occlusion). This is to be distinguished from clogging, which is the filling of the membrane channels with solids due to poor hydrodynamic performance, a more common phenomenon than fouling in membrane bioreactors (MBRs) (Section 3.6.2). The membrane resistance is fixed, unless its overall permeability is reduced by components in the feedwater permanently adsorbing onto or into the membrane. The resistance imparted by the interfacial region is, on the other hand, dependent on the total amount of fouling material residing in the region. This in turn depends upon the thickness of the interface, the feedwater composition (and specifically its foulant content) and the flux through the membrane. The feedwater matrix and the process operating conditions thus largely determine process performance.

2.1.2. Membrane Materials

There are mainly two different types of membrane material, these being polymeric and ceramic. Metallic membrane filters also exist, but these have very specific applications which do not relate to MBR technology. The membrane material, to be made useful, must then be formed (or configured) in such a way as to allow water to pass through it.

A number of different polymeric and ceramic materials are used to form membranes. Membranes generally comprise a thin surface layer which provides the required perm-selectivity on top of a more open, thicker porous support which provides mechanical stability. A classic membrane is thus anisotropic in structure, having symmetry only in the plane orthogonal to the membrane surface (Fig. 2.3). Membranes are usually fabricated both to have a high surface porosity, or per cent total surface pore cross-sectional area, and narrow pore size distribution to provide as high a throughput and selectivity as possible. The membrane must also be mechanically strong (i.e. have structural integrity). Lastly, the material will normally have some resistance to thermal and chemical attacks, that is, extremes of temperature, pH and/or oxidant concentrations that normally arise when the membrane is chemically cleaned

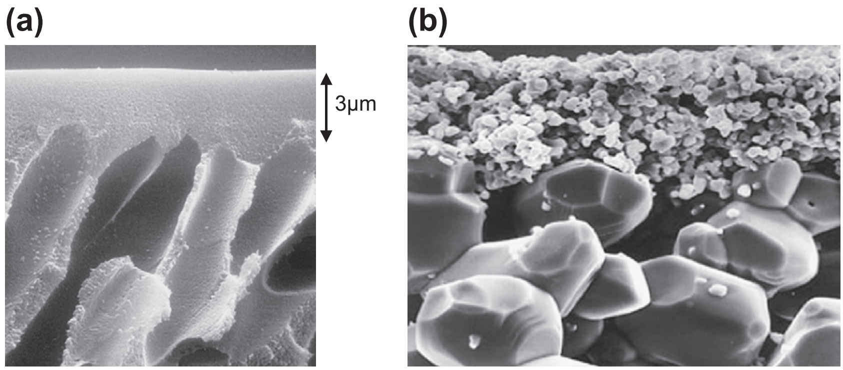

FIG. 2.3 Anisotropic UF membranes: (a) polymeric (thickness of 'skin' indicated) and (b) ceramic *(with kind permission from GE and Pall, respectively).*

(Sections 2.1.4.3 and 2.3.9.2), and should ideally offer some resistance to fouling.

Commercial UF/MF membranes span the range from fully hydrophilic polymers such as cellulose acetate (CA), which is not used at all in commercial MBRs, to fully hydrophobic polymers such as polypropylene (PP), polyethylene (PE) and fluoropolymers such as polytetrafluoroethylene (PTFE). Between the two extremes, there is the polysulfone (PS)/polyethersulfone (PES) family, polyacrylonitrile (PAN) and polyvinylidene difluoride (PVDF). Though these polymers are basically hydrophobic, membranes made from them can be modified to some extent through the use of additives, such as co-polymers, or by post-treatment. To be cost effective in large-scale applications, the membrane polymer must be a commodity product of low or medium price. Whilst the earliest commercial products in the water and wastewater field were based on PS, CA and PP, the MBR market is now supplied with products mainly based on PES, PVDF or on derivatives of PE.

Hollow fibre (HF) and capillary tube (CT) membranes (Section 2.1.3) can be made either by a phase inversion process or a stretching process (Chung, 2008). Stretching, also known as dry spinning, produces characteristic slit-like pores (Fig. 2.4a). UF membranes are produced by phase inversion, whilst MF can be made by either process. Phase inversion enables precise control of the pore size structure over a wide range of size. The phase inversion can be solvent based (the so-called wet spinning process) or temperature based (the thermal-induced phase separation (TIPS) process, Fig. 2.4b). In either case, the membrane is formed by inducing precipitation which occurs when the polymer solution containing the membrane polymer is destabilized. For wet spinning, the membrane polymer is dissolved in the solvent and subsequently precipitated by contact with a non-solvent. CA, PS/PES, PVDF and PAN all lend themselves to wet spinning. In the TIPS process, the precipitation of the

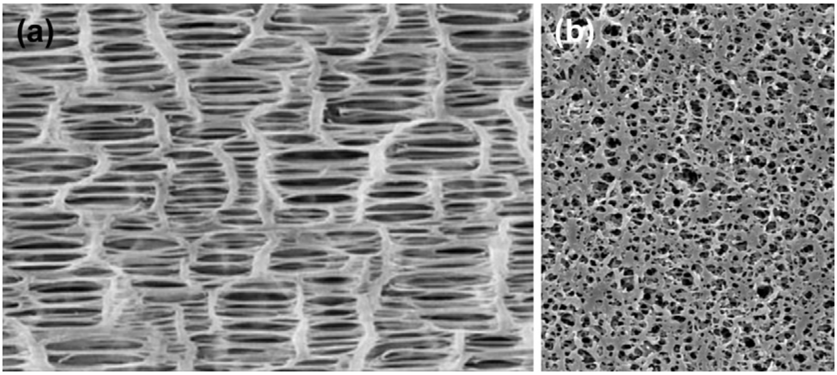

FIG. 2.4 Membrane surface: (a) dry-spun membrane and (b) TIPS *(with kind permission from Korea Membrane Separations and Asahi-Kasei, respectively).*

membrane is induced by a change in temperature, and this process can be used for PVDF membranes.

The solubility of PS/PES makes these polymers suitable for polymer blend membranes since other polymers can be co-dissolved, allowing the hydrophilic properties of the finished membrane to be modified and enhanced. Wet spinning allows the pore size and other membrane properties to be widely varied according to the spinning conditions. PS/PES can be made more hydrophilic with the appropriate blend of polymers (Boom, Wienk, van den Boomgaard, & Smolders, 1992), thereby gaining some of the advantages of the hydrophilicity of CA, whilst avoiding the primary disadvantages of this material, namely biodegradability and poor tolerance to caustic cleaning chemicals. PVDF membranes can be formed by wet spinning (Bottino, Roda, Capanelli, & Munari, 1991) and TIPS processes (Lloyd & Kinzer, 1990), but it is more difficult to modify since hydrophilic additives tend to lead to macrovoid formation and a weakening of the structure (Fortanova, Jansen, Cristiano, Curcio, & Drioli, 2006). Accordingly, most commercial PVDF membranes are relatively hydrophobic with surface characteristics close to the unmodified polymer. Dry spinning is used for hydrophobic polymers such as PP and PE. It tends to produce satisfactory MF membranes, though permeability may be low due to low pore density. Dry spinning produces slit-shaped pores (Fig. 2.4a), unlike the more unidimensional pores of wet spinning (Fig. 2.4b), generally yielding a slightly wider pore size distribution. PE can be made hydrophilic by post-treatment, but although derivatives of these membranes are used successfully in membrane bioreactors (MBRs), they are not used in conventional MF products. PP is used in MF products, but cannot be made hydrophilic.

In fabricating UF and MF membranes both the surface characteristics and the supporting sub-structure must be controlled, with the overall material

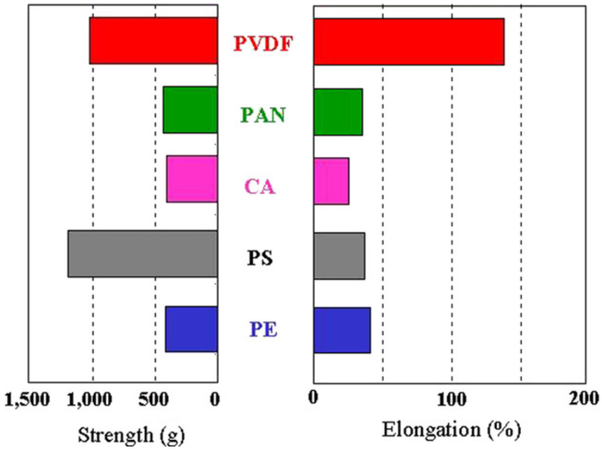

FIG. 2.5 Summary of polymer properties *(Pearce, 2007)*.

mechanical properties combining robustness with respect to bursting and collapsing strength (for HF and CT configurations) with a reasonable degree of flexibility. They must also have good chemical resistance, tolerating a wide pH range and high chlorine concentrations and so enabling rigorous and regularly applied chemical cleaning (Sections 2.1.4.3 and 2.3.9.2). In addition, thermal resistance should be such that moderate elevated temperatures can be used without detriment to the membrane properties or life. PS and PES are considered the most mechanically strong of the polymers (Fig. 2.5), with PVDF combining strength and flexibility. The left-hand side of Fig. 2.5 measures the tensile strength of various membrane polymers at their breaking point. On the right, the per cent elongation before the fibre breaks provides a measure of flexibility, and thus 100% would indicate that the membrane doubles in length before breaking. Flexibility is required when air scouring of HF membranes is employed, producing lateral movement of the fibres. Whilst general comparisons between materials can be made, a recent survey (Gijsbertsen-Abrahamse, Cornelissen, & Hofman, 2006) has shown that different polymer sources can affect the physical properties. Also, the membranes derived from polymers can vary widely in their properties according to the preparative conditions, membrane dimensions and other aspects.

The chemical resistance is equally important. Of the polymers discussed above, the PS/PES family has good chemical resistance, and can tolerate a pH range from 1.5 to 13, as well as moderate chlorine levels. PVDF tolerates acids down to a pH of 1, but is limited to operation below a pH of 11. However, it is very highly tolerant of chlorine, making it ideal for MBRs where cleaning with hypochlorite solutions is ubiquitous. PAN has similar pH tolerance to PVDF, combined with a moderate chlorine tolerance (probably similar to PS/PES). CA is much more limited in its chemical resistance, since its natural hydrophilicity makes it susceptible to hydrolysis in the presence of acids below pH 4 and caustic above pH 8. It tolerates chlorine but is biodegradable, which makes it sensitive to bacterial attack and is thus unsuitable for use in MBRs. The polyolefin family, PP and PE, has good tolerance to acid and caustic but low tolerance to chlorine. PTFE has the widest chemical tolerance of any membrane polymer with a pH range of 1−14 and the ability to withstand continuous exposure to a 20% hypochlorite solution. It is the only membrane polymer capable of resisting stronger oxidants such as ozone at any significant concentration.

All of the membrane materials employed for MBRs have a thermal resistance significantly greater than that suggested by the rating of the commercial modules in which they arise, which are typically warranteed up to a temperature of 40 °C. The module rating is normally dictated by the potting compound used, the pressure rating of the containment vessel or the thermal expansion limits of the potting/vessel interface.

A key surface property of the membrane is the hydrophilicity, or the extent of wetting by water. The degree of hydrophilicity is measured by the contact angle of the water droplet with the surface. If completely wetted, as is the case for CA, the contact angle is zero. For a strongly hydrophobic surface, such as PP, the contact angle is >90°. In water treatment, a hydrophilic membrane has some obvious advantages. First, the membrane is easily wetted, promoting high permeabilities relative to the pore size. Ready wetting also ensures that air flushing can be carried out without the risk of drying out; repeated contact of air with a hydrophobic surface will lead to a progressive loss of wetting. Second, the fouling constituents often present in surface water sources are organic in nature and readily attach to a hydrophobic surface (Section 2.3.6.2). A hydrophilic surface tends to resist attachment due to absorption by organics, and such a surface is referred to as a low fouling surface. However, many factors influence fouling (Fig. 2.28), such that predicting the most suitable membrane for a particular application is not a straightforward exercise.

In summary, a cursory review of the commercial UF/MF membranes used for water and wastewater treatment shows them to be produced from one of six polymers or polymer families, each with its own advantages and disadvantages. The different membrane characteristics inherent in each of these families have led to different products and operating methods to capitalize on the strengths of the various membranes. Important properties for a membrane are strength and flexibility, pore size and permeability, chemical resistance and hydrophilicity. Whilst in principle a membrane can be formed from a very wide range of polymeric materials, to be cost effective for large-scale applications a membrane polymer needs to be made from a commodity product. The PS/PES family and PVDF are now emerging as the dominant polymers of choice for the water industry, but with PP, PE, PAN and, more recently, PTFE also being available. Both of the two main polymer families have properties conducive to their application to MBRs: PS/PES copolymers can provide a hydrophilic membrane, a narrow pore size distribution of UF rating and excellent all round chemical tolerance. PVDF provides excellent strength and flexibility, with very high chlorine tolerance, and can be formed into a UF/fine MF membrane. Further discussion of recent research relating to membrane properties is provided in Section 2.3.5.

2.1.3. Membrane Configurations

The configuration of the membrane, that is, its geometry and the way it is mounted and oriented in relation to the flow of water, is crucial in determining the overall process performance. Other practical considerations concern the way in which the membrane elements, that is the individual discrete membrane units themselves, are housed in containers (or 'shells') to produce modules, the complete vessels through which the water flows.

Ideally, the membrane should be configured so as to have:

(a) a high membrane area to module bulk volume ratio (or packing density),
(b) a high degree of turbulence for mass transfer promotion on the feed side,
(c) a low energy expenditure per unit product water volume,
(d) a low cost per unit membrane area,
(e) a design that facilitates cleaning and
(f) a design that permits modularization.

All membrane module designs, by definition, permit modularization (f), and this presents one of the attractive features of membrane processes. This also means that membrane processes provide limited economy of scale with respect to membrane costs, since these are directly proportional to the membrane area which relates directly to the flow. However, some of the remaining listed characteristics are mutually exclusive. For example, promoting turbulence (b) results in an increase in the energy expenditure (c) and is adversely affected by high packing densities (a). On the other hand, low packing densities can also be associated with higher unit membrane costs (d). Finally, it is not possible to produce a high-membrane packing density (a) without narrowing the retentate flow channels, which will then compromise turbulence promotion (b) and ease of cleaning (e).

There are six principal configurations currently employed in membrane processes, which all have various practical benefits and limitations (Table 2.2). The configurations are based on either a planar or cylindrical geometry and comprise:

1. Plate-and-frame/flat sheet (FS)
2. Hollow fibre (HF)
3. (Multi)tubular (MT)
4. Capillary tube (CT)
5. Pleated filter cartridge (FC)
6. Spiral-wound (SW)

TABLE 2.2 Membrane Configurations

Configuration	Turbulence Promotion	Backflushable?	Application
FC	Very poor	No*	**DEMF,** **low TSS waters**
FS	*Fair*	*No***	***ED***, *UF, RO*
SW	Poor	No	**RO/NF**, UF
MT	*Very good*	*No*	***CFMF/UF,*** ***high TSS waters***, *NF*
CT	Fair	Yes	**UF**
HF	*Very poor–fair*#	*Yes*	***MF/UF***, *RO*

Most important applications *MBR configurations.*
DE = Dead-end, CF = crossflow, TSS = total suspended solids.
**Some cartridge filters are backflushable, but not pleated filters.*
***Some newer FS panels are backflushable (Section 4).*
#Degree of turbulence promotion dependent on packing density and air–liquid contacting.

Of the above configurations, only the first three are suited to MBR technologies (Figs 2.6 and 2.7), principally for the reasons outlined previously: the modules must permit turbulence promotion and regular effective cleaning. Turbulence promotion can arise through passing either the feedwater or an air/water mixture along the surface of the membrane to

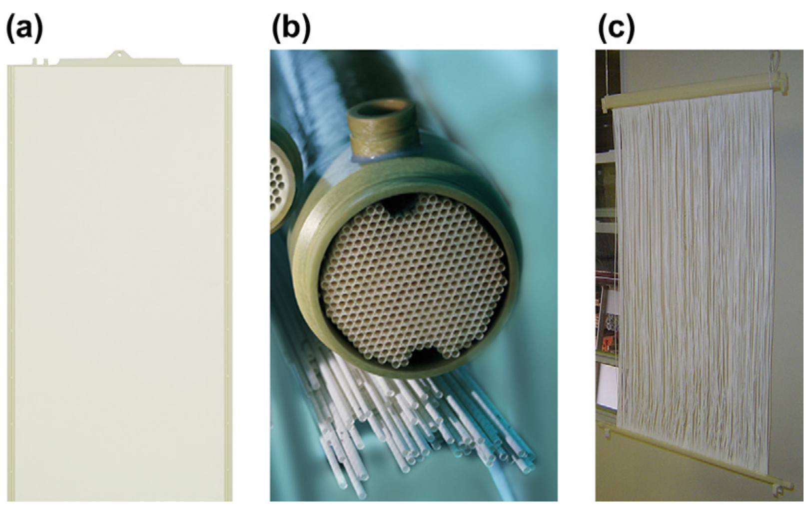

FIG. 2.6 Membrane configurations: (a) FS, (b) MT and (c) HF modules *(with kind permission from Toray, Berghof and Superstring, respectively).*

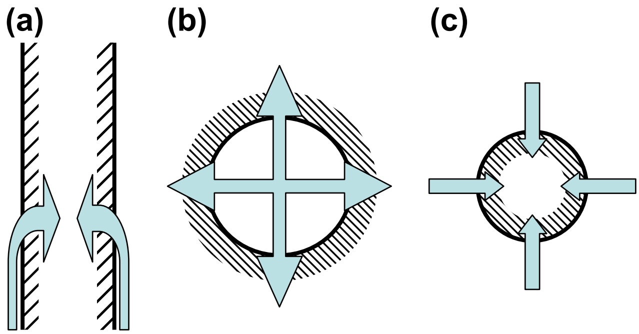

FIG. 2.7 Schematics showing flow through membrane configured as: (a) FS, (b) CT or MT and (c) HF.

aid the passage of permeate through it. This crossflow operation (Section 2.1.4.2) is widely used in many membrane technologies, and its efficacy increases with increasing membrane interstitial distance (i.e. the membrane separation).

Because the MT module operates with flow passing from inside to outside the tube (‘lumen-side’ to ‘shell-side’), whereas the HF operates outside-to-in, the interstitial distance is defined by (Fig. 2.7):

- the tube diameter for an MT,
- the distance between the filaments for an HF and
- the channel width for an FS.

The membrane packing density of the HF and FS modules is thus crucial, since too high a packing density will reduce the interstitial gap to the point where there is a danger of clogging (Section 3.6.2). CT modules, which are, to all intents and purposes, HF modules with reversed flow (i.e. lumen-side to shell-side, Fig. 2.7), are too narrow in diameter to be used for MBR duties as they would be at high risk of clogging.

Physical cleaning may be achieved by reversing the flow (i.e. backflushing), at a rate 1−3 times higher than the forward flow, back through the membrane to remove some of the fouling layer on the retentate side. For this to be feasible, the membrane must have sufficient inherent integrity to withstand the hydraulic stress imparted. In other words, the membrane must be strong enough not to break or buckle when the flow is reversed. This generally limits backflushing of polymeric membranes to those configured as capillary tubes or HFs, since at low filament diameters the membranes have a high enough wall thickness:filament diameter ratio to have the inherent strength to withstand stresses imposed by flow reversal. However, ceramic membranes are all backflushable, since they are rigid, and backflushable FS modules are now commercially available, with more being developed.

2.1.4. Membrane Process Operation

2.1.4.1. Flux, Pressure, Resistance and Permeability

The key elements of any membrane process relate to the influence of the following parameters on the overall permeate flux:

(a) the membrane resistance,
(b) the operational driving force per unit membrane area,
(c) the hydrodynamic conditions at the membrane:liquid interface and
(d) the fouling and subsequent cleaning of the membrane surface.

The flux (normally denoted J) is the quantity of material passing through a unit area of membrane per unit time. This means that it takes SI units of $m^3/(m^2\,s)$, or simply m/s, and is occasionally referred to as the permeate or filtration velocity. Other non-SI units used are litres per m^2 per hour (or LMH) and m/day, which tend to give more accessible numbers. The most usual units for MBRs are m/day, with Imperial units of gallons per square foot per day (GFD) still used in the USA. MBRs generally operate at fluxes between 10 and 150 LMH; the flux relates directly to the driving force (i.e. the transmembrane pressure, or TMP, for conventional MBRs) and the total hydraulic resistance offered by the membrane and the interfacial region adjacent to it.

Although for conventional biomass separation of MBRs the driving force for the process is the TMP, for extractive or diffusive MBRs (Sections 2.3.2–2.3.4) it is respectively the concentration or partial pressure gradient. Whereas with conventional pressure-driven MBRs the permeate is the purified product, for extractive MBRs the contaminants are removed from the water across the membrane under the influence of a concentration gradient and are subsequently biologically treated, the retentate forming the purified product. For diffusive bioreactors neither water nor contaminants permeate the membrane; in this case the membrane is used to transport a gas into the bioreactor.

Resistance R (per m) is the ratio of the pressure difference ΔP to the flux and viscosity η, and hence given by $\Delta P/(\eta J)$, and is inversely related to the permeability K which normally takes the most convenient units of LMH/bar. In the USA permeability is often termed 'specific flux', and takes Imperial units of GFD/psi. ΔP can refer to either the TMP (ΔP_{m}) or individual components which contribute to the pressure drop and so the resistance offered. The resistance R may then include a number of components, namely:

(a) the membrane resistance,
(b) the resistance of the fouling layer (adsorbed onto the membrane surface) and
(c) the resistance offered by the membrane:solution interfacial region.

The membrane resistance is governed by the membrane material itself, and mainly the pore size, the surface porosity (percentage of the surface area covered by the pores) and the membrane thickness. The fouling layer resistance is associated with the filtration mechanism, which is then dependent on the membrane and filtered solids characteristics. The membrane:solution interfacial region resistance is associated with concentration polarization (CP, Section 2.1.4.4) which, for the more perm-selective processes such as RO, produces a solution osmotic pressure at the membrane surface which is higher than that in the bulk solution as well as exacerbating fouling. The resistance offered by foulants is often further delineated into generic types according to their characteristics, behaviour and origin (Sections 2.3.6.2—2.3.6.6). However, in general, the membrane resistance only dominates when fouling is either absent (i.e. the feedwater is almost free of fouling materials) or is suppressed by operating under specific conditions (Section 2.3.9).

2.1.4.2. Dead-end and Crossflow Operation

Conventional pressure-driven membrane processes with liquid permeation can operate in one of two modes. If there is no retentate stream then operation is termed 'dead-end' or 'full-flow'; if retentate continuously flows from the module outlet then the operation is termed crossflow (Fig. 2.8). Crossflow implies that, for a single passage of feedwater across the membrane, only a fraction is converted to permeate product. This parameter — the ratio of permeate to feed flow — is termed the 'conversion' or 'recovery'. The recovery is reduced further if product permeate is used for maintaining process operation, usually for membrane cleaning (Section 2.1.4.3).

Filtration always leads to an increase in the resistance to flow. In the case of a dead-end filtration process, the resistance increases according to the thickness of the cake formed on the membrane, which would be expected to be roughly proportional to the total volume of filtrate passed. Rapid permeability decay then results, at a rate proportional to the solids concentration and flux, demanding periodic cleaning (Fig. 2.9). For crossflow processes, this deposition continues until the adhesive forces binding the cake to the membrane are balanced by the scouring forces of the fluid (either liquid or a combination of air and liquid) passing over the membrane. All other things being equal,

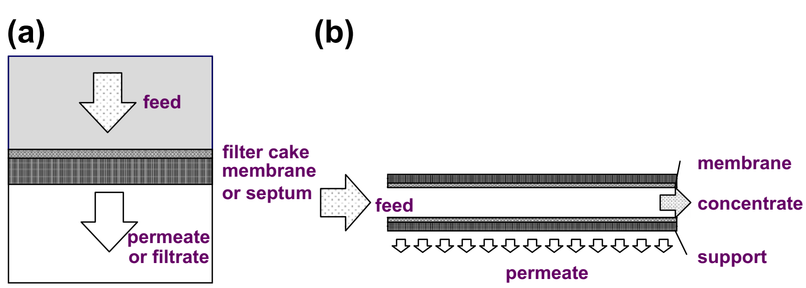

FIG. 2.8 (a) Dead-end and (b) crossflow filtration.

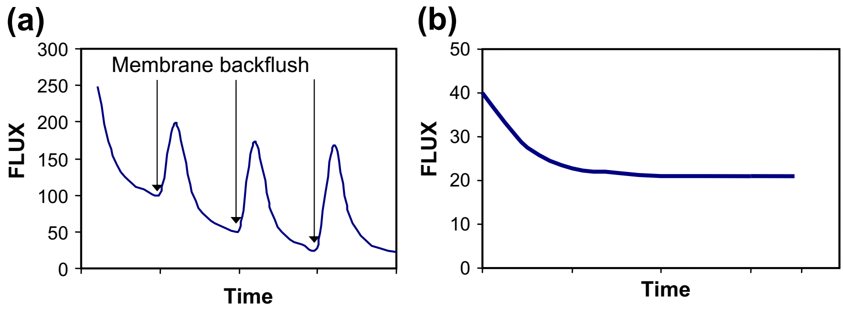

FIG. 2.9 Flux transients for (a) dead-end and (b) crossflow filtration for constant pressure operation.

a crossflow filtration process would be expected to attain steady-state conditions determined by the degree of CP (Section 2.1.4.4). In practice, only pseudo-steady-state (or stabilized) conditions are attained due to the unavoidable deposition or adsorption of fouling material.

Filtration proceeds according to a number of widely recognized mechanisms, which have their origins in early filtration studies (Grace, 1956), comprising (Fig. 2.10):

- complete blocking,
- standard blocking,
- intermediate blocking and
- cake filtration.

All models imply a dependence of flux decline on the ratio of the particle size to the pore diameter. The standard blocking and cake filtration models appear most suited to predicting initial flux decline during colloid filtration (Visvanathan & Ben Aim, 1989) or protein filtration (Bowen, Calvo, & Hernandez, 1995). All of the models rely on empirically derived information and some have been refined to incorporate other key determinants. On the other hand, a number of empirical and largely heuristic expressions have been proposed for particular matrices and/or applications. Classical dead-end

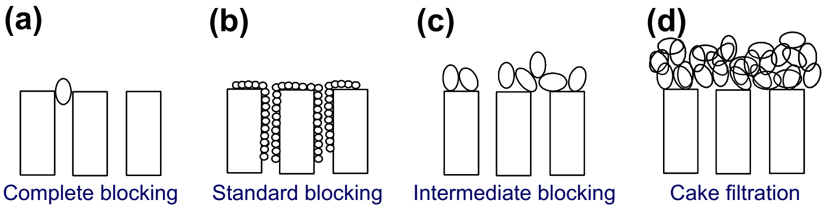

FIG. 2.10 Fouling mechanisms.

filtration models can be adapted for crossflow operation if the proportion of undeposited solute material can be calculated.

2.1.4.3. Physical and Chemical Cleaning

Since the flux and driving force are interrelated, either one can be fixed for design purposes. For conventional pressure-driven water filtration, it is usual to fix the value of the flux and then determine the appropriate value for the TMP. The main impact of the operating flux is on the period between cleaning, which may be by physical or chemical means (Fig. 2.11), and usually both. In MBRs, physical cleaning is normally achieved either by backflushing, i.e. reversing the flow, or relaxation, which is simply ceasing permeation whilst continuing to scour the membrane with air bubbles. These two techniques may be used in combination, and backflushing may be enhanced by combination with air. Chemical cleaning is carried out with mineral or organic acids, caustic soda or, more usually in MBRs, sodium hypochlorite, and can be performed either in situ ('cleaning in place' or CIP) or ex situ. Alternatively, a lower concentration of chemical cleaning agent can be added to the backflush water to produce a 'chemically enhanced backflush' (CEB), usually performed only periodically.

Physical cleaning is less onerous than chemical cleaning on a number of bases. It is generally a more rapid process than chemical cleaning, lasting no more than two minutes. It demands no chemicals and produces no chemical waste, and also is less likely to incur membrane degradation. On the other hand, it is also less effective than chemical cleaning. Physical cleaning removes gross solids attached to the membrane surface, generally termed 'reversible' or 'temporary' fouling, whereas chemical cleaning removes more tenacious material often termed 'irreversible' or 'permanent' fouling, which is obviously something of a misnomer. Since the original virgin membrane permeability is never recovered once a membrane is fouled through normal operation, there remains a residual resistance which can be defined as 'irrecoverable fouling'.

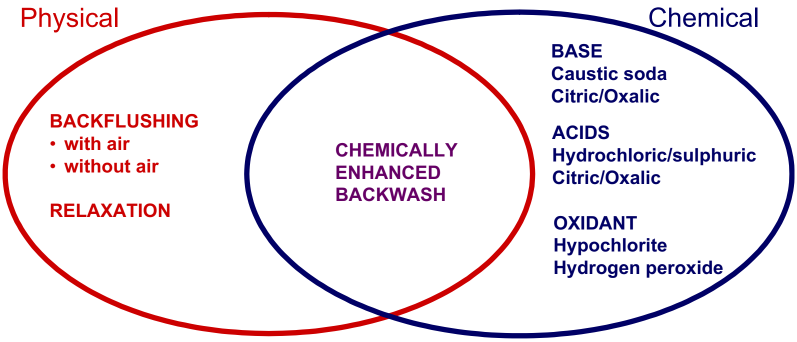

FIG. 2.11 Membrane cleaning methods.

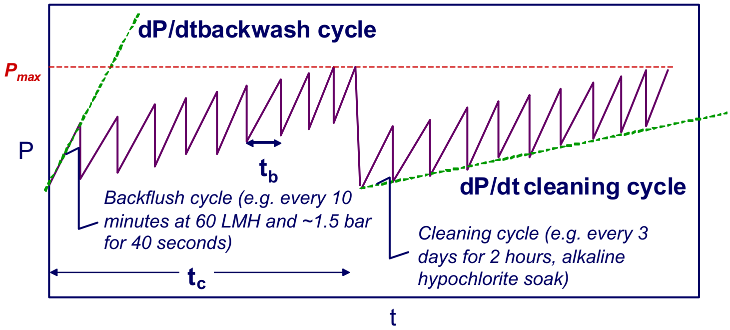

FIG. 2.12 Pressure transient for constant flux operation of a dead-end filter.

It is this fouling which builds up over a number of years and may ultimately determine membrane life.

Since flux, amongst other things, influences the permeability decline rate dK/dt (or pressure increase dP/dt), it also determines the period between physical cleaning (backflushing or relaxation), that is, the physical cleaning cycle time. If backflushing is used, this period can be denoted t_p and, assuming no changes to other operating conditions, increasing the flux decreases t_p. Since backflushing does not, in practice, return the permeability to the original condition, only a finite number of backflush cycles can be performed before a threshold pressure is reached (P_{max}) beyond which operation cannot be sustained. At this point chemical cleaning must be conducted to return the pressure to close to the original baseline value (Fig. 2.12). As with physical cleaning, chemical cleaning never recovers the original membrane permeability but is normally considerably more effective than physical cleaning. For some crossflow modules backflushing is not normally an option due to the nature of the membrane module (Table 2.2), and membrane permeability is thus maintained by a combination of relaxation and chemical cleaning.

2.1.4.4. Concentration Polarization

For membrane filtration processes, the overall resistance at the membrane:solution interface is increased by a number of factors which each place a constraint on the design and operation of membrane process plant:

(a) the concentration of rejected solute near the membrane surface;
(b) the precipitation of sparingly soluble macromolecular polymeric and inorganic (gel layer formation and scaling, respectively) at the membrane surface; and
(c) the accumulation of retained solids on the membrane (cake layer formation).

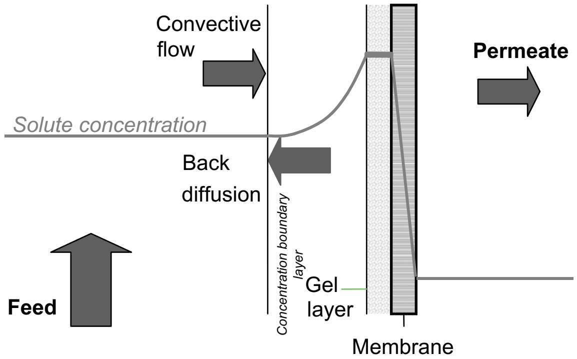

FIG. 2.13 Concentration polarization.

All of the above contribute to membrane fouling, and (a) and (b) are promoted by concentration polarization. CP describes the tendency of the solute to accumulate at the membrane:solution interface within a concentration boundary layer, or liquid film, during crossflow operation (Fig. 2.13). This layer contains near-stagnant liquid, since at the membrane surface itself the liquid velocity must be zero. This implies that the only mode of transport within this layer is diffusion, which can be two orders of magnitude slower than convective transport in the bulk liquid region. However, it has been demonstrated (Romero & Davis, 1991) that transport away from the membrane surface is much greater than that governed by Brownian diffusion and is actually determined by the amount of shear imparted at the boundary layer; such transport is referred to as 'shear-induced diffusion'.

Rejected materials nonetheless build up in the region adjacent to membrane, increasing their concentration over the bulk value, at a rate which increases roughly exponentially with increasing flux. The thickness of the boundary layer, on the other hand, is determined entirely by the system hydrodynamics, decreases and when turbulence is promoted. For crossflow processes, the greater the flux, the greater the build-up of solute at the interface; the greater the solute build-up, the steeper the concentration gradient, and so the faster the diffusion. Under normal steady-state operating conditions, there is a balance between those forces transporting the water and constituents within the boundary layer towards, through and away from the membrane. This balance is determined by CP.

2.1.4.5. Fouling Control

In MBRs, as with many other membrane filtration processes, it is the balance between the flux, physical and chemical cleaning protocol and, when relevant, the control of CP which ultimately determines the extent to which fouling is successfully suppressed. CP-related fouling can be reduced by two methods: (i) promoting turbulence (which then decreases the thickness of the boundary layer) and (ii) reducing the flux. For sidestream MBRs (sMBRs, Fig. 1.1a), turbulence can be promoted simply by increasing the crossflow velocity (CFV), whereas for an immersed system (iMBR, Fig. 1.1b) this can only reasonably be achieved by increasing the membrane aeration. Whereas pumped flow of liquid along a tubular or parallel plate channel, as with sidestream systems, allows estimation of the degree of turbulence through calculation of the Reynolds number (density × velocity × tube diameter/viscosity), determination of turbulence for an immersed aerated membrane (Section 2.3.7.1) is more challenging.

2.1.4.6. Critical Flux

The critical flux concept was originally presented by Field, Wu, Howell, and Gupta (1995). These authors stated that: 'The critical flux hypothesis for MF/UF processes is that on start-up there exists a flux below which a decline of flux with time does not occur; above it, fouling is observed'. Two distinct forms of the concept have been defined. In the strong form, the flux obtained during sub-critical flux is equated to the clean water flux measured under the same conditions. However, clean water fluxes are rarely attained for most real feedwaters due to irreversible adsorption of some solutes. In the alternative weak form, the sub-critical flux is the flux rapidly established and maintained during start-up of filtration, but does not necessarily equate to the clean water flux. Alternatively, stable filtration operation, that is, constant permeability for an extended time period, has been defined as sub-critical operation even when preceded by an initial decline in flux (Howell, 1995). Such conditions would be expected to lead to lower critical flux values than those obtained for constant permeability operation, however, since an initial permeability decline implies foulant deposition.

A number of slightly different definitions of sub-critical flux operation have been proposed, largely depending on the method employed. The most microscopically precise definition equates the critical flux to that flux below which no deposition of colloidal matter takes place. Kwon and Vigneswaran (1998) equated critical flux to the lift velocity as defined by the lateral migration theory of Green and Belfort (1980). This rigorous definition is difficult to apply because of the relative complexity of the determination of the lift velocity, particularly for heterogeneous matrices. On the other hand, experimental determination of critical flux by direct observation of material deposition onto the membrane has been conducted using model homodispersed suspensions of polystyrene latex particles (Kwon & Vigneswaran, 1998), and some authors have also used mass balance determinations (Kwon, Vigneswaran, Fane, & Ben Aim, 2000).

Given the limitations of applying particle hydrodynamics to the identification of the critical flux in real systems, recourse generally has to be made to experimental determination. By plotting flux against the TMP it is possible to

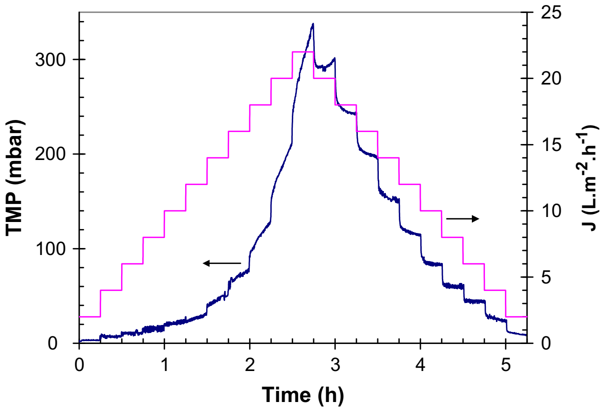

FIG. 2.14 Flux and pressure relationship for 15-min flux steps, MBR biomass *(Le-Clech et al., 2003b).*

observe the transition between the linearly pressure-dependent flux and the onset of fouling, where deviation from linearity commences. The flux at this transition has been termed 'secondary critical flux' (Bouhabila, Ben Aim, & Buisson, 1998) and the concept of 'sustainable flux' has since been introduced, defined as the flux for which the TMP increases gradually at an acceptable rate, such that chemical cleaning is not necessary (Ng, Tan, Ong, Toh, & Loo, 2006). There remains some debate as to the fundamental definition of critical and/or sustainable flux for MBRs.

Whilst potentially useful in providing a guide value for the appropriate operating flux, the absolute value of the critical flux obtained is dependent on the exact method employed for its determination and, specifically, the rate at which the flux is varied with time. A common practice is to increase incrementally the flux for a fixed duration for each increment, giving a stable TMP at low flux but an ever-increasing rate of TMP increase at higher fluxes (Fig. 2.14). This flux-step method defines the highest flux for which TMP remains stable as the critical flux. This method is preferred over the corresponding TMP-step method since the former provides a better control of the flow of material deposition on the membrane surface, as the convective flow of solute towards the membrane is constant during the run (Defrance & Jaffrin, 1999). No single protocol has been agreed for critical flux measurement, making comparison of reported data difficult. A practical method based on a threshold permeability change was proposed by Le-Clech, Jefferson, Chang, & Judd (2003b), and further refinements in irreversible fouling measurements for determining the critical flux have been made in more recent years (Huyskens, Brauns, Van Hoof, and De Wever, 2008).

It is also apparent from bench- and pilot-scale studies that irreversible fouling of MBR membranes can take place at operation well below the critical flux. Pertinent studies have been summarized by Pollice, Brookes, Jefferson, and Judd (2005) and Meng et al. (2009). Sub-critical flux fouling appears to be

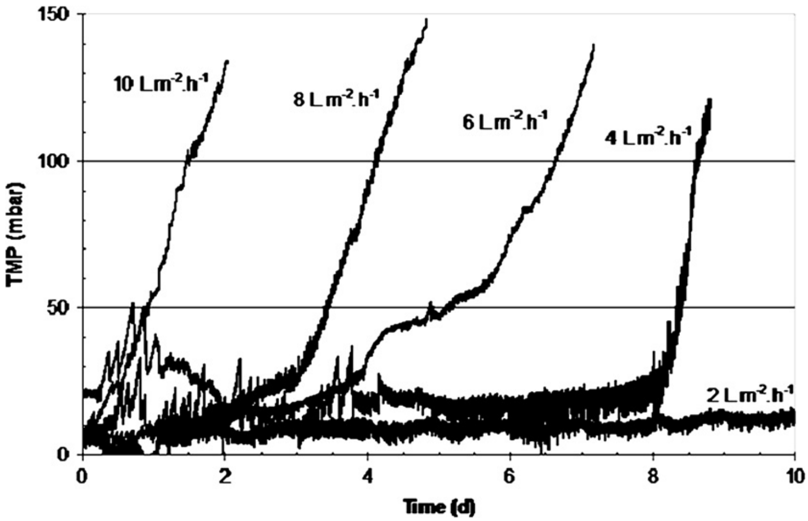

FIG. 2.15 TMP transients for sub-critical flux operation *(Brookes et al., 2006)*.

characterized by a sudden discontinuity of the TMP (Fig. 2.15, 4 LMH line) at very low flux operation after some extended time period (Brookes et al., 2004; Ognier, Wisnieswski, & Grasmick, 2001; Wen, Bu, & Huang, 2004; Meng et al., 2009) and a steady neo-exponential increase at fluxes closer to the notional critical flux (Fig. 2.15, 10 LMH line). Sub-critical fouling is discussed further along with MBR membrane fouling mechanisms in Section 2.3.8.

2.2. BIOTREATMENT

2.2.1. Biotreatment Rationale

Biological treatment (or biotreatment) processes are those which remove dissolved and suspended organic chemical constituents through biodegradation, as well as suspended matter through physical separation. Biotreatment demands that the appropriate reactor conditions prevail so as to maintain sufficient levels of viable (i.e. living) micro-organisms (or, collectively, biomass) to achieve removal of organics. The latter are normally measured as biochemical or chemical oxygen demand (BOD or COD, respectively); these are indirect measurements of organic matter levels since both refer to the amount of oxygen utilized for oxidation of the organics. The micro-organisms that grow on the organic substrate on which they feed generate cellular material from this organic matter, and can be aerobic (oxygen-dependent) or anaerobic (oxygen-independent). They are subsequently separated from the water to leave a clarified effluent that has a reduced level of organic matter.

The most attractive feature of biological processes is the very high chemical conversion efficiency achievable. Unlike chemical oxidation processes, aerobic processes can quantitatively mineralize large organic molecules, that is, convert them to the end mineral constituents of CO_2, H_2O and inorganic nitrogen products, at ambient temperatures without significant onerous chemical by-product formation, though a solid waste (sludge) is produced. In doing so a variety of materials are released from the biomass in the reactor which are collectively referred to as extracellular polymeric substances (EPS) and which contain a number of components which contribute to membrane fouling in an MBR (Section 2.3.6). The relative and overall concentrations of the various components are determined both by the feed characteristics and operational facets of the system, such as microbial diversity. Anaerobic processes generate methane as an end product, a possible thermal energy source, and similarly generate EPS. Biotreatment processes are generally robust to variable organic loads, create little odour (if aerobic) and generate a waste product (sludge) which is readily processed. On the other hand, they are slower than chemical processes, susceptible to toxic shock and consume energy associated with aeration in aerobic systems and mixing in all biotreatment systems.

2.2.2. Biotreatment Processes

Processes based on biodegradation can be classified according to the process configuration, feeding regime and oxidation state (Table 2.3). Process configuration defines the way in which the water is contacted with the biomass, which can form a layer on some supporting media to form a fixed biofilm or be suspended in the reactor, or sometimes a combination of these. Suspended growth systems tend to provide higher mass transfer (although submerged aerated media processes also provide reasonably high mass transfer), but the biomass subsequently needs to be separated from the water. Both configurations generate excess biomass which needs to be disposed of. Feeding regime defines the way in which the feedwater is introduced, which can be either continuous or batch-wise. Feeding in batches allows the same vessel to be used both for biodegradation and separation, thus saving on space. This is the case for the sequencing batch reactor (SBR). Finally, the reduction–oxidation (redox) conditions are defined by the presence of either dissolved oxygen (DO) (aerobic conditions) or some other compound capable of providing oxygen for bioactivity (anoxic conditions) or the complete absence of any oxygen (anaerobic conditions). The different redox conditions favour different microbial communities and are used to affect different types of treatment.

Aerobic treatment is used to remove organic compounds (BOD or COD) and to oxidize ammonia to nitrate. Aerobic tanks may be combined with anoxic and anaerobic tanks to provide biological nutrient removal (BNR). BNR, the removal of nitrogen and phosphorus, is discussed further in Section 2.2.4.4, and the various facets of biological processes in general are described in detail in various reference books (Tchobanoglous et al. 2003; Henze, van Loosdrecht, Ekama, & Brdjanovic, 2008; Grady, Daigger, & Love, 2010). However, almost all biological processes are configured according to the sub-categories listed in Table 2.3, and their function and performance depend on which specific sub-categories apply. Moreover, unit biotreatment processes can be combined so as to achieve multiple functions. So, for example, within an individual bioreactor, both aerobic and anoxic processes can be designed to occur within different zones.

TABLE 2.3 Examples of Biological Processes and Their Characteristics

	Process Configuration		Feeding Regime		Redox Conditions		
	Fixed Film	Suspended Growth	Continuous	Fed-Batch	Aerobic	Anoxic	Anaerobic
AD		X	(X)	(X)			X
AF	X		X				X
(C) ASP		X	X		X	(X)	(X)
BAF	X		X		X		
IFAS		X					
MBBR	X		X		X		
RBC	X		X		X		
SAF	X		X		X		
SBR		X		X	X	(X)	
TF	X				X		
UASB		X	X				X
MBR		**X**	**X**		**X**	**(X)**	**X**

Key:
AD Anaerobic digestion
AF Anaerobic filter
(C)ASP (Conventinal) activated sludge process
BAF Biological aerated filter
IFAS Integrated fixed film in activated sludge
MBBR Moving bed bioreactor
RBC Rotating biological contactor
SAF Submerged aerated filter
SBR Sequencing batch reactor
TF Tricking filter
UASB Upflow anaerobic sludge blanket

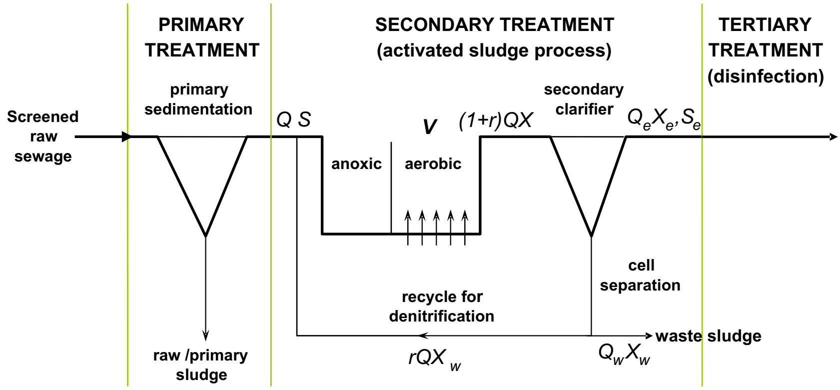

FIG. 2.16 Classic sewage treatment process, with mass flows for the ASP indicated.

The classic sewage treatment process (Fig. 2.16) is the combination of screening of gross solids, and then sedimentation of settlable solids followed by a biological process. Various configurations that include a preliminary anaerobic zone to remove phosphorus biologically are also available. Aerobic processes may be configured either as suspended growth, the (conventional) activated sludge process or (C)ASP, or fixed film, predominantly as a trickling filter or TF. Total removal of organic nitrogen (ON) from the feedwater can be achieved by recycling the nitrate-rich sludge from the CASP to some point upstream of the aerobic process where anoxic conditions then prevail; nitrification and denitrification are thus carried out sequentially. Aerobic MBRs can be configured similarly since, in essence, the biological function remains unaltered by the membrane.

In all biotreatment processes, the treated water must be separated from the biomass. Fixed film process effluent is notionally low in biological material since the latter forms a biofilm on the growth media, although biofilms can slough off into the product water, whereas in the CAS process separation is normally by sedimentation. This means that CASPs rely on the solids (which are flocculated particles and referred to as flocs) growing to a size where they can be settled out, which means that they must be retained in the bioreactor for an appropriate length of time. The solid retention time (SRT) is thus coupled with the hydraulic retention time (HRT), the retention time being the time taken for the solids and water respectively to pass through the reactor. For commercial MBR technologies, separation is by membrane filtration, eliminating the requirement for substantial floc growth and the associated long HRTs. The key advantage offered by the MBR process, with specific respect to biotreatment, is thus the uncoupling of the HRT and SRT; the implications of this regarding design are discussed in Section 2.2.4.5. However, MBRs can also be configured as fixed film processes, using the membrane to support a biofilm. These types of MBR are discussed in Sections 2.3.2 and 2.3.3.

2.2.3. Microbiology

Biological treatment relies on conversion of organic and inorganic matter into innocuous products by micro-organisms and, as such, the biological community must be healthy and sustainable. Figure 2.17 illustrates the food chain in a biotreatment environment. Higher forms of micro-organisms such as protozoa and rotifers play crucial roles in consuming suspended organic matter and controlling sludge concentration by scavenging bacteria. Larger biological species such as nematode worms and insect larvae may contribute to the consumption of particulate organic matter, especially in TF systems.

There is some evidence to suggest that higher organisms, protozoa, filamentous organisms, nematodes and ciliates, are present at lower concentrations in MBRs than in conventional activated sludge systems (Cicek, Franco, Suidan, Urbain, & Manem, 1999; Witzig et al., 2002; Wei, van Houten, Borger, Eikelboom, & Fan, 2003). However, higher concentrations of protozoa, particularly flagellates and free ciliates, have been reported for MBRs compared with a CASP operating at the same SRT (Ghyoot & Verstraete, 2000). These experiments were performed on a system with long HRT (20−74 h); hence the shorter HRT associated with MBRs may be responsible for the absence of protozoa in other studies. Predatory organisms have a negative effect on nitrification (Lee & Welander, 1994) and overgrowth of protozoa has been shown to create a complete breakdown of nitrification (Bouchez et al., 1998). This predation (or grazing) in activated sludge is accounted for in the death coefficient (k_e), and recent research suggests that this effect has a greater impact on sludge concentration than previously thought in an activated sludge system (Van Loosdrecht & Henze, 1999). In contrast, the

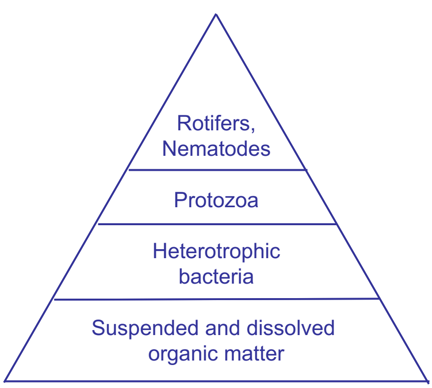

FIG. 2.17 Ecology of activated sludge systems.

sludge concentration in an MBR is limited by the energy provided and cell decay. Higher filamentous organisms, such as *Nocardia*, have been shown to develop in full-scale MBRs and produce significant foaming problems (Smith, 2006; Sections 2.3.6.3 and 3.6.5).

Conditions can be created in an MBR whereby the sludge is accumulated to the point where the biomass concentration is such that all of the energy available is used for cell maintenance. The high sludge concentration compared to the food available creates an environment where bacteria face starvation conditions so the bacteria are not in a physiological state for cell growth (Müller, Stouthamer, Vanverseveld, & Eikelboom, 1995). Oxygen uptake rates in an MBR system compared with a conventional activated sludge system are lower, indicating that the MBR is carbon rather than oxygen limited (Witzig et al., 2002). Even if the cells in an MBR system are not growing, new bacteria are constantly being introduced with the influent wastewater; if fewer grazing organisms exist, then cell decay is necessary to keep the biomass concentration constant.

The microbial community in any biological system comprises a large number of different bacterial species. Micro-organisms can be classified according to the redox conditions in which they prevail (Table 2.4), and hence the process type, and their energy requirements. Heterotrophs use organic carbon as an energy source and for synthesis of more cellular material, and are responsible for BOD removal and denitrification. Autotrophs use inorganic reactions to derive energy, for example, oxidation of iron(II) to iron(III) or hydrogen to water, and obtain assimilable material from an inorganic source (such as carbon from carbon dioxide) to carry out such processes as nitrification, sulphate reduction and anaerobic methane formation. Autotrophs are generally less efficient at energy gathering than heterotrophs and therefore grow more slowly.

Microbial growth relies on appropriate conditions of total dissolved solids (TDS) concentration, pH and temperature. Most micro-organisms can only

TABLE 2.4 Microbial Metabolism Types in Wastewater Biotreatment

Component	Process	Electron Acceptor	Type
Organic-carbon	Aerobic biodegradation	O_2	Aerobic
Ammonia	Nitrification	O_2	Aerobic
Nitrate	Denitrification	NO_3^-	Facultative
Sulphate	Sulphate reduction	SO_4^{2-}	Anaerobic
Organic-carbon	Methanogenesis	CO_2	Anaerobic

function in relatively dilute solutions, around neutral pH and at ambient temperature, though some can grow under extreme conditions: *Thiobacillus* growth is optimum at pH 1.5−2.0. Some MBRs are based on growth of specific cultures, such as for nitrification (Section 2.2.4.4), or recalcitrant organics biodegradation in extractive MBRs (Section 2.3.2). Classification of micro-organisms according to the temperature at which they are most active provides the terms psychrophilic, mesophilic and thermophilic for optimum growth temperatures of 15, 35 and 55 °C, respectively. While most aerobic biological processes are operated at ambient temperatures, the micro-organisms usually have mesophilic temperature optima, such that pumping operations in sMBRs can provide additional benefit in raising the reactor temperature to both increase biotreatment efficacy and reduce liquid viscosity (Van Dijk & Roncken, 1997). Some examples exist of MBRs operating under thermophilic conditions (Section 5.3.1.7), and this mode appears to offer some promise for treatment of heavy COD loads and/or recalcitrant organic matter.

In both an MBR and activated sludge system, the dominant group of autotrophic bacteria has been shown to be β-subclass Proteobacteria (Manz, Wagner, Amann, & Schleifer, 1994; Sofia, Ng, & Ong, 2004); all currently characterized ammonia oxidizers (i.e. nitrifiers) belong to this group. Although these bacteria are dominant in an MBR, a higher proportion of other bacteria (52−62%) were recorded in these studies, suggesting that the long SRT shifted the microbial population away from Proteobacteria-β (Luxmy, Nakajima, & Yamamoto, 2000; Sofia et al., 2004). *Nitrosomonas* and *Nitrosospira* are the autotrophic ammonia-oxidising bacteria found in activated sludge, and *Nitrobacter* and *Nitrospira* are the nitrite-oxidising bacteria, and it is thus between these groups that the nitrification process is carried out (Wagner, Rath, Amann, Koops, & Schleifer, 1996; Wagner et al., 1998). A number of studies (Sofia, Ng, & Hong, 2004; Li et al., 2005a; Whang, Yang, Huang, & Cheng Whang, 2008) have determined the predominant nitrifiers to be *Nitrosospira* and *Nitrospira*, whilst Witzig et al. (2002) showed no *Nitrosomonas*, *Nitrobacter* or *Nitrosospira* to be found in membrane-filtered sludge. This implies that the ammonia-oxidizing bacteria are system-specific and that *Nitrospira* are responsible for the reduction of nitrite. Nitrifying autotrophs are known to be slow-growing bacteria. The long residence times available in an MBR system are thus highly advantageous for nitrification.

2.2.4. Process Design and Operation Fundamentals

Monod kinetics can be used to design biological systems for a limiting substrate (S kg/m^3), usually organic carbon provided as BOD or COD, or ammonia (NH_4-N). Using known biokinetic constants, the system kinetics and mass balance can be used to define the rate of substrate degradation, nitrification, biomass growth and sludge production (Tchobanoglous et al., 2003).

2.2.4.1. Substrate Degradation

The rate of substrate removal determines the loading rate (the rate at which organic matter is introduced into the reactor, kg BOD/m^3), as dictated by Monod kinetics. Accordingly, the rate of reaction is first order with respect to a limiting substrate up to a maximum specific growth rate, after which growth is unaffected by any further increase in substrate concentration:

$$\mu = \frac{\mu_m S}{K_s + S}, \tag{2.1}$$

where μ and μ_m are the growth rate and maximum specific growth rate per day respectively, S the limiting substrate concentration and K_s is the saturation coefficient, both in g/m^3. It follows that there is a maximum specific substrate utilization rate which is defined as:

$$k = \frac{\mu_m}{Y}, \tag{2.2}$$

where Y is the biomass yield, i.e. the mass of cells formed per mass of substrate consumed in g volatile suspended solids (VSS) per g BOD. Y can be controlled by manipulating environmental factors such as temperature and pH, but such changes may be detrimental to biodegradation in the reactor (Eckenfelder & Grau, 1998). Substituting terms defined by Monod kinetics into a mass balance expression for the system and rearranging produces an expression for the effluent dissolved substrate S_e in g/m^3:

$$S_e = \frac{K_s(1 + k_e \theta_{x,\text{aer}})}{\theta_{x,\text{aer}}(Yk - k_e) - 1}, \tag{2.3}$$

where $\theta_{x,\text{aer}}$ is the aerobic SRT or sludge age (days) and k_e is the decay rate constant. This expression assumes a continuous stirred tank reactor (CSTR); for plug flow the SRT is less, but the high recirculation ratios employed in an MBR are designed to produce complete mixing. The SRT is an important design parameter in suspended growth systems. Since the MBR membrane retains all solids in the reactor complete control of the SRT is provided. The decay rate constant accounts for endogenous metabolism, i.e. the utilization by cells of stored materials, and the EPS (Section 2.3.6.5) associated with the biomass. It also accounts for grazing of the biomass by predatory organisms, as previously discussed. k_e for conventional activated sludge and anaerobic processes is typically in the range of 0.04–0.075/day (Gu, 1993; Tchobanoglous et al., 2003), and according to some authors takes similar values for MBRs (Fan, Urbain, Qian, & Manem, 1996; Wen, Huang, & Qian, 1999). However, experiments by Huang, Gui, and Qian (2001) showed that the endogenous decay in an MBR is higher (0.05–0.32/day) than for an ASP (0.04–0.075/day), and Al-Malack (2006) showed that it varies over a wider range (0.0261–0.151/day) than for an ASP.

Y_{obs}, the observed yield in g/(g/day), is always lower than Y due to the effects of cell decay (i.e. k_e). The relationship between Y_{obs} and Y is governed by the aerobic SRT, $\theta_{x,aer}$, and is defined by:

$$Y_{obs} = \frac{Y}{1 + k_e\theta_{x,aer}} + \frac{f_d k_e \theta_{x,aer}}{1 + k_e\theta_{x,aer}}, \tag{2.4}$$

where f_d is the fraction of the biomass that remains as cell debris, usually 0.1−0.15 g VSS/g substrate (Tchobanoglous et al., 2003). Observed yields (Y_{obs}) are approximately 0.6/day for conventional aerobic processes and an order of magnitude lower for anaerobic ones. Y_{obs} is used to calculate the mass flow rate of heterotrophic biomass produced by a biological system ($M_{x,het}$) for a given flow rate (Q m^3/day):

$$M_{X,het} = Y_{obs}Q(S - S_e). \tag{2.5}$$

2.2.4.2. Nitrification Kinetics

Besides carbonaceous degradation, effluent regulations usually require removal of ammonia, present as the ammonium ion (NH_4^+). Nitrification is the biological generation of nitrate from ammoniacal nitrogen under aerobic conditions (nitrification), and takes place in two distinct stages:

$$2NH_4^+ + 3O_2 \rightarrow 2NO_2^- + 2H^+ + 2H_2O \quad (\text{ammonia} \rightarrow \text{nitrite}) \tag{2.6}$$

$$2NO_2^- + 2O_2 \rightarrow 2NO_3^- \quad (\text{nitrite} \rightarrow \text{nitrate}) \tag{2.7}$$

Overall:

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O \tag{2.8}$$

Since the second step proceeds at a much faster rate than the first, nitrite does not accumulate in most bioreactors. Nitrification relies on sufficient levels of carbon dioxide, ammonia and oxygen, the carbon dioxide providing carbon for cell growth of the autotrophs. Since nitrifiers are obligate aerobes, dissolved oxygen (DO) concentrations need to be 1.0−1.5 mg/L in suspended growth systems for their survival. As already stated, autotrophic nitrifying bacteria grow more slowly than heterotrophic organisms, and thus longer SRTs are required than those for organic carbon degradation; nitrification is thus the determining factor for the aerobic SRT $\theta_{x,aer}$:

$$\theta_{x,aer} = \frac{1}{\mu_n}, \tag{2.9}$$

where μ_n is the specific growth rate of nitrifying bacteria, which can be found from:

$$\mu_n = \left(\frac{\mu_{n,m}N_e}{K_n + N_e}\right)\left(\frac{DO}{K_o + DO}\right) - k_{e,n}, \tag{2.10}$$

where $\mu_{n,m}$ is the maximum specific growth rate of nitrifying bacteria, K_n the half saturation coefficient for nitrification, $k_{e,n}$ the decay rate coefficient for nitrifying bacteria, N_e the required effluent ammonia concentration, DO the dissolved oxygen concentration in the aerobic tank and K_o is the half saturation coefficient for oxygen. The oxygen switching function indicates that nitrification is greatly inhibited at low DO concentrations, inherent to the obligate aerobic nature of nitrifying autotrophs.

Literature values for the nitrification constants, along with the heterotrophic biokinetic data, can be found in Appendix B. Sludge production from nitrification ($M_{x,aut}$, in kg/day) is given by:

$$M_{X,aut} = \frac{QY_n NO_x}{1 + k_{e,n}\theta_{x,aer}}, \tag{2.11}$$

where Y_n is the nitrification sludge yield (gVSS/gNH$_4$-N) and NO_x is the concentration of TKN that is oxidized (mg/L) to form nitrate. To calculate the NO_x, a nitrogen balance can be performed on the system:

$$NO_x = TKN - N_e - 0.12M^{X,bio} \Big/ Q, \tag{2.12}$$

where *TKN* is the influent total Kjeldahl (biochemically oxidizable) nitrogen concentration (TKN, mg/L), while $P_{x,bio}$, the biomass yield, is the sum of $P_{x,het}$ and $P_{x,aut}$. As NO_x is used to determine $M_{x,bio}$, NO_x can be estimated at the first attempt and iterated to find values for NO_x and $M_{x,bio}$.

Near-complete and stable nitrification tends to be more common in full-scale MBR municipal installations than in CASPs (Munz et al., 2008), which is sometimes attributed to the smaller floc size in MBRs which facilitate oxygen transfer within the flocs (Manser, Gujer, & Siegrist, 2005a). However, as in CAS, nitrification is greatly temperature-sensitive with the growth rate of nitrifiers decreasing by 50% for a decrease in temperature of 6 °C (Ekama et al., 1984). As a consequence, ammonia removal generally decreases below 10 °C. An aerobic SRT of around 10 days is required to allow full growth of the nitrifying community (Huang et al., 2001) and, due to the aerobic SRT required for growth of nitrifiers being much longer than that for substrate degradation, it can be reasonably assumed that carbonaceous degradation is complete when nitrification occurs (Tchobanoglous et al., 2003; Grady et al., 2010).

2.2.4.3. Sludge Yield

$P_{x,bio}$ represents the biomass yield (gVSS/d) and results from the growth of heterotrophic biomass, endogenous decay and growth of nitrifying biomass. However, wastewaters often contain fractions of volatile particulate non-biodegradable organics *nb*VSS (up to 25–33% of VSS) and inert total suspended solids (*i*TSS) that accumulate in the mixed liquor and are not degraded. To obtain the total sludge yield $M_{x,TSS}$ (g mixed liquor suspended solids (MLSS)/day), it is necessary to consider contributions from both fractions:

$$M_{X,\mathrm{TSS}} = M^{X,\mathrm{bio}}/R_{\mathrm{SS}} + Q(nb\mathrm{VSS} + i\mathrm{TSS})$$
$$= \frac{\left(\dfrac{QY(S-S_{\mathrm{e}})}{1+k_e\theta_{x,\mathrm{aer}}} + \dfrac{f_{\mathrm{d}}k_eQY(S-S_{\mathrm{e}})\theta_{x,\mathrm{aer}}}{1+k_e\theta_{x,\mathrm{aer}}} + \dfrac{QY_{\mathrm{n}}(\mathrm{NO}_x)}{1+k_{e,n}\theta_{x,\mathrm{aer}}}\right)}{0.85} + Q(\mathrm{nbVSS} + \mathrm{iTSS}), \quad (2.13)$$

where R_{SS} is the VSS/MLSS ratio in the biomass and is ~0.85, based on the typical composition of bacterial cells (Tchobanoglous et al., 2003). Equation (2.13) shows that zero sludge production is not possible in an MBR; at infinite $\theta_{x,\mathrm{aer}}$ and zero VSS and TSS influent levels the daily sludge production corresponds to $fdYQ(S-S_{\mathrm{e}})$ pertaining to the cell debris, i.e. the non-degradable component of the micro-organisms formed through growth on the biodegradable substrate. This is also the theoretical minimum sludge production possible.

Changing θ_x (i.e. the SRT) has by far the greatest impact on sludge production (Xing, Wu, & Tardieu, 2003) and allows a desired operational MLSS concentration (X_{aer} g/m^3) to be set. The MLSS concentration then affects the aeration demand (through the α-factor, Section 2.2.5) and membrane fouling and clogging propensity (Section 2.3.6.3). Using a design MLSS and aerobic SRT (as determined by the required effluent ammonia concentration N_{e}) the aeration tank volume V_{aer} can be calculated by obtaining the mass of solids being aerated, and then using the aerobic MLSS to convert that mass to the volume which those solids occupy:

$$V_{\mathrm{aer}} = \frac{M_{X,\mathrm{TSS}}\theta_{x,\mathrm{aer}}}{X_{\mathrm{aer}}}. \quad (2.14)$$

2.2.4.4. Nutrient Removal

If effluent limits mandate removal of total nitrogen (TN), addition of a denitrification step to remove the nitrate produced in the nitrification process is required. Denitrification occurs under anoxic conditions when oxidation of the organic carbon takes place using the nitrate ion (NO_3^-), generating molecular nitrogen (N_2) as the primary end product:

$$C_{10}H_{19}O_3N + 10NO_3^- \rightarrow 5N_2 + 10CO_2 + 3H_2O + NH_3 + 10OH^-, \quad (2.15)$$

where in this equation '$C_{10}H_{19}O_3N$' represents dissolved organic material wastewater. Under anoxic conditions, facultative micro-organisms, which normally remove BOD under aerobic conditions, are able to convert nitrates to nitrogen gas. Denitrification requires a sufficient carbon source for the heterotrophic bacteria. This can be provided by the raw wastewater, such that the nitrate-rich waste from the aerobic zone can be recycled to mix with the raw wastewater. Most full-scale MBR sewage treatment plants are also designed to achieve denitrification, with an anoxic zone usually incorporated before the aerobic and membrane tanks. The required volume of the anoxic zone can be derived through an iterative process, subject to the constraint that the denitrification capacity (NO_r, in kgNO_3-N/day) has to be larger than the nitrate load (NO-loading, in kgNO_3-N/day) inferred from the aerobic zone through the recirculation flow (Q_{int}, in m^3/day):

$$NO_r = V_{anox} X_{b,anox} SDNR, \tag{2.16}$$

$$\text{NO-loading} = Q r_{int} NO_x. \tag{2.17}$$

The nitrate recirculation ratio r_{int} and the active biomass in the anoxic zone $X_{b,anox}$ can be respectively found from:

$$r_{int} = \frac{NO_x}{NO_e} - 1. \tag{2.18}$$

$$X_{b,anox} = \left(\frac{Q\theta_{x,aer}}{V_{aer}}\right)\left(\frac{Y(S-S_e)}{1+k_e\theta_{x,aer}}\right)\left(\frac{r_{int}}{r_{int}+1}\right), \tag{2.19}$$

where NO_e is the desired effluent nitrate concentration (g/m^3). The specific denitrification rate (SDNR, in gNO_3-N/gVSS) can be determined empirically (Fig. 2.18) from the ratio of food to active biomass in the anoxic zone F/M_b:

$$F/M_b = \frac{QS}{V_{anox} X_{b,anox}}, \tag{2.20}$$

where the significance of the *F/M* ratio is discussed in Section 2.2.4.5. For a determined denitrification capacity insufficient to denitrify the incoming nitrate load, V_{anox} has to be adjusted and the procedure reiterated until a value for V_{anox} arises which ensures sufficient denitrification capacity.

Most wastewaters treated by biological processes are carbon limited, and hence phosphorus is not significantly removed. This applies as much to MBRs

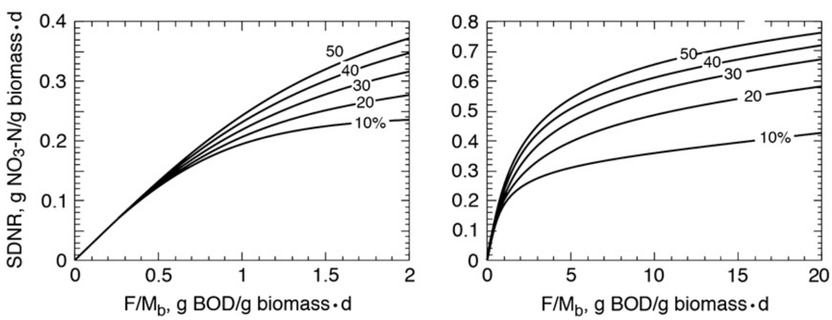

FIG. 2.18 Nitrification kinetics curves *(from Tchobanoglous et al., 2003).*

as to conventional plants. It appears that membrane separation offers little or no advantage regarding phosphorus removal (Yoon, Kang, & Lee, 1999). Enhanced biological phosphate removal can be achieved by the addition of an anaerobic zone at the front of an activated sludge plant and returning nitrate-free sludge from the anoxic zone (Ekama et al., 1984; Yeoman, Stephenson, Lester, & Perry, 1986). This has been applied to some full-scale MBR plants where constraints on discharged P levels have been imposed (Daigger, Rittmann, Adham, & Andreottola, 2005; Daigger, Crawford, & Johnson, in press), and has formed the basis of a great many MBR studies (Lesjean et al., 2005a; Bracklow, Drews, Vocks, & Kraume, 2007; Abegglen, Ospelt, & Siegrist, 2008). When bio-P removal is combined with dosing of chemicals such as metal coagulants or lime, which can form sparingly soluble precipitates, effluent P concentrations lower than 0.1 mg/L can reliably be achieved (Fleischer et al., 2005). However, design and process control is critical when enhanced nutrient control is to be achieved (Daigger et al., 2010), viz.: (1) the membrane recirculation flow has to be directed to the aerobic zone; (2) intense mixing has to be provided at the inlets of the anaerobic and anoxic zones; (3) internal recirculation rates have to be controlled to maintain the desired MLSS distribution; and (4) supplemental metal salt addition has to be carefully controlled in proportion to the residual phosphorus following biological P removal. A possible added benefit of adding alum to increase P-removal is that it may have a beneficial effect on membrane fouling, reducing organic fouling and improving floc structure and strength (Holbrook et al., 2004; Fleischer et al., 2005).

More authoritative and extensive treatises on the biological nutrient process are available elsewhere in biological wastewater treatment textbooks (Tchobanoglous et al., 2003; Henze et al., 2008; Grady, Daigger, & Love, 2010).

2.2.4.5. Process SRT

The slow rate of microbial growth demands relatively long HRTs (compared with chemical processes), and hence large-volume reactors. Alternatively, retaining the biomass in the tank either by allowing them to settle out and then recycling them, as in an ASP, fixing them to porous media, such as in a TF, or rejecting them with a perm-selective barrier, as with an MBR, permits longer SRTs without requiring the HRT to be commensurately increased. As stated above, controlling the aerobic SRT in a biological system determines the rate of substrate degradation, nitrification, excess sludge production and biomass concentration (Equations (2.3), (2.9), (2.13) and (2.14), respectively). The total process SRT can be found by adding the anoxic and aerobic SRT, and is controlled by periodically discharging some of the solids (sludge) from the process:

$$\theta_{x,\mathrm{process}} = \theta_{x,\mathrm{aer}} + \theta_{x,\mathrm{anox}} = \frac{V_{\mathrm{aer}}X_{\mathrm{aer}} + V_{\mathrm{anox}}X_{\mathrm{anox}}}{Q_{\mathrm{w}}X_{\mathrm{aer}}}, \tag{2.21}$$

where Q_w is the sludge wastage rate (m^3/day), and it is assumed that the solids wasted from the reactor are at the same concentration as those within it. In order to correctly control the aerobic SRT, the volume of sludge wasted Q_w becomes:

$$Q_w = \frac{V_{aer}}{\theta_{x,aer}}. \tag{2.22}$$

An often-quoted ASP empirical design parameter is the food-to-microorganism ratio (F/M in units of inverse time), which defines the rate at which substrate is fed into the tank (SQ, Q being the volumetric feed flow rate in m^3/day) compared to the mass of reactor solids:

$$F/M = \frac{SQ}{VX}. \tag{2.23}$$

This relates to the aerobic SRT $\theta_{x,aer}$ and the process efficiency E (%) by:

$$\frac{1}{\theta_{x,aer}} = Y(F/M)\frac{E}{100} - k_e. \tag{2.24}$$

SRT values for activated sludge plants treating municipal wastewaters are typically in the range of 5−15 days with corresponding F/M values of 0.2−0.4/day. Increasing SRT increases the reactor biomass (or MLSS) concentration. Conventional ASPs operating at SRTs of ~8 days have an MLSS of around 2.5 g/L, whereas one with an SRT of ~40 days might have an MLSS of 12−15 g/L. A low F/M ratio implies a high MLSS and a low sludge yield, such that increasing SRT is advantageous with respect to waste generation. On the other hand, high MLSS values are to some extent detrimental to process performance. First they lead to an accumulation of inert compounds which is reflected in a decrease in the MLVSS/MLSS ratio where MLVSS represents the volatile (organic) fraction of the MLSS, though this does not appear to be the case in practice (Huang, Gui, & Quian, 2001; Rosenberger, Kraume, & Szewzyk, 1999); high MLSS levels do not appear to impair biodegradation (Pollice, Laera, Saturno, & Giordano, 2008). Second, high solids levels increase the propensity for clogging or 'sludging' − the accumulation of solids in the membrane channels (Section 3.6.2). Lastly, and possibly most significantly, high MLSS levels reduce aeration efficiency (Section 2.2.5).

There have been a number of studies where the characteristics and performance of CASPs and MBRs have been compared when these processes operated under the same conditions of HRT and SRT. Massé, Spérandio and Cabassud (2006) reported that sludge characteristics and biological performance differ for ASP and MBR, and that the difference increases as SRT increases. Deterioration in effluent concentration arises for the ASP due to poor sludge settlability, filamentous bacteria and protein and polysaccharide (or carbohydrate) release. The MBR effluent quality was superior, mainly due to complete retention of TSS. Ghyoot and Verstraete (2000), in their studies using a skimmed milk-based analogue feed, observed sludge yields to be lower for an MBR than for an ASP (0.22 vs 0.28 and 0.18 vs 0.24 gVSS/d for operation at 12 and 24 days SRT, respectively). This trend was also reported by Smith, Judd, Stephenson and Jefferson (2003), who noted the greatest impact of the membrane separation to be on K_s, which decreased from 125 ± 22 to 11 ± 1 g/m^3 for the ASP compared to a corresponding increase from 2 ± 1.6 to 73 ± 22 g/m^3 for the MBR. Al-Malack (2006) also reported MBR K_s values (289−2933 g/m^3) significantly higher than those typically reported for CAS. Given that K_s is inversely proportional to substrate affinity, the generally lower values of K_s in the case of an MBR suggest a greater biomass substrate affinity, and also that the growth rate is less influenced by substrate concentration. Smith and co-workers proposed that this related to the difference in floc size, since the corresponding specific surface areas of the two biomasses at 30-day SRT were 0.098 m^2/g for the MBR and 0.0409 m^2/g for the ASP, revealing that the MBR biomass provides over 230% more surface area at about the same MLSS concentration. This was corroborated by Manser, Gujer, & Siegrist (2005a), who suggested that the smaller floc size in MBR also benefits nitrification kinetics through the oxygen half saturation coefficient, the maximum nitrification rate being unaffected (Manser, Gujer, & Siegrist, 2005b).

2.2.5. Aeration

2.2.5.1. Mass Balance

In conventional aerobic biological wastewater treatment processes, oxygen is usually supplied as atmospheric air, either via immersed air-bubble diffusers or surface aeration. Diffused air bubbles (via fine bubble aeration) are delivered to the bulk liquid (as in an ASP, a biological/submerged aerated filter (BAF/SAF), fluidized bioreactors, etc.), or oxygen transfer occurs from the surrounding air to the bulk liquid via a liquid/air interface (as for a TF or a rotating biological contactor (RBC)).

The oxygen requirement to maintain a community of micro-organisms and degrade BOD and ammonia and nitrite to nitrate can be found by a mass balance on the system (Tchobanoglous et al., 2003):

$$R_0 = Q(S - S_e) - 1.42P_{X,\mathrm{bio}} + 4.33Q\mathrm{NO}_x - 2.86Q(\mathrm{NO}_x - NO_e), \quad (2.25)$$

where R_o is the total oxygen required (kg/d). The first term in Equation (2.25) refers to substrate oxidation, the second refers to biomass respiration, the third refers to nitrification and the final one to denitrification (Section 2.2.6). Certain terms thus disappear from the expression depending on whether or not the system is nitrifying and/or denitrifying.

2.2.5.2. Mass Transfer

Mass transfer of oxygen into the liquid from air bubbles is defined by the overall liquid mass transfer coefficient (k_L m/s) and the specific surface area for mass transfer (a m^2/m^3). Because of the difficulties associated with measuring k_L and a individually, the two are usually combined to give the volumetric mass transfer coefficient k_La (per unit time). The standard method accepted for determining k_La in clean water is detailed in ASCE (1992). The rate of oxygen transfer into a liquid can be determined by:

$$\mathrm{OTR}_{\mathrm{cleanwater}} = k_La(C^* - C), \tag{2.26}$$

where C and C^* are the dissolved and saturated oxygen concentration values in kg/m^3. For pure water and equilibrium conditions C is found using Henry's Law. This can be converted to process conditions by the application of three correction factors (α, β and ϕ) which account for those sludge properties which impact on oxygen transfer (Section 2.2.5.3):

$$\mathrm{OTR}_{\mathrm{cleanwater}} = \frac{\mathrm{OTR}_{\mathrm{process}}}{\alpha\beta\phi}. \tag{2.27}$$

Aeration also provides agitation to ensure high mass transfer rates and complete mixing in the tank. There is thus a compromise between mixing, which demands larger bubbles, and oxygen dissolution, which demands small, indeed microscopic, bubbles (Garcia-Ochoa, Castro, & Santos, 2000). Consequently oxygen utilization, the amount of oxygen in the supplied air which is used by the biomass, can be as low as 10%, and decreases with increasing biomass concentration (Equation (2.25)). This can be quantified by the standard aeration efficiency (kg O_2/kWh):

$$\mathrm{SAE} = \frac{\mathrm{OTR} \cdot V}{W}, \tag{2.28}$$

where W is the power demand. The OTR into the mixed liquor can be increased by using oxygen-enriched air, but this increases costs and is rarely used other than for high-strength effluents when the oxygen limitation is reached. In an iMBR, additional aeration is also required for scouring of the membrane (Section 2.3.7.1).

Changes in airflow have been shown to produce the largest changes in mass transfer in a coarse bubble aeration system (Ashley, Mavinic, & Hall, 1992), with k_La increasing with gas velocity in an air-lift reactor (Lazarova, Julián, Laurent, & Jaques, 1997; Masoud, Sohrabi, Vahabzadeh, & Bonakdarpour, 2001). Nordkvist, Grotkjaer, Hummer and Villadsen (2003) proposed that both the liquid and gas velocities impact on mass transfer, confirmed by experiments based on a jet loop MBR by Kouakou, Salmon, Toye, Marchot and Crine (2005). However, the authors of this paper also noted a linear relationship between the mass transfer coefficient and the liquid recirculation velocity. Also, increasing horizontal velocity has been shown to increase the value of k_La in an oxygen ditch in both pilot (Gillot, Capela, & Heduit, 2000) and full-scale plants (Deronzier, Gillot, Duchène, & Héduit, 1996).

2.2.5.3. Correction for Temperature and Sludge Characteristics

Φ in Equation (2.27) relates to the effect of temperature on the mass transfer:

$$k_{\mathrm{L}}a_{(T)} = k_L a_{(20\,^\circ\mathrm{C})}\Phi^{(T-20)}, \tag{2.29}$$

where T is the temperature (°C) and Φ is a constant. Typical values are between 1.015 and 1.040 with 1.024 being the ASCE standard (Iranpour et al., 2000) for temperature correction of viscosity η:

$$\frac{\eta_T}{\eta_{20}} = 1.024^{(20-T)}. \tag{2.30}$$

Salts and particulates in wastewater both impact on the oxygen transfer rate. Comparative tests on synthetic wastewater and tap water performed by Lazarova, Julián, Laurent, & Jaques. (1997) showed that below 2 g/L salt concentration has little effect on the oxygen transfer. Kouakou et al. (2005) performed comparative studies between clean water and wastewater with a salt concentration of 0.48 g/L and found the mass transfer coefficients did not significantly vary. The effect of such constituents is accounted for by the β factor which is defined as:

$$\beta = \frac{C^*_{(\mathrm{wastewater})}}{C^*_{(\mathrm{cleanwater})}}, \tag{2.31}$$

and is usually around 0.95 for wastewater (EPA, 1989).

Both biomass characteristics and aeration system design impact on oxygen transfer (Müller, Boyle, & Popel, 2002), and impacts have been reviewed by Schwarz, Rittmann, Crawford, Klein and Daigger (2006). Biomass is a heterogeneous mixture of particles, micro-organisms, colloids, organic polymers and cations of various sizes and surface properties which can all impact on oxygen transfer through contact area and surface energy. Bubble characteristics differ depending on the aerator type and bubble stability, the latter being influenced by the biomass characteristics and promotion of bubble coalescence. At the same time, biological and physical characteristics of the mixed liquor are affected by the shear imparted by the airflow, which can fragment flocs (Abbassi, Dullstein, & Rabiger, 1999) and cause the release of chemicals, as well as impacting on biodiversity. The inter-relationships developed between aeration and various system facets and parameters are thus complex, especially given that, for an iMBR, aeration is also used for membrane scouring (Fig. 2.19). This complex relationship is usually accounted for by the α-factor. The α-factor has the most significant impact on aeration efficiency of all three conversion factors. It is accepted that α-factor is a function of SRT (mean cell retention time), air and liquid

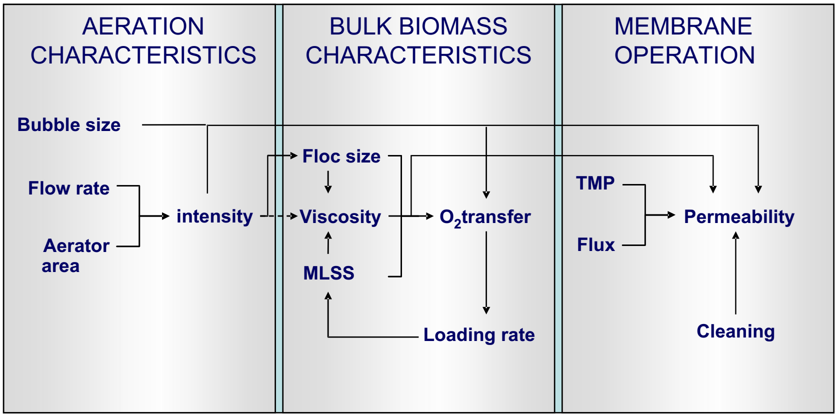

FIG. 2.19 Aeration impacts in an iMBR *(adapted from Germain, 2004)*.

flow rate and of tank geometry for a given wastewater (Rosso & Stenstrom, 2005), and is defined as:

$$\alpha = \frac{k_L a_{\text{wastewater}}}{k_L a_{\text{cleanwater}}}. \tag{2.32}$$

Wastewater composition and, in particular, the level of surfactants affect the bubble size, shape and stability. Surfactants are found in detergents of all kinds, including washing up liquid, laundry powder and soap. A high concentration of contaminants builds up on the outside of the bubble, reducing both the diffusion of oxygen into solution and the surface tension. Reduced surface tension has the beneficial effect of reducing bubble size, thereby increasing the water–air interfacial area (a). Fine bubble aeration systems are most negatively affected by surfactants, since bubbles produced are already small and cannot be further reduced in size by a reduction in surface tension (Stenstrom & Redmon, 1996). It has been shown from experiments testing oxygen transfer in waters containing different surfactants that the ratio of mass transfer from surfactant water to clean water varies between 1.03 and 0.82 (Gillot, Capela, & Hedvit, 2000). However, surfactants have a negative effect on ASP processes overall due to the promotion of foaming (Sections 2.3.6.3 and 3.6.5).

Studies of the impact of solids concentration on oxygen transfer in biological wastewater treatment systems have all indicated a decrease in OTR with increasing solids concentration regardless of the system studied, though the relationship is system and feedwater dependent (Chang, Lee, & Ahn, 1999; Chatellier and Audic, 2001; Fujie, Hu, Ikeda, & Urano, 1992; Germain et al., 2007; Günder, 2001; Krampe & Krauth, 2003; Lindert, Kochbeck, Pruss, Warnecke, & Hempel, 1992; Müller et al., 1995). In a number of studies of sewage treatment, an exponential relationship between α-factor and MLSS

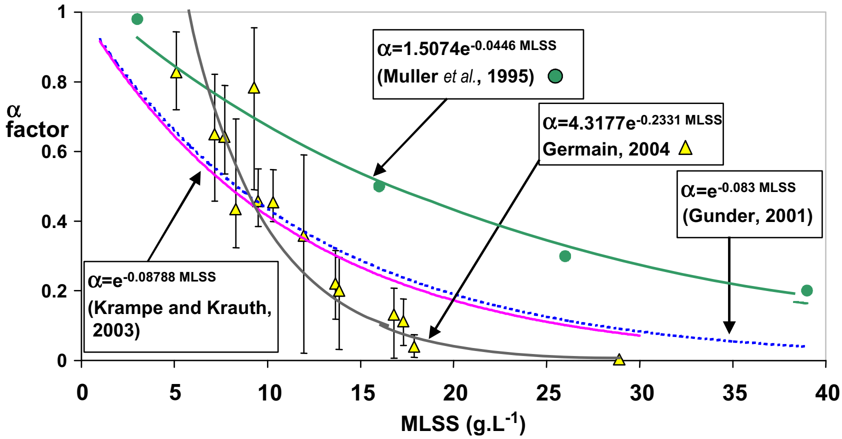

FIG. 2.20 α-Factor vs MLSS concentration.

concentration has been observed. Müller et al. (1995), recorded α-factor values of 0.98, 0.5, 0.3 and 0.2 for MLSS concentrations of 3, 16, 26 and 39 g/L, respectively, yielding an exponent value of -0.045 and an R^2 of 0.99 (Fig. 2.20). Günder (2001) and Krampe and Krauth (2003) observed the same exponential trend with exponent values of -0.083 and -0.088, respectively, whereas an even higher exponent value of -0.23 was recorded by Germain et al. (2007). Studies on model or simplified systems by a number of authors (Verlaan & Tramper, 1987; Freitas & Teixeira, 2001; Özbek & Gayik, 2001) appear to indicate that the principal impact of solids concentration is on the interfacial area a, which decreases with increasing solids level whilst leaving the mass transfer coefficient k_L largely unaffected. This has been attributed to the promotion of bubble coalescence by suspended solids (Klein et al., 2002), and the effect is also aeration rate-dependent (Freitas & Teixeira, 2001). Since MBRs run at longer SRTs than a conventional ASP the oxygen demand, and thus the volumetric aeration demand, for biotreatment is somewhat higher.

The impact of particle size is more complex than particle concentration, since aeration, mass transfer and particle size are interrelated. For fine particles, ~0.01 mm, k_La has been shown to increase with increasing solids concentration up to a certain level and remain stable, before decreasing with further increased solids concentration (Saba, Kumazawa, Lee, & Narukawa, 1987; Smith & Skidmore, 1990). With larger particles, 1−3 mm, k_La appears to decrease with concentration (Koide, Shibata, Ito, Kim, & Ohtaguchi, 1992; Lindert et al., 1992; Komaromy & Sisak, 1994; Hwang & Lu, 1997; Nakao, Harada, Furumoto, Kiefner, & Popovic, 1999). Experiments examining excess sludge production, in which the DO concentration was adjusted independently of aeration intensity, have indicated that higher mixing intensity created by raising the airflow has almost the same impact on floc break-up and therefore on particle size. At a sludge loading of 0.53 kg BOD_5/(kg MLSS day), the excess sludge production was reduced by 22% by raising the oxygen concentration from 2 to 6 mg/L (Abbassi, Dullstein, & Rabiger, 1999). However, the principal impact of particle size in an MBR is on filter cake permeability, as indicated by the Kozeny Carman equation.

Viscosity correlations are complicated by the non-Newtonian pseudoplastic nature of the sludge, but it has nonetheless been shown to have a negative influence on the oxygen transfer coefficient (Koide et al., 1992; Garcia-Ochoa, Casfro, & Santos, 2000; Jin, Yu, Yan, & van Leeuwen, 2001; Badino, Facciotti, & Schmidell, 2001; Özbek & Gayik, 2001). This has variously been attributed to bubble coalescence and solubility impacts, with larger bubbles forming (Özbek & Gayik, 2001) and greater resistance to mass transfer recorded (Badino, Facciotti, & Schmidell, 2001) at higher viscosity. Air is also less well distributed at higher viscosities, the smaller bubbles becoming trapped in the reactor (Jin Yu, Yan, & van Leeuwen, 2001).

Correlations between the α-factor and viscosity (η in kg/(ms)) have been presented, these correlations being more pronounced than those between α-factor and MLSS concentration (Wagner, Cornel, & Krause, 2002). Relationships presented take the form:

$$\alpha = \eta^{-x}, \tag{2.33}$$

where $x = 0.45$ (Günder, 2001) or 0.456 (Krampe & Krauth, 2003) at a shear rate of 40 s^{-1} in activated sludge of high MLSS concentrations. The correlation is shear-dependent: increasing shear stress decreases viscosity (Dick & Ewing, 1967; Wagner, Cornel, & Krause, 2002). An increase in aeration rate therefore offers the dual benefit to oxygen transfer in that it increases the amount of available oxygen and also decreases biomass viscosity by increasing shear stress. Viscosity has been shown to increase both exponentially and linearly with increasing MLSS concentration (Manem & Sanderson, 1996; Rosenberger, Kraume, & Szewzyk, 1999), in both cases impacting negatively on both oxygen transfer and membrane fouling (Section 2.3.6.3).

2.2.6. Anaerobic Treatment

Compared with aerobic processes, anaerobic biological treatment is characterized by (Stephenson, Judd, Jefferson and Brindle, 2000):

- a lower energy demand due to the absence of aeration
- slower microbial growth
- a lower COD removal (generally 60–90%)
- no nitrification
- greater potential for odour generation
- longer start-up (months cf. weeks)
- higher alkalinity
- lower sludge production
- biogas (methane) generation.

Conventional anaerobic treatment process configurations are all designed to achieve both good mixing and sludge separation. A number of configurations exist:

(a) Simple contacting coupled with external sludge separation (by sedimentation, rotary vacuum filtration, etc.) and/or digestion before returning the clarified liquid to the reactor. This is a simple and relatively easily controlled process but is also made expensive by the pumping operations.
(b) Anaerobic filters, which are flooded media filters based on either packed or structured media.
(c) Upflow clarification using the upflow anaerobic sludge blanket (UASB, Fig. 2.21a) reactor (Lettinga & Vinken, 1980), in which sludge particles settle at the same rate as the water flows upwards, forming a stationary 'blanket' of sludge in the reactor. This process relies on the formation of a dense granular sludge bed that is readily retained in the reactor, in much the same way as secondary clarification in the ASP relies on the growth of large settleable particles. The process is augmented in the

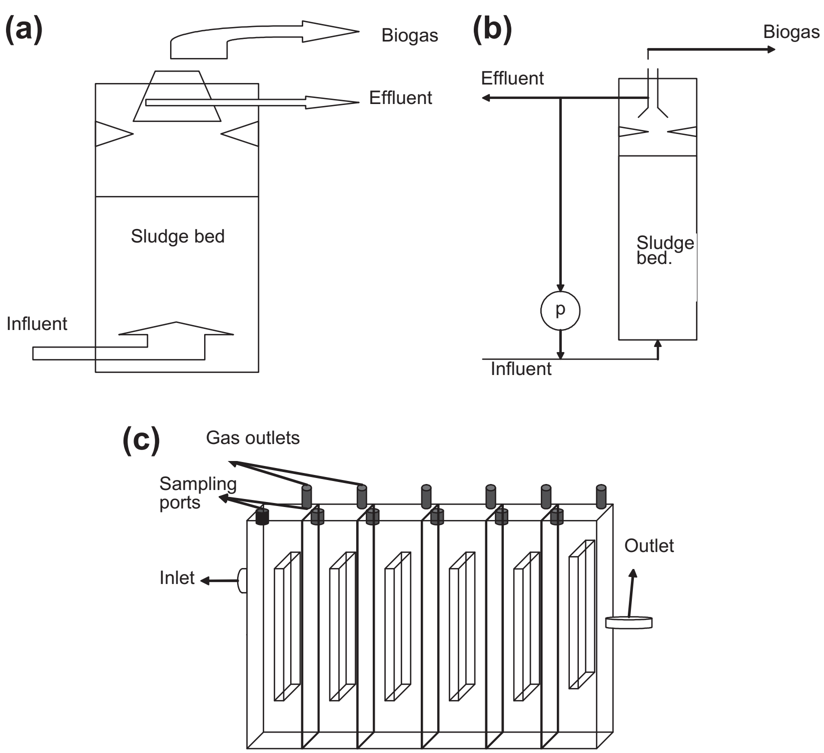

FIG. 2.21 Schematics: (a) UASB and (b) EGSB reactors *(modified from Seghezzo et al., 1998, 2002)* and (c) the ABR *(modified from Dama et al., 2002).*

expanded granular sludge bed (EGSB, Fig. 2.21b) which provides better influent distribution to improve contact between the sludge and wastewater and so promotes more efficient use of the entire reactor volume (Seghezzo, Zeeman, Van Lier, Hamelers, & Lettinga, 1998). In this configuration, the sludge bed is expanded by operating at higher upflow rates and the reactor behaves as a completely mixed tank (Rinzema, 1988).

(d) Staged reactor systems, based on plug flow (Van Lier, 1995) and using sequentially operated reactors or compartments within a single reactor. Staged reactors include anaerobic baffled reactors (ABRs, Fig. 2.21c) in which baffles are used to direct the flow of wastewater in an upflow mode through a series of sludge blanket reactors (Tchobanoglous et al., 2003). The process is more tolerant to non-settling particles than the UASB and EGSBs whilst still providing long, solid retention times.

Anaerobic treatment is generally only considered for high-strength wastes and where low feed temperatures are less likely to be encountered. Low feed temperatures and strength imply low biomass growth yield and growth rate, such that the biomass concentration in the reactor is more difficult to sustain, particularly when substantial biomass wash-out from the reactor can occur. MBRs ameliorate this problem to a large extent, such that the range of anaerobic process operation can be extended to lower limits. This is achieved by the retention of the biomass in the reactor by the membrane independently of the HRT in the same way as for aerobic systems; significant quantities of residual organic matter are hydrolysed and biodegraded as a result. However, the membrane fouling propensity of the bioreactor liquor is significantly higher for anaerobic treatment (Section 2.3.6.6), such that fluxes and permeabilities are generally much lower than for the aerobic counterparts.

2.3. MEMBRANE BIOREACTOR TECHNOLOGY

A classical MBR comprises a conventional activated sludge process (CASP) coupled with membrane separation to retain the biomass. Since the effective pore size is generally below 0.1 μm, the MBR produces a clarified and substantially disinfected effluent. In addition, it concentrates up the biomass and, in doing so, reduces the necessary tank size and also increases the efficiency of the biotreatment process. MBRs thus tend to generate treated waters of higher purity with respect to dissolved constituents such as organic matter and ammonia, both of which are significantly removed by biotreatment. Moreover, by removing the requirement for biomass sedimentation (Section 2.2), the flow rate through an MBR cannot affect product water quality through impeding solids settling, as is the case for the CASP. On the other hand, hydraulic and organic shocks can have other onerous impacts on the operation of an MBR.

2.3.1. MBR Configurations

2.3.1.1. *Aerobic Processes*

The word 'configuration' can be used with reference to both the MBR process (and specifically how the membrane is integrated with the bioreactor) and the membrane module (Section 2.1.3). There are two main MBR process configurations (Fig. 1.1): submerged or immersed (iMBR), and sidestream (sMBR). There are also two modes of hydraulic operation: pumped and air-lift. These configurations and bulk liquid transfer modes are employed commercially for what can be referred to as conventional biomass rejection MBRs, as outlined above. However, there are also two other membrane process modes, these being extractive (eMBR) and diffusive (dMBR) (Fig. 2.22), which employ a membrane for a purpose other than to separate the biomass from the treated water. Finally, whilst a number of membrane geometries and configurations exist in the membrane market place in general (Table 2.2), three predominate in existing commercial MBR technologies, these being flat sheet (FS), hollow fibre (HF) and multitube (MT). Examples of each type of commercial technology are detailed in Chapter 4.

iMBRs are generally less energy-intensive than sMBRs, since employing membrane modules in a pumped sidestream crossflow incurs an energy penalty due to the high pressures and volumetric flows imposed. To make the most use of this latent energy, the flow path must be as long as possible, such that the maximum amount of kinetic energy intrinsic in the liquid flowing at high pressure is harnessed for permeation. To achieve a reasonable conversion of 40–50% conversion along the length of the module, a long flow path is required, often in excess of 10 m. This then demands a large number of

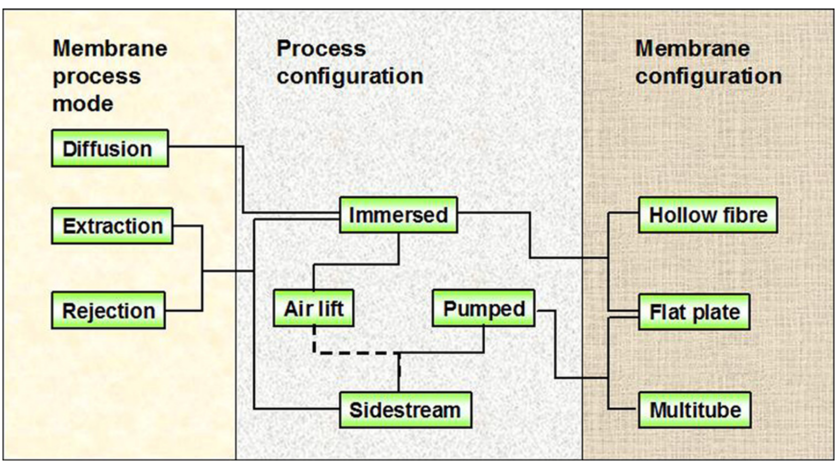

FIG. 2.22 Principal configurations of MBR technologies.

FIG. 2.23 sMBR pilot plant: the total retentate fluid path is four times the length of one horizontal module.

membrane modules in series (Fig. 2.23) incurring a significant pressure drop along the retentate flow channels.

With sMBRs, there is always a trade-off between pumping energy demand and flux. To maximize the flux, a high TMP is required combined with a high crossflow velocity (CFV or retentate velocity U_R). Since the energy demand is directly proportional to $Q_R \Delta P$ (retentate flow rate × pressure), then it is of interest to reduce both these parameter values as much as possible. However, since Q_R determines U_R ($U_R = Q_R/A_t$, A_t being the tube cross-sectional area) and ΔP relates to TMP, reducing $Q_R \Delta P$ inevitably reduces flux. Moreover, if Q_R is reduced by decreasing the cross-sectional area A_t, this has the effect of increasing the pressure drop along the length of the module on the retentate side, since the resistance to flow is inversely proportional to A_t.

sMBRs have an inherently higher fouling propensity than iMBRs since higher flux operation always results in lower permeabilities because fouling itself increases with increasing flux, particularly above the so-called 'critical flux' (Section 2.1.4.6). Moreover, it is thought that the higher shear imparted by liquid pumping of the sidestream imparts sufficient shear stress on the flocs to cause them to break up (Tardieu, Grasmick, Geaugey, & Manem, 1999; Wisniewski & Grasmick, 1998). This both reduces particle size and promotes the release of foulant materials bound within the flocs (EPS, Section 2.3.6.5). Wisniewski and Grasmick (1998) studied the effects of sludge recirculation on the particle size in an sMBR. Without recirculation, floc size ranged from 20 μm to more than 500 μm. Only 15% of the particles were lower than 100 μm. With recirculation, reduction in particle size was directly proportional to the magnitude of the shear stress and the experiment time; at 5 m/s CFV 98% of the particles were smaller than 100 μm. iMBRs are therefore higher in energy efficiency, manifested as the specific energy demand in kWh per m^3 permeate product, than sMBR technologies. The immersed configuration employs no liquid pumping for permeation, instead relying on aeration to promote mass transfer of liquid across the membrane (i.e. enhancing flux) by generating significant transient shear at the membrane:solution interface (Section 2.1.4.4). Shear can also be promoted by directly moving the membrane, such as in the recently introduced Grundfos Biobooster system (Section 4.4.4).

Whilst sMBRs cannot provide the same low energy demand as the immersed configuration, they do offer a number of advantages:

1. Fouling has been shown to decrease linearly with increasing crossflow velocity (CFV). For example a bench-scale study revealed that CFV values of 2 and 3 m/s were sufficient to prevent the formation of reversible fouling in UF (30 kDa) and MF (0.3 μm) systems, and that fouling was suppressed for CFV values up to 4.5 m/s (Choi, Zhang, Dionysiou, Oerther, & Sorial, 2005b).
2. The membranes can also be chemically cleaned in place (CIP) easily without any chemical risk to the biomass.
3. Membrane 'loops' can be easily brought on- and off-line during periods of high and low flow, respectively.
4. Maintenance and plant downtime costs, particularly with reference to membrane module replacement, are generally slightly lower because of the accessibility of the modules which can be replaced in 10−20 min.
5. Precipitation of sparingly soluble inorganic solids (i.e. scalants) and organic matter (gel-forming constituents) is more readily managed in sidestream MT systems by control of the hydrodynamics both during the operation and the CIP cycle.
6. It is generally possible to operate sMBRs at higher MLSS levels than HF iMBRs.
7. Aeration can be optimized for high oxygen transfer, rather than demanding a compromise between membrane aeration and oxygen dissolution, as would be the case for single-tank iMBRs.

A configuration which would appear to combine some of the advantages of both the immersed and sidestream configuration is the air-lift sidestream (a-lsMBR). In this configuration, the multitube modules are oriented vertically outside the tank and a combination of sludge and air pumping used to flush the sludge along the length of the modules. There are currently a number of installations based on this configuration, with applications tending to be for domestic/municipal wastewater (Section 5.4.1.3) rather than the higher strength industrial effluents for which the pumped sidestream configuration is generally preferred.

2.3.1.2. *Anaerobic Processes*

The process configuration options for the anaerobic MBR (anMBR) are essentially the same as for the aerobic one: pumped and gas-lift sidestream (Fig. 2.24a) and immersed (Fig. 2.24b), with a further configuration employing vacuum extraction from an immersed sidestream module (Fig. 2.24c). For the anMBR, however, membrane scouring with air is obviously not an option. Instead scouring and lifting of the sludge through the membrane channels must either employ liquid pumping or the generated biogas.

As with aerobic systems, the sidestream anaerobic configuration (ansMBR) was commercially established before the immersed one, with the latter only introduced in the late 1990s, and full-scale fermentation applications for FS configurations have been reported (Kanai, Ferre, Wakahara, Yamamoto, & Moro, 2010). AnsMBRs generally operate at CFVs and TMPs of 1−5 m/s and 2−7 bar, respectively, to provide reasonable fluxes. Much lower pressures (0.2−1 bar) arise in immersed systems, though these are still higher than in corresponding aerobic iMBRs. CFVs in immersed systems have been reported as being less than 0.6 m/s (Bérubé & Lei, 2006b).

An example of a proprietary pumped ansMBR, the BIOREK® process, is given in Section 4.4.5, and that of an aniMBR given in Section 5.2.1.8. Both these examples, as with most membrane-based anaerobic bioreactor processes, use the CSTR. More limited application of membranes is found in upflow anaerobic sludge bed (UASB) reactors. For the CSTR the membrane is exposed to the MLSS solids at 10−40 g/L, whereas in the UASB the membrane is in contact only with the supernatant at <1 g/L, and start-up is faster for the CSTR design.

2.3.2. Extractive and Diffusive MBRs

Extractive and diffusive MBR processes (EMBRs and DMBRs) are still largely at the developmental stage and are likely to be viable only for niche, high-added value applications. In an extractive system, specific problem contaminants are extracted from the bulk liquid across a membrane of appropriate perm-selectivity. The contaminant then undergoes biotreatment on the permeate side of the membrane, normally by a biofilm formed on the membrane surface. In the case of the diffusive MBRs, a gas permeable membrane is used to introduce into the bioreactor a gas in the molecular, or 'bubbleless' (Côté, Bersillon, & Fau, 1988; Ahmed & Semmens, 1992a,b), form. This again normally feeds a biofilm at the membrane surface. Hence, both extractive and diffusive systems essentially rely on a membrane both for enhanced mass transport and as a substrate for a biofilm, and also operate by diffusive transport: the pollutant or gas for the

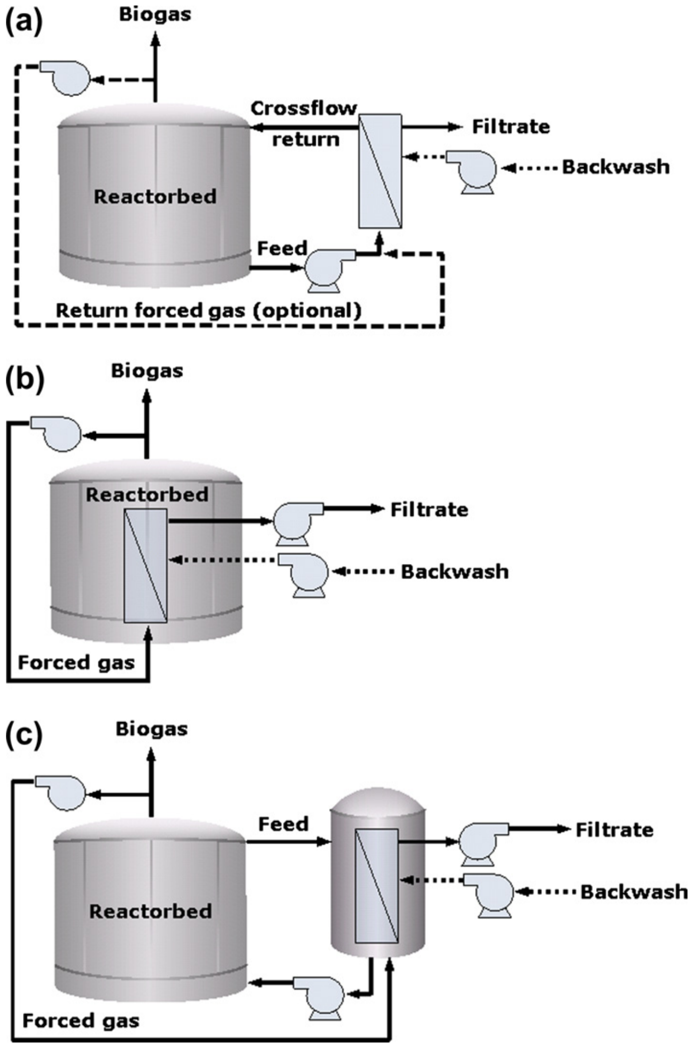

FIG. 2.24 Membrane flow arrangement in an anMBR: (a) pressurized external crossflow, (b) immersed vacuum driven and (c) sidestream vacuum driven *(from Papukchiev, 2009)*.

extractive or diffusive MBR, respectively travels through the membrane under a concentration gradient.

In principle, any gas can be transported across the membrane, though obviously the choices are limited if it is to be used to feed a biofilm. Diffusive systems are generally based on the transfer of oxygen across a microporous membrane and are thus commonly referred to as Membrane Aeration Bioreactors or MABRs (Brindle, Stephenson, & Semmens, 1999). They present an attractive option for very high organic loading rates (OLRs) when oxygen is likely to be limiting, whilst retaining the advantages of a fixed film process (i.e. no requirement for downstream sedimentation and high OLRs). MABRs present an alternative to more classical high-gas transfer processes for oxygenation using pure oxygen such as a Venturi device. However, whereas these devices provide high levels of oxygenation (i.e. high OTRs, Equation (2.26)), it is not necessarily the case that they also provide high levels of utilization by the biomass (oxygen utilization efficiency, OUE). MABRs, on the other hand, have been shown to provide 100% OUEs (Ahmed & Semmens, 1992b; Pankhania, Brindle, & Stephenson, 1999), organic removal rates of 0.002–0.005 kg $m^{-2}\,d^{-1}$ from analogue effluents (Suzuki, Miyahara, & Tokeishi, 1993; Yamagiwa & Ohkawa, 1994; Brindle, Stephenson, & Semmens, 1998) and OLRs of almost 10 kg $m^{-3}\,d^{-1}$ (Pankhania, Stephenson, & Semmens, 1994) – around five times that of conventional MBRs. This means much less membrane area is required to achieve organic removal, but removal efficiencies also tend to be lower.

Recent reviews of diffusive MBRs (sometimes termed 'membrane biofilm reactors' or MBfRs) have demonstrated continued interest in this configuration for ammonia removal (Hwang et al., 2008) and biotreatment generally (Syron & Casey, 2008). Whilst the efficiencies offered and the intensivity of the process make it notionally attractive, it is generally acknowledged that its sustainability is very much constrained by the requirements to control biofilm formation to prevent the reactor from becoming clogged (Pankhania Stephenson, & Semmens, 1994; Celmer, Oleszkiewicz, Cicek, & Husain, 2006). The energy input demanded by the control strategies in this regard detract from the process efficacy to some extent, but the MBfR configuration nonetheless appears to hold some promise for simultaneous nitrification and denitrification by providing nitrification at the membrane/biofilm surface and dentrification in the bulk (Downing & Nerenberg, 2007).

Extractive MBRs allow the biodegradable contaminant to be treated ex situ. This becomes advantageous when the wastewater requiring biotreatment is particularly onerous to micro-organisms which might otherwise be capable of degrading the organic materials of concern. Examples include certain industrial effluents having high concentrations of inorganic material, high acidity or alkalinity, or high levels of toxic materials. Extraction of priority pollutants specifically using a perm-selective membrane, such as a silicone rubber membrane used to extract selectively chlorinated aromatic compounds from effluents of low pH or of high ionic strength (Livingston, 1993, 1994), allows them to be treated under more benign conditions than those prevailing in situ.

Whilst the diffusive and extractive configurations offer specific advantages over biomass separation MBRs, they are also subject to a major disadvantage. Neither process presents a barrier between the treated and untreated stream, although this is not necessarily the case for the extractive system (Section 2.3.3). This means that little or no rejection of micro-organisms takes place and, in the case of diffusive systems, there is a risk of sloughing off of biomass into the product stream in the same way as in the case of a trickling filter.

More recently, there has been interest in extractive and diffusive MBR systems for the specific applications of:

(a) nitrate removal and
(b) combined biotreatment and desalination.

Biological nitrate removal (i.e. denitrification) is achievable through any one of the three process configurations (Section 2.3.3). Most recently, there has been interest shown in the use of extractive MBR hybrid processes, specifically the combination of membrane distillation and forward osmosis (FO), for desalination.

2.3.2.1. Membrane Distillation and Forward Osmosis Hybrid MBRs

The concept of combining an MBR with a membrane desalination process (Table 2.1) has arisen as an alternative to the two-stage MBR-RO process for water reuse, which has been implemented in a number of sites worldwide. For both the MD-MBR and FO-MBR processes, and indeed the NF-MBR process in which the conventional UF/MF membrane is replaced with a nanofilter, the selectivity of the membrane means that minerals are retained in the bioreactor leading to increased mixed liquor salinity. A review of the impacts of elevated salt concentration in an MBR (Lay, Liu, & Fane, 2010) concluded that:

(a) Detrimental physicochemical aspects included reduced oxygen transfer, increased scalant precipitation, higher colloid levels (generated biologically) and increased osmotic back pressure. The latter three all led to a diminution of the membrane permeability and, in the case of colloidal materials, also impaired the product water quality.
(b) Acclimation of the microbial community may be possible up to salt concentrations of 30 g/L with satisfactory biological carbon removal, but above this salinity addition of halophilic or halotolerant micro-organisms is likely to be necessary. Acclimation for nitrification is more challenging and may take significantly longer periods even at the lower salt concentrations of around 10 g/L, though denitrification was considered to be largely unaffected by high salinities, corroborating findings elsewhere (McAdam & Judd, 2008a,c).

The pressure-driven NF process is thus more exposed to these limitations, which exacerbate the extremely onerous fouling conditions promoted by the high organic and solids loading (compared with conventional NF applications, which operate on low-turbidity waters). Permeabilities below 0.5 LMH/bar have been reported for this configuration (Choi, Fukushi, & Yamamoto, 2007), based almost exclusively on cellulose acetate HF membranes, with apparent breakthrough of dissolved organic carbon in some cases. Only in a recent study based on UF PDVF membranes grafted with POEM (poly(oxyethylene methacrylate)) to create thin film composite (TFC) membranes have permeabilities exceeding 5 LMH/bar been reported (Asatekin et al., 2006), for periods of up to 16 h operating in dead-end mode against activated sludge containing 1750 mg/L VSS. Rejection analysis using model organic compounds demonstrated the TFC membrane to exceed 92% retention compared to <69% for the base PVDF UF membrane. However, NF has limited capability for salinity removal, being more effective against multivalent ions.

MD and, in particular, FO processes are less constrained by fouling because they operate with lower TMPs. MD achieves separation through a difference in vapour pressure across the membrane, achieved through a combination of applying a vacuum on the permeate side and increasing the retentate temperature, normally within the range 30−80 °C. The feed thus requires preheating before entering the bioreactor and the permeate water is cooled and recirculated through the module (Fig. 2.25). Permeate fluxes of 2−15 LMH are potentially achievable depending on the operating temperature (Phattaranawik, Fane, Pasquier, & Wu, 2008), with 2−5 LMH reported for operation at 55 °C using 0.22 μm hydrophobic PVDF capillary tube membranes. The temperature determines both the flux and whether operation is mesophilic or thermophilic. The latter appears more common, and can offer advantages of high COD removal efficiency and low sludge net yield, as well as relatively rapid start-up. On the other hand, difficulties with thermophilic operation can be encountered with respect to community proliferation, and elevated temperatures also increase the carbonate scaling propensity associated with the rejected hardness ions (Lay, Liu, & Fane, 2010). There is limited information from extended studies of the process, and thus the extent of long-term insidious fouling is

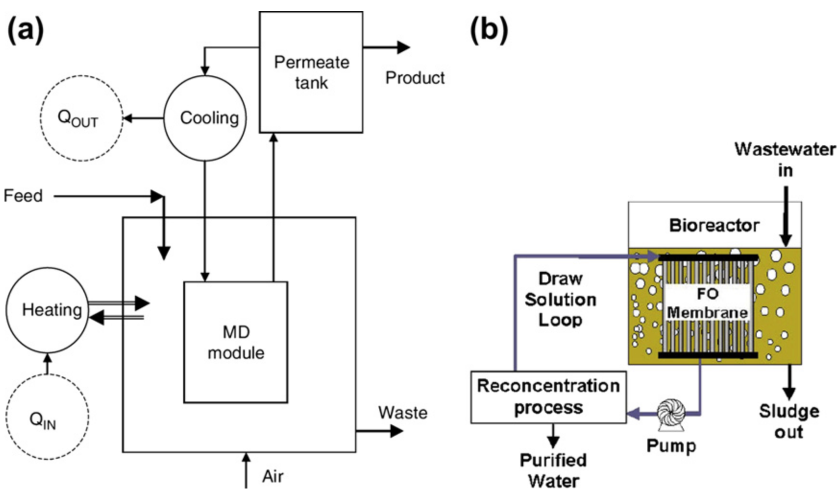

FIG. 2.25 Hybrid process schematics: (a) MD-MBR and (b) FO-MBR *(taken from Phattaranawik et al., 2008; Achilli et al., 2009).*

unknown. The process may also need a source of low-grade waste steam to be viable.

To date, the reported fluxes of the hybrid MBRs have been around or below 10 LMH (Achilli, Cath, Marchand, & Childress, 2009; Choi, Fukushi, & Yamamoto, 2007; Cornelissen et al., 2008; Phattaranawik, Fane, Pasquier, & Wu, 2008). The highest reported fluxes recorded for ambient temperatures − around 9 LMH with almost no observed reversible or irreversible fouling effects (Achilli, Cath, Marchand, & Childress, 2009) − arises for the FO-MBR hybrid technology, and the process typically achieves TOC rejection in excess of 98%. The benign nature of the mass transport of water under zero applied pressure means that the backflushing may be required no more than once a week. Conventional TFC RO and NF membranes have been tested in this hybrid configuration (Cornelissen et al., 2008), as well as dedicated FO membranes (Achilli, Cath, Marchand, & Childress, 2009) which appear to be the most effective since their structure means that they are not prone to internal concentration polarization. Highest fluxes are attained with monovalent draw solutions (e.g. NaCl at 50−70 g/L) and cellulose tri-acetate membranes, but this is accompanied by highest solute diffusion rates of the draw solute into the reactor which exacerbates the build-up of salinity in the reactor and may ultimately demand seeding with halotolerant bacteria (if salt is used as the draw solution) or else operating at low sludge ages to control salinity (Lay, Liu, & Fane, 2010; McAdam & Judd, 2008a,c). Most critically, the viability of the process demands that the recovery of the permeate water from the draw solution is essentially low in energy demand.

2.3.3. Denitrification

The three alternative membrane process modes can all be employed for the removal of nitrate from potable water supplies (McAdam & Judd, 2006). Denitrification, the biochemical reduction of nitrate (Section 2.2.4.4), is conventionally configured as a packed bed process in which denitrification is achieved by a biofilm formed on the packing material. Full-scale schemes for potable duty based on this technology can nonetheless encounter problems of (a) sloughed biomass and (b) residual organic carbon (OC) arising in the treated product.

Biological anoxic denitrification is extensively employed in wastewater treatment, though the configuration employed is more often a suspended growth process. However, various permutations have been trialled for drinking water denitrification (Matějů, Čižinská, Krejčí & Janoch, 1992; Soares, 2000). The motivation for potable reactors is to replace ion exchange since the waste brine generated from this application incurs significant disposal costs (McAdam & Judd, 2008a). To date, full-scale application has been limited, due to poor retention of both the microbial biomass and the electron donor − an exogenous organic substrate. Electron donors trialled have included methanol

(Mansell & Schroeder, 1999), ethanol (Fuchs, Schatzmayr, & Braun, 1997), acetic acid (Barrieros, Rodrigues, Crespo, & Reis, 1998), hydrogen (Haugen, Semmens, & Novak, 2002) and sulphur (Kimura, Nakamura, & Yoshimasa, 2002), all designed to promote the appropriate heterotrophic or autotrophic conditions necessary for denitrification, and each having its own limitations. As already stated, MBRs can be employed to augment denitrification and negate disadvantages traditionally associated with denitrification of potable water (Table 2.5). The three MBR configurations principally under development include the following:

(a) selective extraction of nitrate with porous (Fuchs, Schatzmayr, & Braun, 1997; Mansell, & Schroeder, 1999) or dense (ion exchange) membranes

TABLE 2.5 System Facets of Denitrification MBR Configurations

Configuration	Advantages	Disadvantages
Extractive microporous	Separation of biomass and carbon source from product water	Requires further downstream processing Carbon source breakthrough Pumping costs
Extractive ion exchange	Dense membrane significantly reduces risk of carbon source breakthrough	Requires further downstream processing Potentially complex operation Unknown impact of fouling Comparatively high membrane cost Pumping costs
Diffusive	Non-toxic and low cost electron donor Good nitrate removal Low biomass yield	Requires further downstream processing Biomass breakthrough Potential for fouling to limit mass transfer Health and safety risk with respect to hydrogen gas dissolution Autotrophs, slow to adapt
Biomass rejection	Retention of biomass/active denitrifiers Limited further downstream processing High rate nitrate removal Proven at full scale Appropriate dose control to limit breakthrough Comparatively low cost Comparatively simple to operate	Potential for carbon source breakthrough Limited knowledge of fouling potential

(Velizarov, Rodrigues, Reis, & Crespo, 2003; Matos, Velizarov, Reis, & Crespo, 2008a; Matos, Fortunato, Velizarov, Reis, & Crespo, 2008b);

(b) supply of gas in molecular form (Ho, Tseng, & Chang, 2001; Lee & Rittmann, 2002); or

(c) rejection of biomass (Nuhoglu, Pekdemir, Yildiz, Keskinler, & Akay, 2002; Urbain, Benoit, & Manem, 1996; McAdam, & Judd, 2008b).

2.3.3.1. Extractive Microporous MBR

In this configuration (Fig. 2.26a), also known as a 'confined cell' or 'fixed membrane biofilm reactor', nitrate is extracted from the pumped raw water by molecular diffusion through a physical barrier to a recirculating solution containing the denitrifying biomass. Pressure should ideally be equalized to reduce the influence of diffusion (Mansell & Schroeder, 2002). Various materials have been researched to effectively separate the solutions, including calcium alginate gel, polyacrylamide/alginate copolymer, an agar/microporous membrane

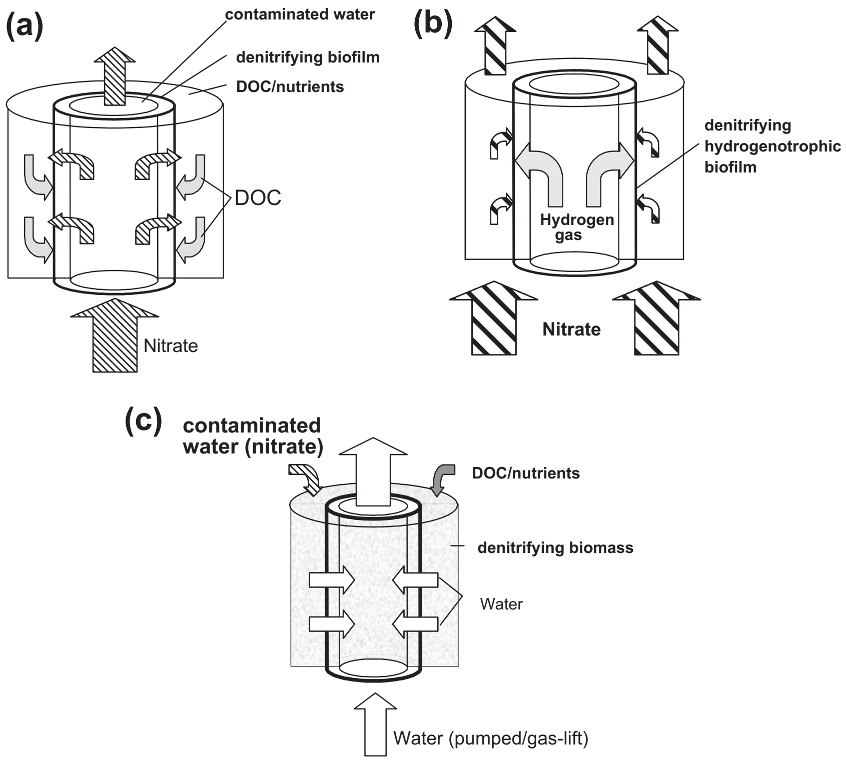

FIG. 2.26 System configurations, denitrifying MBR: (a) nitrate extraction (eMBR), (b) membrane hydrogenation (MHBR) and (c) (conventional) biomass rejection (rMBR).

composite structure and various microporous membranes (Mansell & Schroeder, 2002). Membrane configurations have typically been either FS (Reising & Schroeder, 1996) or MT (Ergas & Rheinheimer, 2004) modules, though a recent study employing UF polymeric HF membranes, more synonymous with classical pressure-driven MBRs, has been reported (Fabricino & Petta, 2007). The advantage of this process is that both the electron donor and the heterotrophic denitrifying biomass are separated from the product water. Whilst the membrane can permit electron donor transport, biofilm formation should theoretically aid donor retention (Fuchs, Schatzmayr, & Braun, 1997).

Between 90% and 99% removal of nitrate has been reported at nitrate levels as high as 200 mg NO_3-N/L (Ergas & Rheinheimer, 2004; Fuchs, Schatzmayr & Braun, 1997; Mansell & Schroeder, 1999). The main limitation of this system appears to be permeation of the electron donor (such as methanol) into the product water, with 8% transfer and 4 mg total organic carbon (TOC)/L product water concentration being, respectively, reported by Ergas and Rheinheimer (2004) and Mansell and Schroeder (1999) in controlled addition experiments. It has been suggested that this problem can be ameliorated by continuous, rather than batch, operation and appropriate control of biofilm growth (Reising & Schroeder, 1996), a postulate corroborated to some extent by studies by Fuchs, Schatzmayr & Braun (1997). However, Fabricino and Petta (2007) adopted a commercial Zeeweed ZW 500 module for concept demonstration at influent flows of up to 0.15 m^3/h (lumen side flow). The authors observed several practical limitations including inorganic precipitation within the fibre lumen, water losses into the biological compartment and TOC contamination of the product water imposed by the positive pressure induced by the biological compartment. Problems of organic carbon breakthrough into the product water can be obviated by using hydrogen as the electron donor, coupled with a bicarbonate carbon source (Mansell & Schroeder, 2002). However, the process then becomes limited by dissolution of hydrogen.

2.3.3.2. Extractive Ion-exchange MBR

This configuration (IEMBR) is identical to the extractive process except that the microporous membrane is replaced by an ion-exchange (IEX) membrane. This membrane is then, in principle, more selective for nitrate which is then removed under a concentration gradient. By appropriate membrane selection, the electron donor concentration in the product water can be reduced to below 1 mg/L (Fonseca, Crespo, Almeida, & Reis, 2000; Velizarov, Rodrigues, Reis, & Crespo, 2000) coupled with 85% nitrate removal at a feed concentration of 135–350 mgNO_3/L (Fonseca et al., 2000). However, Matos et al. (2008b) identified that the membrane constituted >73% of the mass transfer resistance, and thus the IEX membrane provides lower nitrate removal rates than that of extractive microporous membrane systems. As with the extractive technology, the use of the membrane simply to extract nitrate implies that further processing of the product water is required. Moreover, the potential for membrane fouling from both organic materials and hardness (Oldani, Killer, Miquel, & Schock, 1992) has yet to be explored. The relative expense of the ion-exchange membrane (Crespo, Velizarov, & Reis, 2004) has been recognized as a constraint to further development. Consequently, numerous commercially available ion exchange materials to support the extraction and biodegradation of NO_3^-, perchlorate (ClO_4^-), bromate (BrO_3^-) and mercury (Hg^+) within the IEMBR have been evaluated (Velizarov et al., 2008; Matos et al., 2008a). Critically, low-cost membranes exhibited lower preferential selectivity towards the target monovalent anions, and allowed permeation of divalent species. However, complete retention of the exogenous organic carbon was achieved indicating the potential viability of lower cost alternatives (Matos et al., 2008b).

2.3.3.3. Diffusive MBRs

As already stated, the use of hydrogen (H_2) as the electron donor combined with either carbon dioxide or bicarbonate as the carbon source (Mansell & Schroeder, 2002) obviates the organic carbon contamination issue since molecular hydrogen, as a non-polar slightly soluble gas, does not remain dissolved in the water. Such autotrophic (or 'hydrogenotrophic') denitrification can be considered both inexpensive and non-toxic (Haugen, Semmens, & Novak, 2002), as well as producing a relatively low biomass yield (Lee & Rittmann, 2002). As with MABRs, diffusive H_2 MBRs (or membrane biofilm reactor, MBfRs) typically employ microporous HF membranes or dense polydimethyl siloxane (PDMS) tubes (Ho, Tseng, & Chang, 2001) to deliver gas directly to biomass (in this case a denitrifying biofilm) attached to the shell-side of the membrane (Fig. 2.26b) providing up to 100% gas transfer (Mo, Oleszkiewicz, Cicek, & Rezania, 2005). Although around 40% slower than heterotrophic denitrification according to some batch measurements (Ergas & Reuss, 2001), high nitrate removal rates have nonetheless been reported in hydrogenotrophic MBRs (Haugen et al., 2002; Ho et al., 2001). They have also been demonstrated during in-situ (groundwater well) and ex situ MBfR experiments on simulated waters with DO concentrations approaching saturation (Schnobrich, Chaplin, Semmens, & Novak, 2007; Ziv-El & Rittmann, 2009). High nitrate removal rates are attained through high H_2 gas mass transfer rates sustained by limiting fouling. Fouling is mainly manifested as thick and dense biofilms (Roggy, Novak, Hozalski, Clapp, & Semmens, 2002) and, possibly, scalants (Ergas & Reuss, 2001; Lee & Rittmann, 2002) at the membrane surface. Celmer, Oleszkiwicz, and Cicek (2008) demonstrated that high shear forces introduced by nitrogen gas sparging could improve specific denitrification rates by reducing biofilm thickness; specific denitrification rates as high as 0.93 $gNO_3^-N/(m^2\,d)$ were obtained when the biofilm thickness was $<500\ \mu m$. Inorganic precipitation at the membrane surface is strongly correlated to specific denitrification rate (Hwang, Cicek, & Oleszkiewicz, 2009a). However, the impact of mineral precipitation does not appear to adversely affect H_2 transfer in all studies (Lee & Rittmann, 2003; Roggy et al., 2002).

Rezania, Cicek, and Oleszkiewicz (2006), identified β-3 calcium phosphate ($\beta Ca_3(PO_4)_2$) as the predominant foulant on a microporous polypropylene membrane for H_2 gas delivery, and demonstrated precipitation within the pores and lumen of the hydrophobic HF module due to co-transport of precipitates in water vapour. The authors recommend composite or dense materials for future gas diffuser studies.

The disadvantage of gas transfer MBRs is, as with the extractive processes, that the membrane is not used for filtration and also does not retain the biomass. As such, product water is prone to 'sloughed' biomass and other organic matter in the same way as any fixed film biological process. An increase in the effluent total suspended solids in the range of 2−26 mg/L has been reported where shear has been introduced to control biofilm thickness for optimization of specific denitrification capacity (Hwang, Cicek, & Oleszkiewicz, 2010). Also, regulation of the gas flux through the membrane due to the partial pressure drop along the membrane fibre length (Ahmed & Semmens, 1992a,b) can produce uneven biofilm growth; performance can therefore potentially be unpredictable and will also increase the likelihood of H_2 leakage to the bulk (Terada, Kaku, Matsumoto, & Tsuneda, 2006). In addition, doubts about the safety of dosing water with hydrogen remain, notwithstanding the apparent near quantitative retention of hydrogen by the biomass. Finally, the poor adaptability of autotrophic bacteria under drinking water denitrification conditions demonstrated in several studies by long acclimatization periods of 40 and 70 days (Ergas & Reuss, 2001; Ho et al., 2001) demands process modifications such as that demonstrated by Terada et al. (2006) in which a specific denitrification activity of 4.35 gN/(m^3 d) was attained after only 10 days operation by incorporating a fibrous composite matrix at the gas membrane surface.

2.3.3.4. Biomass Rejection MBR

In the conventional configuration the membrane is actually used to filter the water, and both nitrate and the electron donor enter the developed biofilm in the same direction (Fig. 2.26c). Both heterotrophic and autotrophic systems have been investigated using acetate (Barrieros et al., 1998), ethanol (Chang, Manem, & Beaubien, 1993; Delanghe, Nakamura, Myoga, Magara, & Guibal, 1994; Urbain, Benoit, & Manem, 1996; McAdam & Judd, 2007) and elemental sulphur (Kimura, Nakamura, & Yoshimura, 2002) as electron donors. Early studies were based on externally configured MBR to avoid the use of air for membrane scour (Chang, Manem, & Beaubien, 1993; Urbain et al., 1996; Barrieros et al., 1998). More recent drinking water denitrification studies focus on HF iMBR using inert gases or recirculated gases for membrane sparging (McAdam, Judd, Cartmell and Jefferson, 2007; Rezania, Loeszkiewicz, & Cicek, 2007). Kimura et al. (2002), on the other hand, used rotating disc modules to impart shear by centrifugal force, and sustained a flux of 21 LMH for 100 days with limited fouling. Fouling in denitrification sMBRs has not been characterized, though available data would suggest that the biomass has a higher fouling propensity than that generated from sewage treatment (Delanghe et al., 1994; Urbain et al., 1996). Urbain et al. (1996) identified that fouling could be controlled to some extent by reducing the flux to below 50 LMH and operating at crossflows of 2 m/s in sMBRs. Greater characterization of the organics within the iMBR demonstrated that the choice of exogenous carbon substrate was important in minimizing fouling; ethanol supported growth of strong flocs and limited production of primary particles (cf. acetic acid) which suppressed fouling (McAdam, Judd, Cartmell, & Jefferson, 2007). In addition, the low MLSS concentration developed during potable denitrification with an iMBR has permitted limited gas scouring and near dead-end conditions (McAdam & Judd, 2008b); a specific gas demand (SGD_m) of 0.019 $Nm^3/(m^2\,h)$ was sufficient to sustain low fouling conditions to ~21 times below that demanded by constant gas sparging. Under these conditions, air could be used as the sparge gas in place of nitrogen to simplify process design (McAdam, Eusebi, & Judd, 2010b). Nitrate removal efficiencies of up to 98.5% have generally been reported in these studies although, as with previous configurations, investigators have reported organic carbon (Delanghe et al., 1994) and elevated assimilable organic carbon (AOC) concentrations (Kimura et al., 2002) in the product water.

A full-scale 400 m^3/day (0.4 megalitres per day (MLD)) nitrate removal MBR process was constructed in Douchy, France, incorporating powdered activated carbon (PAC) dosing for pesticide removal. Stabilized fluxes between 60 and 70 LMH were obtained at full scale and, contrary to previous investigations, having optimized C:N dosing, treated water of low organic carbon concentration as well as tri-halo methane formation potential (THMFP) was reported. The authors hypothesized that the low effluent organic content was a consequence of effective membrane rejection of biomass by-products of high-molecular-weight organic matter (Urbain et al., 1996). An alternative perspective is to employ denitrification MBR for IEX brine treatment rather than to replace IEX. This has two advantages: (i) limiting process scale and (ii) removing the denitrification MBR from direct contact with the product water, thus removing the risk of organic contamination (McAdam & Judd, 2008c). Recent MBR brine studies have demonstrated high nitrate removal efficiency >90% and reasonable fluxes (~12 LMH) at saline concentrations >100 gNaCl/L with limited impact on permeability from system perturbations (Cyplik, Grajek, Marecik, Króliczak, & Dembczyński, 2007; McAdam, Pawlett, & Judd, 2010a).

2.3.3.5. Hybrid MBR Systems

A system ingeniously using electrolysis to generate hydrogen and feed a biofilm on a granular activated carbon (GAC) support, coupled with a downstream membrane to filter the water, has been trialled (Prosnansky, Sakakibara, & Kuroda, 2002). This system provided treated water nitrate levels of 5–10 mg NO_3-N/L once optimized by employing high-specific area GAC. The low nitrate removal rates were a consequence of influent DO concentration affecting hydrogen dissolution, difficulties with pH control, hydrodynamic limitations and the influence of the anode on nitrate migration on increasing electric field intensity. Furthermore, the process is somewhat limited in application due to an intensive energy requirement and, once again, formation of hydrogen bubbles which impose a safety risk. In a further study, a similar process has been established within the MBfR by replacing H_2 in the lumen with methane (Modin, Fukushi, Nakajima, & Yamamoto, 2008). Methane is supplied simultaneously with oxygen to the biofilm comprised of methanotrophs which synthesize organic compounds to be utilized as an electron donor for denitrifiers sited at the outer extremities of the biofilm. However, current nitrate removal performance, AOC leakage and proximity to a suitable gas source limit the scope for potable applications.

Recent research by Mo, Oleszkiewicz, Cicek, & Rezania. (2005) has focused on incorporating both gas transfer and immersed pressure-driven membranes into the same reactor. The authors focused treatment on suspended biomass rather than biofilms to minimize mass transfer problems previously reported with biofilm development (Ergas & Reuss, 2001; Crespo, Velizarov, & Reis, 2004). Nitrate loading rates between 24 and 192 mg NO_3-N L/day were trialled, with all but the higher loadings resulting in 100% removal performance. However, average effluent dissolved organic carbon (DOC) concentrations of approximately 8 mg/L were also recorded, possibly due to the regular mechanical removal of biofilm from the membrane surface which would otherwise act to reject organic matter.

2.3.3.6. Synopsis

The use of MBRs for drinking water denitrification is still at the developmental stage, with research expanding to target more oxyanions (Rittmann, 2006) in water and wastewater applications and synergies between configurations, for example, to achieve TN consents (Hwang, Cicek, & Oleszkiewicz, 2010), being actively sought. Three different MBR configurations have been studied and whilst challenges for all three configurations remain, specifically organic carbon (exogenous or endogenous) contamination of the treated water, there is now evidence to suggest exogenous carbon in heterotrophic systems can be controlled in biomass rejection MBR (McAdam et al., 2007). If the membrane is not used for direct filtration, as is the case for diffusive and extractive denitrification MBRs, further downstream processing is required for colour, taste and turbidity improvements and for disinfection. Additionally, there are health and safety concerns to address with the hydrogenotrophic systems. The conventional biomass rejection configuration may thus hold the most promise since, as a barrier system, disinfection is also achieved. Based on early economic evaluation (McAdam & Judd, 2008c) and the lower risk to product water quality, implementation of denitrification MBRs for the treatment of IEX brine (Van Ginkel et al., 2008; McAdam et al., 2010a) as a sidestream process appears immediately more feasible (Cyplik et al., 2007; McAdam et al., 2010a) than the direct replacement of ion exchange.

2.3.4. Elements of an Immersed Biomass-rejection MBR

MBRs using immersed membranes to reject biomass represent the most widely employed of all MBR configurations, since they incur the lowest specific energy demand and therefore become the most economically viable for large-scale applications. There are essentially five main elements of the iMBR process key to its design and operation (Fig. 2.27). These are the following:

1. the membrane, its design and the sustaining of permeability;
2. feedwater, its characteristics and its pre-treatment;
3. aeration of both membrane and the bulk biomass;
4. sludge withdrawal and residence time; and
5. bioactivity and nature of the biomass.

These elements are obviously largely inter-related (Fig. 2.28), in particular the latter three which relate to operation. The rate at which sludge is withdrawn controls the residence time (i.e. the SRT) which then determines the concentration of biomass (or, strictly speaking, the mixed liquor suspended solids). The MLSS concentration then impacts both upon the biological properties, i.e. the bioactivity and microbial speciation (Section 2.2.3), and on the physical properties such as the viscosity and oxygen transfer rate (Section 2.2.5). The feedwater chemistry provides the biggest impact upon MBR operation, in that the membrane fouling propensity of the mixed liquor is generally mainly dictated by the nature of the feedwater from which it is generated. Similarly, the rigour of the pre-treatment of the feedwater by screening has a significant impact on the clogging propensity.

Whilst governing principles and the nature of inter-relationships can be appreciated (Fig. 2.28), actual operating conditions and the associated absolute operating parameter values can generally only be arrived at heuristically. Having said this, an understanding of the fundamentals of MBR design, operation and maintenance can proceed through a comprehensive examination of the biological, chemical and physical phenomena occurring in MBRs, since these interact to generate fouling through a number of mechanisms. In the following sections, the elements of the iMBR are considered in turn, namely the membrane itself (Section 2.3.5), the feedwater and biomass characteristics (Section 2.3.6) and the operation and maintenance aspects (Section 2.3.7), with

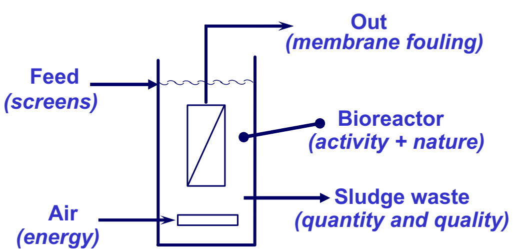

FIG. 2.27 Elements of an iMBR.

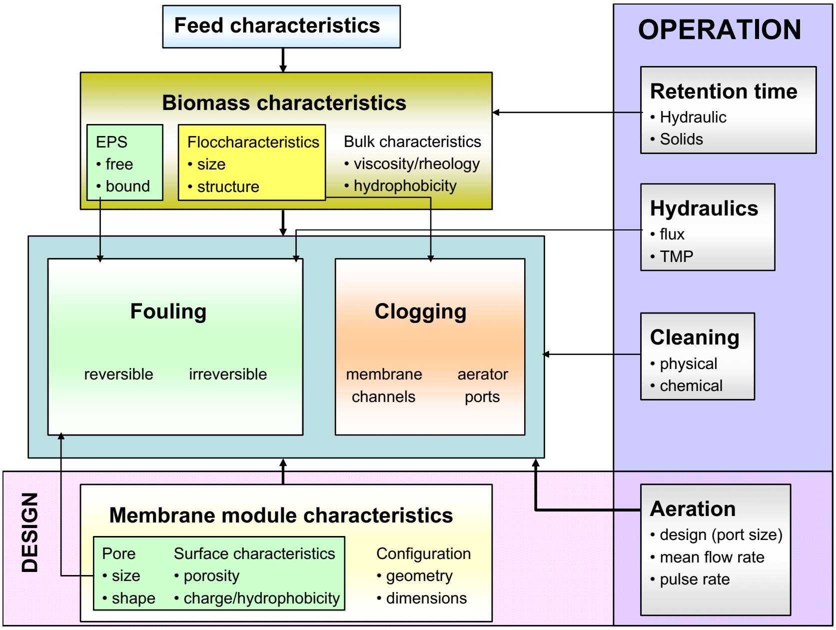

FIG. 2.28 Inter-relationships between MBR parameters and fouling.

a view to appraising mechanisms of fouling (Section 2.3.8) and, ultimately, developing methods for its control (Section 2.3.9). Performance with respect to permeate water quality is appraised in Section 2.3.10.

2.3.5. Membrane Characteristics

2.3.5.1. Physical Parameters

Pore Size

The effects of pore size on membrane fouling are strongly related to the feed solution characteristics and, in particular, the particle size distribution (Section 2.3.6.1). This has led to conflicting trends reported in the literature (Table 2.6), with no consistent general trend noted between pore size and hydraulic performance. This can, in part, be attributed to the complex and changing nature of the biological suspension in MBR systems and the comparatively large pore size distribution of the membranes used (Chang, Le Clech, Jefferson, & Judd, 2002a; Le-Clech, Jefferson, & Judd, 2003c), along with operational facets such as the system hydrodynamics and the duration of the test. A direct comparison of MF and UF membranes at a CFV of 0.1 m/s has shown an MF membrane to provide a hydraulic resistance of around twice that of a UF membrane (Choi et al., 2005b). Interestingly, the DOC rejection of both membranes was similar following 2 h of operation, indicating the dynamic

TABLE 2.6 Effect of Pore Size on MBR Hydraulic Performances

Membranes Tested	Optimum	Test Duration	Other	Ref.
0.1, 0.22, 0.45 μm	0.22 μm	20 h	–	Zhang et al. (2006a)
20, 30, 50, 70 kDa	70 kDa	110 min	Concentrated feed, anaerobic	He et al. (2005)
	50 kDa	110 days		
70 kDa, 0.3 μm	70 kDa	8 h	–	Choi et al. (2005a)
30 kDa, 0.3 μm	30 kDa, 0.3 μm	2 h	CFV = 0.1 m/s CFV = 3.5 m/s	Choi et al. (2005b)
0.1, 0.2, 0.4, 0.8 μm	0.8 μm	n/a	–	Lee et al. (2005)
200 kDa, 0.1, 1 μm	1 μm	3 h	Flux-step test	Le-Clech et al. (2003c)
0.3, 1.5, 3, 5 μm	5 μm	25 min	–	Chang et al. (2001b)
	0.3 μm	45 days		
0.4, 5 μm	0.4 μm	1 day	–	Gander et al. (2000)
	No effect	From 50 days		
0.01, 0.2, 1 μm	No effect	A few hours	Flux-step test	Madaeni et al. (1999)
200 kDa, 0.1, 1 μm	0.1 μm	NA	Anaerobic	Choo and Lee (1996a)
0.05, 0.4 μm	0.05 μm	NA	–	Chang et al. (1994)

membrane layer formed on the membranes to have provided the perm-selectivity rather than the membrane substrate itself.

Conventional wisdom considers smaller pores to afford greater protection of the membrane by rejecting a wider range of materials, with reference to their size, thus increasing cake (or fouling layer) resistance. Compared to that formed on membranes having larger pores, the layer is more readily removed and less likely to leave residual pore plugging or surface adsorption. It is the latter and related phenomena which cause irreversible and irrecoverable fouling. However, when testing membranes with pores ranging from 0.4 to 5 μm, Gander, Jefferson, and Judd (2000) observed greater initial fouling for the larger pore-size membranes and significant flux decline when smaller pore-size membranes were used over an extended period of time, though these authors used isotropic membranes without surface hydrophilization.

Characterization of the distribution of MW compounds present in the supernatant of MBRs operated with membranes of four pore sizes (ranging from 0.1 to 0.8 μm) has also been presented (Lee, Jeon, Cho, Chung, & Min, 2005). Although providing a lower fouling rate, the 0.8 μm pore-size MBR nonetheless had a slightly higher supernatant concentration of most of the macromolecules. From these results, it seems unlikely that the small differences in MW distribution could cause the significant variation in fouling rates observed between the four MBR membranes. In another study based on short-term experiments, sub-critical fouling resistance and fouling rate increased linearly with membrane resistance ranging from 0.4 to 3.5×10^9/m, corresponding to membrane pore size from 1 down to 0.01 μm (Le-Clech et al., 2003c). These results suggest that a dynamic layer is created of greater overall resistance for the more porous membranes operating under sub-critical conditions, and support the notion that larger pores decrease deposition onto the membrane at the expense of internal adsorption. Long-term trials have revealed that progressive internal deposition eventually leads to catastrophic increase in resistance (Cho & Fane, 2002; Le-Clech et al., 2003b; Ognier, Wisniewski, & Grasmick, 2002a), as discussed in Section 2.1.4.6. Tests conducted using a very porous support for the formation of a dynamic membrane have yielded reasonable removal efficiencies and permeabilities (Wu, Huang, Wen, & Chen, 2005), and full-scale installations now exist based on such membranes. However, the trade-off between the higher membrane permeability/higher fouling propensity at higher pore sizes compared with lower membrane permeability/more readily recoverable permeability at lower pore sizes means that in practice the range of pore size for most commercial MBR membrane materials is relatively narrow. Early research conducted on the development of (self-forming) dynamic membranes has been summarized by Meng et al. (2009). With very few articles published on that topic in the last few years, it would seem that the concept of dynamic filters has gained little interest, although they may ultimately offer a low-cost alternative to conventional polymeric membranes.

Porosity/pore Size Distribution/roughness

Membrane roughness and porosity have been identified as possible causes of differing fouling behaviour observed when four MF membranes with nominal pore sizes between 0.20 and 0.22 μm are tested in parallel (Fang & Shi, 2005). The track-etched membrane, with its dense structure and small but uniform cylindrical pores, provided the lowest resistance due to its high surface isoporosity whereas the other three membranes were more prone to pore fouling due to their highly porous network. Although all membranes were of similar nominal pore size, the PVDF, mixed cellulose esters (MCE) and PES membranes resulted in relative pore resistance of 2, 11 and 86% of the total hydraulic resistance, respectively. It was suggested that membrane microstructure, material and pore openings all affected MBR fouling significantly (Fang & Shi, 2005). Comparison between two microporous membranes prepared by stretching demonstrated fouling to be influenced by pore aspect ratio (i.e. pore surface length/pore surface width). Whilst both membranes had the same average pore size and pure water flux, reduced fouling was observed with the membrane having higher pore aspect ratio (i.e. more elliptical and less circular) (Kim, Jang, Chio, & Kim, 2004). Surface roughness has been implicated as promoting fouling from studies on an anaerobic MBR system (He, Xu, Li, & Zhang, 2005), although in this study both pore aspect ratio, i.e. more elliptical size, and membrane morphology, were changed simultaneously. This clear relationship between roughness and fouling propensity has presumably compelled MBR suppliers to specify and optimize their method of membrane manufacture. The recent development of microsieve membranes with extremely narrow pore size distribution potentially offers materials of commensurately low fouling propensity (Ning Koh, Wintgens, Melin, & Pronk, 2008; Brans et al., 2006).

Membrane and Process Configuration

As already discussed (Section 2.1.3) the immersed process configuration is generally favoured over the pumped sidestream configuration for medium to large-scale domestic wastewater treatment (Gunder & Krauth, 1999; Fane, Chang, & Chardon, 2002; Judd, 2005; Le-Clech, Jefferson, & Judd, 2005b), although installations based on the air-lift sidestream (a-lsMBR) configuration are increasing in number and size (Section 4.4.1). This relates mainly to the impact of aeration, which suppresses fouling through generating shear (Section 2.3.7.1).

iMBR membranes are largely configured either as HF or FS whereas sMBRs are either FS or, most usually, MT. Whilst HF modules are generally less expensive to manufacture, allow high packing density and tolerate vigorous backflushing, they are also less readily controlled hydrodynamically than FS or MT membranes where the membrane channel dimensions are well defined. It has been recognized for some time that this leads to higher permeabilities for the FS membranes (Gunder & Krauth, 1998), although apparently at the expense of a higher aeration demand.

Packing density, or more specifically the separation between adjacent membranes, has a direct impact on clogging, shear and aeration energy demand. For a given liquid upflow rate, as provided by air-lift, increasing the separation reduces the risk of clogging by gross solids. Reducing the separation will also, for a given bubble volume, retard the rising bubble as a consequence of the gas:membrane contact area increasing, thereby increasing the downward drag force. This might then be expected to decrease the flux because shear forces are reduced as a result. On the other hand, for a given liquid upflow velocity and thus the same shear, increasing channel width also increases the aeration energy demand since a larger volume of liquid (dictated by the channel width) is being passed over the same membrane area; it is the volumetric airflow rate that determines energy demand.

Experiments conducted on a bundle of nine fibres revealed the module performance to be significantly worse than that of a module based on an individual fibre (Yeo & Fane, 2005), with the surrounding fibres being less productive. At high feed concentrations and low CFVs, the surrounding fibres became completely blocked and eventually produced negligible flux. A lowering of the packing density by 30% was advised to allow the bundle to perform similarly to individual fibres, since this reduced the impact of cake layers forming on adjacent fibres and allowed greater shear to be generated (Yeo & Fane, 2005). A mathematical model based on substrate and biomass mass balance also revealed the significant role played by packing density in overall MBR performance and the maintenance of MLSS in particular (Vigneswaran, Shim, Chaudhary, & Ben Aim, 2004).

The effects of other membrane characteristics, including HF orientation, size and flexibility, have been reviewed (Cui, Chang, & Fane, 2003). For HF membranes used for yeast filtration, higher critical fluxes were measured for membranes of smaller diameter (0.65 mm) and greater length (80 cm) (Wicaksana, Fane, & Chen, 2006), though contradictory results showing slightly higher permeabilities for shorter membranes (0.3 cf. 1.0 m) have also been reported (Kim & DiGiano, 2006). A significant impact of membrane length is the pressure drop due to lumen-side permeate flow. Ideally, the resistance to permeate flow from the membrane to the outlet of the element should be small compared with that offered by the membrane itself. If this is not the case, then the hydraulic losses across the permeate side of the membrane element may be sufficient to produce a significant flux distribution across the membrane surface. Significant pressure losses (up to 530 mbar) have been measured for fibres longer than 15 cm; below this critical length, pressure loss was reported as being less than 110 mbar (Kim, Jang, Chio, & Kim, 2004). Further discussion of fouling distribution in HF membranes can be found elsewhere (Lipnizki & Field, 2001; Chang & Fane, 2002; Chang, Fane, & Vigneswaran, 2002b; Zheng, Fan, & Wei, 2003; Zhongwei, Liying, & Runyu, 2003; Ognier, Wisniewski, & Grasmick, 2004).

2.3.5.2. *Chemical Parameters*

Since hydrophobic interactions take place between solutes, microbial cells of the EPS and the membrane material, membrane fouling is expected to be more severe with hydrophobic rather than hydrophilic membranes (Chang, Lee, & Alin, 1999; Madaeni, Fane, & Wiley, 1999; Yu et al., 2005a; Yu, Hu, Xu, Wang, & Wang, 2005b). Due to their physical strength and chemical resistance, PVDF, PP, PE and PES are the main materials used to manufacture commercial MBR membranes, with PAN, PVA (polyvinylalcohol) and PTFE also being used (Sections 2.1.2 and 4.5). Most of these materials are, however, relatively hydrophobic, but some can be mixed with additives during fabrication which then allows the use of these membranes for aqueous filtration. Typical polymers used to increase the wettability of the membrane material include polyvinylpyrrolidone (PVP) and methacrylates (Puspitasari, Granville, Le-Clech, & Chen, 2010). As a result, many of the commercially available membranes comprise a specific and original blend of additives, and the comparison of the filtration performance of membranes based on their core material alone becomes less relevant.

In the literature, changes in membrane hydrophobicity are often linked with other membrane modifications such as pore size and morphology, which make the correlation between membrane hydrophobicity and fouling more difficult to assess. The effect of membrane hydrophobicity in an aerobic MBR, from a comparison of two UF membranes of otherwise similar characteristics, revealed greater solute rejection and fouling and higher cake resistance for the hydrophobic membrane (Chang, Bag, & Lee, 2001a). It was concluded that the solute rejection was mainly due to the adsorption onto or sieving by the cake deposited on the membrane, and, to a lesser extent, direct adsorption into membrane pores and at the membrane surface. It has also been suggested (Fang & Shi, 2005) that membranes of greater hydrophilicity are more vulnerable to deposition of foulants of hydrophilic nature, though in this study the most hydrophilic membrane was also the most porous and this can also enhance fouling (Section 2.3.5.1).

Although providing superior chemical, thermal and hydraulic resistance, the use of ceramic membranes in MBR technologies is limited by their high cost to niche applications such as treatment of high-strength industrial waste (Scott, Neilson, Liu, & Boon, 1998; Luonsi et al., 2002) and anaerobic biodegradation (Fan, Urbain, Quian, & Manem, 1996) in sMBRs. A direct comparison of 0.1 μm ceramic and 0.03 μm polymeric multi-channel membrane modules operated in sidestream air-lift mode showed the former to operate without fouling up to at least 60 LMH, the highest flux tested, whereas for the latter criticality was indicated at ~36 LMH (Judd, Alvarez-Vasquez, & Jefferson, 2006). Novel stainless steel membrane modules have been shown to provide good hydraulic performance and fouling recovery when used in an anaerobic MBR (Zhang et al., 2006d), though evidence suggests that ceramic membranes may be more prone to scaling than polymeric ones.

Since fouling is expected to be more severe at higher hydrophobicities, efforts have been focused on increasing membrane hydrophilicity by chemical surface modification. Examples of MBR membrane modification include NH_3 and CO_2 plasma treatment of PP HFs (Yu et al., 2005a) to functionalize the surface with polar groups. In both cases, membrane hydrophilicity significantly increased and the new membranes yielded better filtration performance and flux recovery than those of unmodified membranes. The same group also applied (with more success but at higher production cost) graft polymerization on PP membranes with UV irradiation in acrylamide solutions (Yu, Xu, Lei, Hu, &

Yang, 2007). In another study, addition of TiO_2 nanoparticles to the casting solution and direct pre-filtration of TiO_2 allowed the preparation of two types of TiO_2-immobilized UF membrane, respectively, comprising entrapped and deposited particles, which were used in MBR systems (Bae & Tak, 2005). A lower flux decline was reported for the TiO_2-containing membranes compared to the unmodified materials, the surface-coated material providing the greatest fouling mitigation. When MBR membranes were pre-coated with ferric hydroxide flocs and compared to an unmodified MBR, both effluent quality and productivity were found to increase (Zhang, Bu, Liu, Luo, & Gu, 2004). Other examples of membrane modification include PVA dip-coating on non-woven fibre (Zhang, Yang, Wang, & Chen, 2008a), and self-assembling graft copolymers (Asatekin, Kang, Elimelech, & Mayes, 2007; Asatekin, Olivetti, & Mayes, 2009).

Whilst membrane modification provides a focus of interest, the development of new membrane modification strategies or new chemical additives is rarely accompanied by an assessment of the long-term MBR performances. In addition, the increase in hydrophilicity offered by these new membranes is usually characterized by contact angle measurement, which fails to characterize the detailed interactions between membrane surface and potential foulants. Not only may the nature of these additives and/or coating materials be rapidly altered during repetitive chemical cleaning, but their exact effect on MBR performance is also questionable when the fouling layer is established and covers the membrane surface (Puspitasari et al., 2010). Moreover, membrane performance with regard to fouling resistance has to be considered alongside membrane life, given that both ultimately contribute to operating costs (Section 3.5.2). Information from other membrane applications suggests that increased chemical cleaning tends to decrease membrane life (Yamamura, Kimura, & Watanabe, 2007), yet there is evidence from the more mature MBR installations that existing commercial products are very robust in this regard (Section 3.6.6). However, little research has been conducted thus far quantifying installed MBR membrane mechanical integrity with age.

2.3.5.3. Anaerobic MBRs

Whilst anMBR fouling mechanisms may be similar to those of aerobic technologies, the nature of the foulants can be expected to be different and, as with the aerobic systems, to change with feedwater characteristics, membrane surface and membrane module properties and process operating conditions. A review by Bérubé, Hall, and Sutton (2006a) has been conducted, and aspects of it summarized below.

Cake layers on membranes in anMBRs have been reported to contain both organic matter and precipitated struvite (Choo & Lee, 1996a), though fouling of organic membranes appears to be governed by biological/organic interactions with the membrane rather than by struvite formation. Choo et al. (2000) observed no difference in fouling rate when ammonia, a component of struvite, was removed from the mixed liquor prior to filtration using an organic membrane. It also appears that internal fouling (i.e. membrane pore plugging) by soluble and colloidal material is less significant than cake layer fouling (Choo & Lee, 1996a; Lee, Jung, & Chung, 2001c; Kang, Yoon, & Lee, 2002;) on organic membranes. However, for ceramic membranes, which provide higher fluxes and for which cake layers are much thinner (especially at high crossflows), the bulk of the fouling has been attributed to internal fouling by struvite. This has been concluded from scanning electron micrograph (SEM) studies coupled with a magnesium mass balance (Yoon, Kang, & Lee, 1999), and from observed impacts of ammonia level (Choo et al., 2000). As with aerobic processes, hydrophobicity suppresses fouling on anMBR polymeric membranes to some extent, and membrane charge may also be important in determining fouling (Kang, Yoon, & Lee, 2002) unless the ionic strength is high enough to compress the double layer and thus nullify charge repulsion (Fane, Fell, & Suzuki, 1983).

Membrane pore size effects also follow similar patterns to those of aerobic systems in that large pores provide greater initial fluxes but more rapid subsequent flux decay (Saw, Anderson, James, & Le, 1986; Imasaka, Kanekuni, So, & Yoshino, 1989; Choo & Lee, 1996b; He, Li, Wu, & Gu, 1999, He, Xu, Li, & Zhang, 2005), which has been attributed to either internal or surface pore plugging. However, the optimum pore size appears to depend on the liquor characteristics. Elmaleh and Abdelmoumni (1997), investigating the impact of pore size on the anMBR steady-state permeate flux, recorded highest steady-state fluxes at a pore size of 0.45 μm for an anaerobic mixed liquor, compared with 0.15 μm for a mixed microbial population of methanogens.

The long-term impact of UF membrane pore size on hydraulic performance has been assessed by He et al. (2005) for an anaerobic MBR. The lowest MWCO-rated membrane tested (20 kDa) yielded the largest permeability loss within the first 15 min of filtration when compared to 30, 50 and 70 kDa membranes. However, when operated for an extended time (over 100 days) with regular hydraulic and chemical cleaning, the largest MWCO membrane (70 kDa) experienced the greatest fouling rate, as 94% of its original permeability was lost, compared to only a 70% performance decrease for the other three membranes. The 30 and 50 kDa membranes thus provided the best overall hydraulic performance, possibly indicating an optimum membrane pore size for a given application. These results also reveal the impact of test duration. Similar temporal trends where long-term fouling was exacerbated by larger pores have been demonstrated for MF membranes ranging from 1.5 to 5 μm pore size for aerobic MBRs (Chang, Gander, Jefferson, & Judd, 2001b).

Stuckey and Hu (2003) reported that slightly higher permeate fluxes could be maintained for an HF compared with an FS membrane element in an immersed anMBR. This would appear to have been corroborated by a subsequent comparative study of FS and HF membranes of the same pore size (0.4 μm) used for anaerobic treatment (Hai, Yamamoto, & Fukushi, 2005), where the authors found the FS membrane to foul slightly more than the HF membrane and the permeability was not recovered following cleaning with water. However, surface properties have been shown to be pore size related: the contact angle measurement in an anMBR study revealed the apparent hydrophobicity of PES membranes decreased with increasing MWCO (from 20 to 70 kDa) (He et al., 2005).

2.3.6. Feed and Biomass Characteristics

2.3.6.1. Feed Nature and Concentration

Whilst membrane fouling in physical wastewater filtration depends directly on the water quality (Judd & Jefferson, 2003; Fuchs, Braun, & Theiss, 2005; Schrader, Zwijnenburg, & Wessling, 2005), MBR membrane fouling is mostly affected by the interactions between the membrane and biological suspension rather than feedwater (Choi, Zhang, Dionysiou, Oerther, & Sorial, 2005a). More recalcitrant feedwaters, such as landfill leachate (Section 5.4.3), may undergo more limited biochemical transformation such that the membrane is challenged in part by the unmodified feed. Biological transformations taking place are influenced both by the operating conditions and the feedwater quality (Le-Clech et al., 2003b; Jefferson, Brookes, Le Clech, & Judd, 2004). In particular, the level of biodegradability of the carbon source impacts the characteristics of the biomass, and the fouling propensity; the use of a glucose feed has been shown to generate more EPS than an acetate one (Li & Yang, 2007). Also, the feed protein/carbohydrate has been correlated with EPS production (Arabi & Nakhla, 2008), although SRT is more important in this regard (Section 2.3.7.2). Finally, the presence of inorganic matter in the influent capable of forming sparingly soluble salts can also have a significant impact on MBR fouling, including struvite (NH_4MgPO_4) in anMBRs (Kang et al., 2002) and calcium carbonate ($CaCO_3$) generally (Ognier, Wisniewski, & Grasmick, 2002a; Ognier, Wisniewski, & Grasmick, 2002b), along with many other chemical/biochemical compounds (Meng et al., 2009).

2.3.6.2. Biomass Foulants

Two types of foulant study dominate the MBR scientific literature: characterization and identification. Characterization refers to properties (usually relating to membrane permeability) the foulant demonstrates either in situ, that is, within the MBR, or ex situ in some bespoke or standard measurement, such as capillary suction time (CST) or specific resistance to filtration (SRF). Identification refers to physical and/or chemical classification of the foulant, invariably through extraction and isolation prior to chemical analysis. Of course, foulant isolates may also be characterized in the same way as the MBR biomass.

In general, foulants can be defined in three different ways (Table 2.7):

1. practically, based on permeability recovery;
2. mechanistically, based on fouling mechanism; and
3. by material type, based on chemical or physical nature or on origin.

TABLE 2.7 Foulant Definitions

Practical	Mechanism	Foulant Material Type
Reversible/temporary: ● Removed by physical cleaning Irreversible/permanent: ● Removed by chemical cleaning Irrecoverable*/absolute: ● Not removed by any cleaning regime	Pore blocking/filtration models (Fig. 2.10): ● Complete blocking ● Standard blocking ● Intermediate blocking ● Cake filtration	Size: ● Molecular, macro-molecular, colloidal or particulate Surface charge/chemistry: ● Positive or negative (cationic or anionic) Chemical type: ● Inorganic (e.g. scalants) or organic (e.g. humic materials, EPS) ● Carbohydrate or protein (fractions of EPS) Origin: ● Microbial (autochthonous), terrestrial (allochthonous) or man-made (anthropogenic) ● (Extracted) EPS ((e)EPS) or soluble microbial product (SMP)**

**Irrecoverable fouling is long-term and insidious.*
***eEPS refers to microbial products directly associated with the cell wall; SMP refers to microbial products unassociated with the cell (Fig. 2.32).*

Of these, evidence suggests that it is the physical nature, and specifically the size, of the foulant that has the greatest impact on its fouling propensity. Hence, activated sludge biomass can be fractionated into three categories: suspended solids, colloids and solutes. The fractionation methodology critically affects the measurements made. Typically, the biomass sample is centrifuged. The resulting supernatant is then filtered with a dead-end membrane cell, with the calculated hydraulic resistance being attributed to colloidal and soluble matter combined (R_{col} and R_{sol}, respectively, Fig. 2.29). Another portion of the biomass suspension is then microfiltered at a nominal pore size of 0.5 μm and the fouling properties of this supernatant (R_{sol}) attributed solely to the soluble matter. The relative fouling contributions of the suspended and colloidal matter can then be calculated (Bae & Tak, 2005). The resistance provided by colloidal matter has also been attributed to the difference between the levels of TOC present in the filtrate passing through 1.5 μm filtration paper and in the permeate collected from the MBR membrane (0.04 μm) (Fan, Zhou, & Husain, 2006).

Fractionation methods may vary slightly for different studies, but results are often reported in terms of hydraulic resistance for suspended solids, colloids and soluble matter, the sum of which yields the resistance of the activated

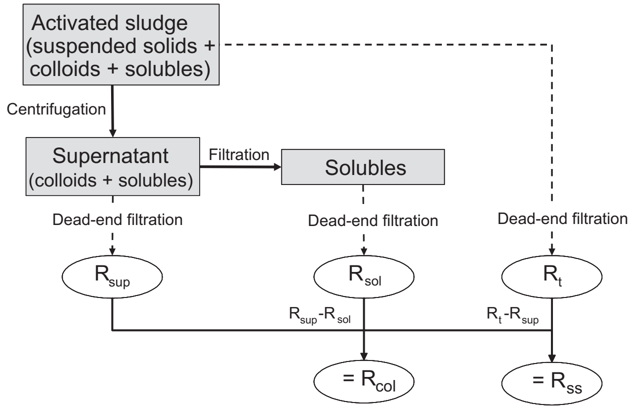

FIG. 2.29 Experimental method for the determination of the relative fouling propensity for the three physical biomass fractions.

sludge. Although an interesting approach to the study of MBR fouling, fractionation neglects coupling or synergistic effects which may occur both among different biomass components and with operating determinants. Such interactions are numerous and include feedwater quality (Li, Gao, Hua, Du, & Chen, 2005b), membrane permeability, particle size and hydrodynamics conditions (Bae & Tak, 2005) (Fig. 2.28). An attempt to compare results obtained from different studies is depicted in Fig. 2.30, where relative fouling resistance contributions have been calculated. This figure clearly reveals the lack of common trends between studies.

The relative contribution of the biomass supernatant to overall fouling ranges from 17% (Bae & Tak, 2005) to 81% (Itonaga, Kimura, & Watanabe, 2004). Such variation is probably attributable to the different operating conditions and biological state of the suspended biomass. It appears from these data that fouling by suspended solids is rather less than that of the supernatant. The latter is generally regarded as comprising soluble microbial product (SMP), which is soluble, and colloidal matter that derives from the biomass (Section 2.3.6.5). With respect to fouling mechanisms, soluble and colloidal materials are assumed to be responsible for membrane pore blockage, whilst suspended solids account mainly for the cake layer resistance (Itonaga, Kimura, & Watanabe, 2004). However, since iMBRs are typically operated at a modest flux, cake formation is limited and deposition of physically smaller species is more likely to take place. Whilst there has been much interest in colloidal materials and their contribution often assessed, it has been argued that their impact on MBR fouling is relatively minor (Teychene, Guigui, Cabassud, & Amy, 2008).

Biofouling can be described as the undesirable deposition of materials of biological origin on a surface (Sommariva, Comite, Capannelli, & Bottino, 2007), and contributes to the reduction of hydraulic performances in MBR

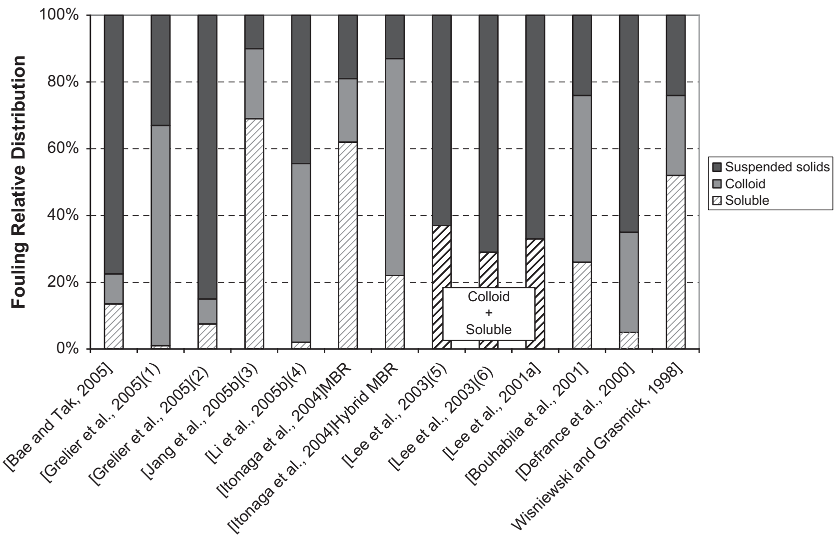

FIG. 2.30 Relative contributions (in %) of the different biomass fractions to MBR fouling: (1−2): For SRT increase from 8 (1) to 40 days (2); (3): *F*/*M* ratio of 0.5, results based on modified fouling index; (4): Based on flux reduction after 600 min of each fraction filtration; and (5−6): For SRT increase from 20 (5) to 60 days (6).

systems. It can be further described as the initial attachment of SMP onto the membrane surface through adhesive forces during either passive adsorption or filtration. During MBR operation, bacteria then attach by cohesive mechanisms to the membrane surface already covered by SMP. As the mixed liquor filters through the fouled membrane, it provides nutrients and DO to the deposited bacteria. As a result, the immobilized bacteria assimilate to the surrounding environment by producing EPS and by forming a complex structure, generally termed the biofilm (Tansel et al., 2006).

Advanced characterisation of the biofilm formed on the membrane surface has formed the basis of much study. Efforts have focused on two main areas: (a) use of state-of-the-art visualization techniques to examine the morphology of the biofouling/biofilm and (b) identification of the microbial community present in the biofilm. The various observation techniques have been recently reviewed, along with the advantages and limitations for their potential use in MBRs (Marselina, Le-Clech, Stuetz, & Chen, 2009); confocal laser scanning microscopy (CLSM) in particular has been shown to offer great potential in characterizing the complex structure of the MBR membrane fouling layer (Yun et al., 2006). In terms of community analysis, methods like fluorescence in situ hybridization (FISH) and polymerase chain reaction denaturing gradient gel electrophoresis (PCR−DGGE) are usually used to study the nature of the microbial population and structure on the membrane surface. From this work, specific bacteria have been identified as having a propensity to deposit and adsorb onto the membrane surface and to initiate biological growth (Zhang, Yang, Wang, & Chen, 2008a). The large number of cultures present makes the unambiguous identification of the microbial communities predominating in the MBR biofilm challenging (Meng et al., 2009). However, of practical importance are the filamentous bacteria, which lead to sludge bulking in clarifiers of CASPs, and promote foaming (Sections 2.3.6.3 and 3.6.5) and EPS production, usually accompanied by TMP increase (Meng & Yang, 2007; Pan, Su, Huang, & Lee, 2010). Strategies proposed to control the development of filamentous bacteria include use of selectors, addition of coagulants, high DO conditions and supplementing alkalinity (Liu and Liu, 2006), though appropriate ameliorative measures rely on the precise identification of the species concerned (Section 3.6.5).

2.3.6.3. Biomass Bulk Parameters

MLSS Concentration

Whilst suspended solids concentration may seem intuitively to provide a reasonable indication of fouling propensity, the relationship between MLSS level and fouling propensity is complex. The impact of increasing MLSS on membrane permeability can be either negative (Chang & Kim, 2005; Cicek et al., 1999), positive (Defrance & Jaffrin, 1999; Le-Clech et al., 2003c) or insignificant (Hong, Bae, Tak, Hong, & Randall, 2002; Lesjean et al., 2005b), as indicated in Table 2.8. The existence of a threshold above which the MLSS concentration has a negative influence has been reported (30 g/L, according to Lubbecke, Vogelpohl, & Dewjanin, 1995). A more detailed fouling trend has been described (Rosenberger, Evenblij, te Poele, Wintgens, & Laabs, 2005), in which an increase in MLSS reduced fouling at low MLSS levels (~6 g/L) whilst exacerbating fouling at MLSS concentrations above 15 g/L. The level of MLSS did not appear to have a significant effect on membrane fouling between 8 and 12 g/L.

Contradictory trends from data obtained in the same study are apparent. For example, the cake resistance (R_c) has been observed to increase and the specific cake resistance (ψ, the resistance per unit cake depth) to decrease with increasing MLSS, indicating that the bulk cake becomes more permeable. Bin, Xiaochang, and Enrang (2004) observed the permeate flux to decrease (albeit at a reduced fouling rate) with increasing MLSS. This was attributed to the rapid formation of a fouling cake layer (potentially protecting the membrane) at high concentration, while progressive pore blocking created by colloids and particles was thought to take place at lower MLSS concentrations when the membrane was less well protected. This may well explain the sub-critical fouling behaviour depicted in Fig. 2.15.

Empirical relationships predicting flux from MLSS level have been proposed in a number of papers (Krauth & Staab, 1993; Sato & Ishii, 1991; Shimizu, Okuno, Uryu, Ohtsubo, & Watanabe, 1996; Fang & Shi, 2005). However, these equations have limited use as they are generally obtained under

TABLE 2.8 Influence of Change in MLSS Concentration (g/L) on MBR Fouling

MLSS change, g/L	Details	Ref.
Fouling increase		
0.09–3.7	Cake resistance: 21–54 × 10^{11} m^{-1} and a: 18.5–0.7 × 10^8 m/kg	Chang and Kim (2005)
2.4–9.6	Total resistance: 9–22 × 10^{11} m^{-1}	Fang and Shi (2005)
7–18	Critical flux: 47–36 LMH (for SRT of 30–100 days)	Han et al. (2005)
2.1–9.6	Critical flux: 13–8 LMH	Bin et al. (2004)
1–10	Critical flux: 75–35 LMH	Madaeni et al. (1999)*
2–15	'Limiting flux': 105–50 LMH	Cicek, Franco, Suidan, and Urbain (1998)*
1.6–22	'Stabilized flux': 65–25 LMH	Beaubien et al. (1996)*
Fouling decrease		
3.5–10	Critical flux: >80, <60 LMH	Defrance and Jaffrin (1999)*
No (or little) effect		
4.4–11.6	No impact from 4 to 8 g/L, Slightly less fouling for 12 g/L	Le-Clech et al. (2003c)
4–15.1	Critical flux decreased from 25 to 22 LMH	Bouhabila et al. (1998)
3.6–8.4		Hong et al. (2002)

**sMBR.*

very specific conditions and are based on a limited number of operating parameters, whilst other parameters are disregarded. A mathematical expression linking MLSS concentration, EPS and TMP with specific cake resistance has been proposed by Cho, Song, and Ahn (2005a). In this study, specific resistance changed little at MLSS levels between 4 and 10 g/L at constant EPS and TMP. The experimental method used for adjusting MLSS concentration can significantly impact on biomass characteristics, since biomass solids levels can be raised by sedimentation coupled with decantation without acclimation, yielding different biological characteristics (Cicek et al., 1999). MLSS concentration also impacts on MBR removal efficiencies; an optimum concentration of ~6 g/L has been identified based on the removal of COD (Ren, Chen, Wang, & Hu, 2005) and phage (Shang, Wong, & Chen, 2005).

The lack of a clear correlation between MLSS concentration and any specific foulant characteristic(s) indicate that MLSS concentration alone is a poor indicator of biomass fouling propensity. Some authors (Brookes et al., 2003b; Jefferson, Brookes, Le Clech, & Judd, 2004) have recommended using fundamental operating parameters such as HRT and SRT as foulant level indicators, supported by relatively stable foulant levels and characteristics under steady-state conditions. Although more recent studies have pointed to non-settlable/colloidal organic substances, rather than MLSS concentration, as being primary indicators of fouling propensity in MBRs (Section 2.3.6.5), the relationship between fouling propensity and some single biomass characteristic or operating condition — or a combination thereof — remains challenging to define.

Viscosity

As with a CASP, biomass viscosity closely relates to solids concentration (Section 2.2.5.3) and contributes to fouling (Yeom, Lee, Choi, Kim, & Lee, 2004). Whilst viscosity has been reported to increase roughly exponentially with MLSS concentration (Manem & Sanderson, 1996; Rosenberger, Kraume, & Szewzyk, 1999), a critical MLSS concentration exists below which viscosity remains low and rises only slowly with the concentration and above which the exponential relationship is observed (Itonaga, Kimura, & Watanabe, 2004). This critical value, generally between 10 and 17 g/L, depending on feedwater quality and process operating conditions, also exists for CST:viscosity correlations (Brookes, Jefferson, Le-Clech, & Judd, 2003a). MLSS viscosity impacts both on flux and air bubble size and can dampen lateral movement of HFs in immersed bundles (Wicaksana, Fane, & Chen, 2006).

Temperature

Temperature impacts on membrane filtration through permeate viscosity. A similar temperature correction can be applied as that used for $k_L a$ (Equation (2.15); Section 2.2.5.3):

$$J = J_{20} 1.025^{(T-20)}, \tag{2.34}$$

where J is the flux at the process temperature T in °C. This correction is not comprehensive, however, as has been demonstrated through normalization of flux data obtained at different temperatures (Jiang, Kennedy, Guinzbourg, Vanrolleghem, & Schippers, 2005). The greater than expected normalized resistance at lower temperatures has been explained by a number of contributing phenomena:

1. impacts of the viscosity of the sludge, rather than that of the permeate, which increases more significantly than permeate viscosity and reduces the shear stress generated by coarse bubbles as a result;
2. intensified deflocculation at low temperatures, reducing floc size (Section 2.3.6.4) and releasing EPS to the solution;

3. particle back transport velocity, which decreases linearly with temperature according to Brownian diffusion; and
4. biodegradation of COD, which decreases with temperature and results in a higher concentration of unbiodegraded solute and particulate COD (Fawehinmi, Lens, Stephenson, Rogalla, & Jefferson, 2004; Jiang et al., 2005).

All of these factors directly impact on membrane fouling and, as such, more extensive deposition of foulant materials on the membrane surface is to be expected at lower temperatures. During the assessment of the effect of seasonal variations on membrane fouling, it has been shown that fouling tends to be more reversible at low temperature, while the relative fraction of irreversible fouling was higher during the summer season (Miyoshi, Tsuyuhara, Ogyu, Kimura, & Watanabe, 2009). In this study, the level of reversibility was also strongly related to the concentration of DOC in the bioreactor.

Higher operating temperatures can be maintained in anMBRs than in aerobic ones (Baek & Pagilla, 2006), and this both decreases viscosity and apparently lowers levels of SMP (Schiener, Nachaiyasit, & Stuckey, 1998; Fawehinmi et al., 2004). Higher temperatures, at least within the mesophilic temperature range, also result in increased COD removal, as demonstrated with a conventional UASB (Singh & Viraraghavan, 2003).

Dissolved Oxygen

The bioreactor DO concentration is controlled by the aeration rate, which provides oxygen to the biomass, and is also used for membrane fouling control. DO impacts on MBR fouling through system biology, e.g. biofilm structure, SMP levels and floc size distribution (Lee et al., 2005). Higher DO levels generally provide better filterability, as manifested in filter cakes of lower specific resistance due to larger particles (Kang, Lee, & Kim, 2003). Against this, some researchers have noted a decrease with DO concentration in mixed liquor dissolved COD levels (Ji & Zhou, 2006). Moreover, the contribution of SMP concentration to membrane filterability was found to be less than that of particle size and porosity (Kang et al., 2003). Ji and Zhou claimed aeration rate to directly control the quantity and composition of SMP, EPS and total polymeric substances in the biological flocs and ultimately the protein/carbohydrate ratio in the membrane surface deposit. In a study of anoxic and aerobic sludge filterability, distinct carbohydrate structures were observed and used to explain the different fouling rates obtained for the two systems. The effect of oxygen limitation causing a lowering of the cell surface hydrophobicity has been reported as another potential cause of MBR fouling (Jang, Ren, Choi, & Kim, 2005a). The level of DO in the bioreactor has also been related to other parameters, specifically molecular weight, of the biopolymers which may then impact on fouling propensity (Min, Ergas, & Mermelstein, 2008). Suspended air, on the other hand, does not appear to contribute significantly to fouling (Jang et al., 2004).

As the thickness of the biological fouling layer increases with extended MBR filtration, some biofilm regions have been observed to become anaerobic (Zhang, Choi, Dionysiou, Sorial, & Oerther, 2006c), therefore impacting on membrane fouling differently to a wholly aerobic film. Endogenous decay, similar to that expected within the fouling layer, was simulated and revealed the foulant levels and, specifically, the carbonate fraction of the EPS (Section 2.3.6.5) to significantly increase. Since the transition between aerobic to anaerobic conditions appears to produce a large amount of EPS, this phenomenon could also contribute to MBR fouling.

Foaming

Foaming in activated sludge plants is described as floating biomass and has been attributed to the combination of anthropogenic surfactants (detergents), biosurfactants (formed from micro-organisms) and/or two groups of filamentous bacteria: *Gordonia* spp. (or *Nocardia* sp.) and *Microthrix parvicella*. Foam in AS plants is regarded as a three-phase matrix, comprising gas bubbles, liquid (wastewater) and solid particles (the hydrophobic filamentous bacteria) (Davenport & Curtis, 2002). Filamentous micro-organisms are those whose cells are not separated following cell division and so tend to grow in the form of 'filaments'. *Gordonia* spp. are filamentous bacteria, known as Actinomycetes, which are extremely hydrophobic due to the presence of mycolic acids on their cell walls. *Microthrix parvicella* is also hydrophobic and utilizes long-chain fatty acids as a carbon source. It can store excess long-chain fatty acids in large globules and has an increased affinity over other bacteria for water-insoluble fats and lipids due to their hydrophobicity. The mycolic acids in their cell walls make them sufficiently hydrophobic to become attached on the gas bubbles present in activated sludge and rise to the surface of the liquid, increasing surface activity and promoting stable foams.

Foaming in AS plants is a well-studied problem by many researchers with significant impacts on the process efficiency. Several studies by various researchers have demonstrated a clear link between the AS foaming and the presence of surfactants, biosurfactants and the mycolic-acid containing micro-organisms. Recent studies (Hug, 2006; Heard, Harvey, Johnson, Wells, & Angove, 2008) have showed that initiation of AS foaming is due to surfactants and biosurfactants, although critical concentrations for foam initiation have not been quantified due to the numerous compounds involved and their variability between different sludges. Foam stabilization is mainly due to the filamentous *Gordonia* and *M. parvicella* but there is evidence suggesting that non-filamentous mycolic acid-containing micro-organisms, of which specific species have not yet been identified, also act as stabilizing agents. The degree of foaming is reported as being related to the protein EPS concentrations (Nakajima & Mishima, 2006). In MBRs foaming sludges appear to yield lower membrane permeabilities (Chang & Lee, 1998), attributed to the higher hydrophobicity of foaming activated sludge. Foaming thus provides an indication of sludge fouling propensity, and severe foaming demands ameliorative measures (Section 3.6.5).

2.3.6.4. *Floc Characteristics*

Floc Size

Comparison of the aggregate size distribution of ASP and MBR sludges (Spérandio, Massé, Espinosa-Bouchot, & Cabassud, 2005) revealed a distinct difference in terms of mean particle sizes (160 and 240 μm, respectively). A bimodal distribution was observed for the MBR sludge (5−20 and 240 μm), the high concentration of small colloids, particles and free bacteria being caused by their complete retention by the membrane. In another study where partial characterization of the MBR flocs up to 100 μm was carried out, floc sizes ranging from 10 to 40 μm were reported with a mean size of 25 μm (Bae & Tak, 2005). Floc size distributions reported for three MBRs operated at different SRTs were similar, although the mean floc size increased slightly from 5.2 to 6.6 μm for SRTs, increasing from 20 to 60 days (Lee, Kang, & Shin, 2003).

Given the large size of the flocculant solids compared to the membrane pore size, pore plugging by the flocs themselves is not possible. Flocs are also to some extent impeded, by drag forces and shear-induced diffusion, from depositing on the membrane surface. They nonetheless contribute to fouling through production of EPS and also directly affect clogging of the membrane channels. The interaction between EPS levels and floc size is discussed in Section 2.3.6.5, and the use of ancillary materials to suppress fouling through floc size and structure discussed in Section 2.3.9.5.

Hydrophobicity and Surface Charge

A number of reports can be found in the literature providing evidence of membrane fouling by highly hydrophobic flocs. Relative floc hydrophobicity can be directly measured by bacterial adhesion/partition using hydrocarbons such as hexane (Jang et al., 2005b), or estimated by contact angle determination (Yu et al., 2005b). Although the direct effect of floc hydrophobicity on MBR fouling is difficult to assess, hydrophobicity measurement of sludge and EPS solutions has been reported by Jang et al. (2005a,b) and Jang, Ren, Cho, and Kim (2006). EPS level and the filamentous index (a parameter related to the relative presence of filamentous bacteria in sludge) directly influence biomass floc hydrophobicity and zeta potential. Excess growth of filamentous bacteria has been reported to yield higher EPS levels, lower zeta potentials, more irregular floc shape and higher hydrophobicity (Meng et al., 2006). Sludge of higher foaming propensity, attributed to its hydrophobic nature, has been shown to produce a flux decline 100 times greater than that of non-foaming sludge (Chang & Lee, 1998). The anionic nature of the functional groups of natural organic materials means that charge and zeta potential of activated sludge flocs (and EPS) tend to be in the 0.2−0.7 meq/gVSS and 20−30 mV regions, respectively (Lee, Kang, & Shin, 2003; Liu & Fang, 2003). Increasing SRT has been shown to produce an increase in both contact angle and surface charge, with an apparently strong correlation with fouling propensity (Lee et al., 2003).

2.3.6.5. Extracellular polymeric substances

Extracted EPS (eEPS) and SMP

As already stated, membrane fouling in MBRs has been largely attributed to EPS (Nagaoka, Ueda, & Miya, 1996; Chang & Lee, 1998; Nagaoka, Yamanishi, & Miya, 1998; Cho & Fane, 2002; Rosenberger & Kraume, 2002), the construction materials for microbial aggregates such as biofilms, flocs and activated sludge liquors. 'EPS' is a general term encompassing all classes of autochthonous macromolecules such as carbohydrates, proteins, nucleic acids, (phosphor)lipids and other polymeric compounds found at or outside the cell surface and in the intercellular space of microbial aggregates (Flemming & Wingender, 2001). These consist of insoluble materials (sheaths, capsular polymers, condensed gel, loosely bound polymers and attached organic material) secreted by the cell, shed from the cell surface or generated by cell lysis (Jang et al., 2005a). Functions of the EPS matrix include aggregation of bacterial cells in flocs and biofilms, formation of a protective barrier around the bacteria, retention of water and adhesion to surfaces (Laspidou & Rittmann, 2002). With its heterogeneous and changing nature, EPS can form a highly hydrated gel matrix in which microbial cells are embedded (Nielson & Jahn, 1999) and can thus help create a significant barrier to permeate flow in membrane processes. Finally, bioflocs attached to the membrane can provide a major nutrient source during biofilm formation on the membrane surface (Flemming, Schaule, Griebe, Schmitt, & Tamachkiarowa, 1997). Their effects on MBR filtration have been reported for more than 15 years (Ishiguro, Imai, & Sawada, 1994) and have received considerable attention in recent years (Meng et al., 2009).

Analysis of EPS relies on its extraction from the sludge flocs (Fig. 2.31). No standard method of extraction exists, making comparison of reported data generated from different extraction methods difficult. The latter include cation exchange resin (Frolund, Palmgren, Keiding, & Nielsen, 1996; Gorner, de Donato, Ameil, Montarges-Pelletier, & Lartiges, 2003; Jang et al., 2005a), heating (Morgan, Forster, & Evison, 1990) and organic solvent (Zhang, Bishop, & Kinkle, 1999). The relative efficacies of these techniques, along with a number of others, have been compared (Liu & Fang, 2003); results suggest formaldehyde extraction to be the most effective in extracting EPS. However, because of its simplicity, the heating method is sometimes preferred (Fig. 2.31). Regardless of the extraction method used, a distinction can be made between EPS which derives directly from the active cell wall and that which is not associated with the cell but is solubilized in the mixed liquor. The former is usually referred to as 'EPS' in the literature, although a less ambiguous term

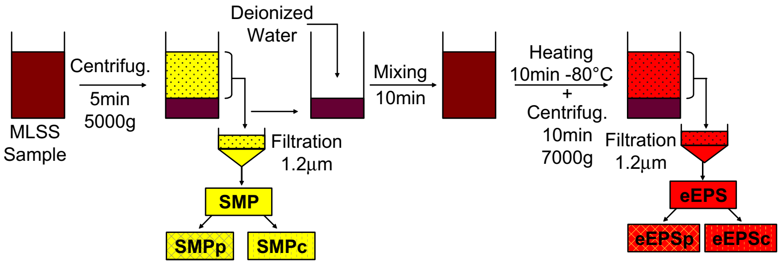

FIG. 2.31 Candidate method for EPS and SMP extractions and measurements.

would be 'eEPS' (*extracted* EPS, Fig. 2.32). The latter is normally termed soluble microbial product (SMP) and invariably refers to clarified biomass, although for some more recalcitrant feedwaters, clarified biomass will inevitably contain feedwater constituents which remain untransformed by the biotreatment process. SMP comprises soluble cellular components released during cell lysis, which then diffuse through the cell membrane and are lost during synthesis or are excreted for some purpose (Laspidou & Rittmann, 2002; Li et al., 2005a). In MBR systems, they can also be provided from the feed substrate. It is widely accepted that soluble EPS and SMP are identical (Jang, Ren, Choi, & Kim, 2005a; Laspidou & Rittmann, 2002; Rosenberger et al., 2005).

Protein and Carbohydrate EPS Fractions

Typically, the EPS solution is characterized according to its relative content of protein (EPSp) and carbohydrate (EPSc), measured by the respective photometric methods of Lowry, Rosebourgh, Farr, and Randall (1951) and Dubois, Gilles, Hamilton, Rebers, and Smith (1956). Reported data are summarized in Table 2.9. While EPSp generally has hydrophobic tendencies, EPSc is more hydrophilic (Liu & Fang, 2003) and may therefore interact more strongly with the membrane. The EPS solution can also be characterized in terms of its TOC level (Cho, Song, Lee, & Ahn, 2005b; Nagaoka & Nemoto, 2005) and, less frequently, its hydrophobicity by measurement of the ultraviolet absorbance per

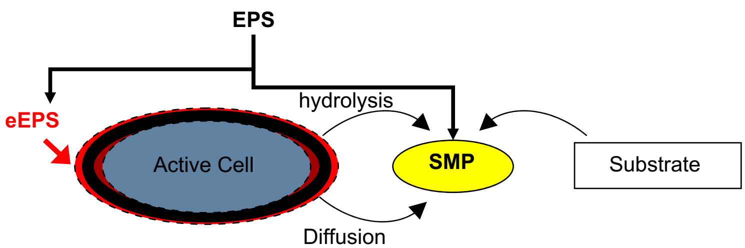

FIG. 2.32 Simplified representation of EPS and SMP.

TABLE 2.9 Concentration of the EPS Components in Different MBR Systems (mg/gSS Unless Otherwise Stated)

EPSp	EPSc	Other	Details	Ref.
25–30	7–8	Humic: 12–13	R, (10)	Spérandio et al. (2005)
29	36	SUVA: 2.8–3.1 L/m.mg	S	Ahn et al. (2005)*
120	40		S, (∞)	Gao et al. (2004b)
31–116	6–15	TOC: 37–65	Four pilot-scale plants, Municipal	Brookes et al. (2003b)
20	14		Pilot-scale plant, Industrial	
11–46	12–40	TOC: 44–47	Three full-scale plants, Municipal	
25	9	TOC: 42	Full-scale plants, Industrial	
		EPSp + EPSc = 8		Jang et al. (2005a)
30–36	33–28		(20–60)	Lee et al. (2003)
73	30		S, (∞)	Le-Clech et al. (2003b)
60	17		R, (∞)	Le-Clech et al. (2003b)
		TOC: 250 mg/L	S, MLSS: 14 g/L	Nagaoka and Nemoto (2005)
		TOC: 26–83 mg/gVSS	(8–80)	Cho et al. (2005b)
116–101	22–24		S, (20)	Ji and Zhou (2006)

S: Synthetic wastewater, R: Real wastewater; (SRT in days in parentheses; ∞ = infinite SRT – no wastage).
**Anaerobic UASB and aerobic MBR (Ahn et al., 2005).*

unit TOC concentration, the specific UV absorbance (SUVA) (Ahn et al., 2005). In many reported cases, EPSp (with a maximum concentration of 120 mg/gSS) is greater than EPSc (maximum concentration of 40 mg/gSS) and the total concentration range reported is surprisingly narrow: 11–120 mg/L for EPSp and 7–40 mg/L for EPSc. Sludge flocs have also been characterized in terms of protein and carbohydrate levels, though colorimetric analysis carried out directly on the washed biomass (Ji & Zhou, 2006) yielded no evident correlation between these indicators and MBR fouling propensity. Finally, the measurement of humic substances, generally overlooked for protein and carbohydrate, has revealed that they arise in significant concentrations in activated sludge liquors (Liu & Fang, 2003).

A functional relationship between specific resistance, MLVSS, TMP, permeate viscosity and EPS has been obtained by dimensional analysis (Cho et al., 2005b). EPS was found to have no effect on the specific resistance below 20 and above 80 mgEPS/gMLVSS, but played a significant role in MBR fouling between these two limits.

Analysis of EPS isolates is normally by UV absorbance, though more extensive analysis has been conducted by a number of authors. In a study of an intermittently aerated MBR, the EPS fraction was found to feature three main peaks at 100, 500 and 2000 kDa following gel chromatographic analysis. EPS larger than 1000 kDa in MW were assumed to be mainly responsible for MBR fouling (Nagaoka & Nemoto, 2005). High-performance size exclusion chromatography (HPSEC), a technique more widely used for potable raw water analysis for allochthonous natural organic matters (NOM) characterization (Nissinen, Miettinen, Martikainen, & Vartiainen, 2001), has been applied to EPS. Analysis of EPS fractions obtained from MBRs at different locations revealed their EPS profiles to be similar (Brookes, Jefferson, Le-Clech, & Judd, 2003a; Brookes et al., 2003b; Jefferson, Brookes, Le-Clech, & Judd, 2004). This would seem to corroborate previous findings from ASP sludge based on size exclusion chromatography combined with infrared micro-spectroscopy techniques (Gorner et al., 2003), where EPS chromatographs exhibited seven distinct peaks. Analysis revealed 45−670 kDa MW proteins and 0.5−1 kDa MW carbohydrates to be present. The existence of low-MW proteins associated with carbohydrates was proposed as being pivotal in floc formation and may therefore be expected to play a significant part in MBR membrane fouling. EPS characterization pertaining to flocculation, settling and dewatering in conventional ASP technologies (Liu & Fang, 2003; Yin, Han, Lu, & Wang, 2004) may also be germane to MBR technologies. More recently, advanced techniques like Fourier transform infrared (FTIR) spectroscopy (Zhou, Yang, Meng, An, & Wang, 2007), nuclear magnetic resonance spectroscopy (NMR), (Kimura, Hara, & Watanabe, 2005) and liquid chromatography − organic carbon and nitrogen detector (LC-OCND), combining the advantages of HPSEC and TOC systems (Siembida, Cornel, Krause, & Zimmermann, in press) have also been successfully used to more fully characterize EPS and other organic compounds in MBR sludge.

Since the EPS matrix features in floc formation (Liu & Fang, 2003), and specifically in the hydrophobic interactions between microbial cells, a decrease in EPS levels may be expected to cause floc deterioration, as indicated by the results from a comparative study of nitrification/denitrification in an MBR (Jang et al., 2005a). This would seem to imply that too low an EPS level is detrimental to MBR performance, though there is no firm experimental evidence to prove this.

Many operating parameters, including gas sparging, substrate composition and organic loading rate (OLR), appear to affect EPS characteristics in the MBR, but SRT is probably the most significant. A decrease in EPS levels has been observed for extended SRTs, with this reduction becoming negligible at SRTs greater than 30 days (Brookes et al., 2003b). Lee et al. (2003) observed an increase in protein concentration (along with stable carbohydrate levels) when SRT was increased. Notwithstanding the crucial role of EPS in the biological activity of the reactor and within the biofilm formed on the membrane, recent research has tended to focus on the effect of soluble EPS (or SMP) on fouling intensity. Other exoploymeric substances of interest in the MBR field include biopolymeric clusters (Wang & Li, 2008; Sun, Wang, & Li, 2008) and transparent exopolymer particles (TEPs). Originally thought to play a major role in biofilms formed on reverse osmosis membranes (Berman & Holenberg, 2005), TEPs have also been extensively studied in MBR processes (De La Torre, Lesjean, Drews, & Kraume, 2008). However, the SMP fraction has been by far the most studied.

Soluble Microbial Products

Whilst the impact of dissolved matter on fouling has been studied for over 15 years, the concept of SMP fouling in an MBR is more recent (Chang, Le-Clech, Jefferson, & Judd, 2002a), with available data being reported within the last 8 years (Table 2.10). Experiments conducted with a dual compartment MBR, where the membrane was challenged ostensibly with the mixed liquor supernatant (i.e. the SMP) rather than the whole biomass (Ng et al., 2005), have revealed greater filtration resistance from the SMP than from the biomass at 4 g/L MLSS concentration. This implies that SMP characteristics have a significant impact on membrane permeability. During filtration, SMP materials are thought to adsorb onto the membrane surface, block membrane pores and/or form a gel structure on the membrane surface where they provide a possible nutrient source for biofilm formation and a hydraulic resistance to permeate flow (Rosenberger et al., 2005). SMP materials appear to be retained at or near the membrane. Biomass fractionation studies conducted by Lesjean et al. (2005b) revealed levels of carbohydrates, proteins and organic colloids to be higher in the SMP than in the permeate, a finding similar to those previously reported (Brookes et al., 2003a; Evenblij & van der Graaf, 2004).

Three methods of separating the water phase from the biomass, so as to isolate the SMP, have been investigated. Simple filtration through filter paper (12 μm) was shown to be a more effective technique than either centrifugation or sedimentation (Evenblij & van der Graaf, 2004). It is likely that removal of colloidal material would demand more selective pre-filtration, e.g. 1.2 μm pore size (Fig. 2.32). As with EPS, the SMP solution can be characterized with respect to its relative protein and carbohydrate content (Evenblij & van der Graaf, 2004), TOC level (Gao, Yang, Li, Yang, & Zhang, 2004b; Lyko et al., 2008) or with SUVA measurement (Shin & Kang, 2003), as well as MW

TABLE 2.10 Concentration of the SMP Components (in mg/L and mg/gSS)

SMPp	SMPc	Other	Operating Conditions	Ref.
8	25	Humic subs: 36	R, (10)	Spérandio et al. (2005)
		TOC: up to 8 mg/L	S, (∞)	Gao et al. (2004b)
0.5–9*	n.d.–10*	4–37* (TOC)	Four pilot-scale plants, Municipal	Brookes et al. (2003b)
0.5–1*	n.d.	11*	Three full-scale plants, Municipal	
0.5*	n.d.	1.5*	Full-scale plant, Industrial	
		TOC: 30–70 mg/L	S, (∞), MLSS 15 g/L	Liu, Huang, Chen, Wen, and Qian (2005)
23	7		R	Evenblij and van der Graaf (2004)
		DOC: 5 mg/L	S, (20)	Shin and Kang (2003)
		TOC: 8–10 mg/L	R, (21)	Tao et al. (2005)
10–34	5–33		R, (from 40 to 8)	Grelier et al. (2005)
4.5–6	4.5–3.7		S, (20)	Ji and Zhou (2006)

n.d.: Not detected, S: Synthetic wastewater, R: Real wastewater, SRT are given in days in brackets, ∞: infinite SRT (i.e. no wastage).
In mg/gSS.

distribution. HPSEC analysis conducted on SMP solutions has revealed the SMP MW distribution to differ significantly across a range of full-scale reactors operated under different conditions (Reid, Liu, & Judd, 2008), although the MW distribution for the eEPS fraction has been found to be similar (Brookes et al., 2003b). However, the SMP solution fingerprint was largely unchanged in weekly analysis conducted on a single reactor, indicating no significant change in SMP characteristics for biomass acclimatized to specific operating conditions. When compared to eEPS MW distribution, the SMP solution featured generally larger macromolecules.

Comparison between acclimatized sludges obtained from MBR and ASP pilot plants revealed similar levels of EPSp, EPSc and EPS humic matter (Spérandio et al., 2005). The membrane did not seem to affect the floc EPS content. However, corresponding levels of the SMP fractions were significantly higher for the MBR sludge. Critical flux tests carried out under the same conditions for both MBR and ASP sludge revealed a higher fouling propensity of the MBR sludge over that of the ASP; critical flux values were around 10–15 and 32–43 LMH, respectively. Since the measured levels of EPS were unchanged, it was surmised that the higher fouling propensity related to the SMP level. During this study, Spérandio and co-workers observed significant biological activity in the MBR supernatant, indicating the presence of free bacteria which may have contributed to fouling.

Between the years 2004 and 2005, a number of different studies indicated a direct relationship between the carbohydrate level in the SMP fraction and MBR membrane fouling directly (Lesjean et al., 2005b), or fouling surrogates such as filtration index and CST (Reid, Judd, & Churchouse, 2004; Evenblij, Geilvoet, van der Graaf, & van der Roest, 2005a; Grelier, Rosenberger, & Tazi-Pain, 2005; Tarnacki, Lyko, Wintgens, Melin, & Natau, 2005), critical flux (Le-Clech et al., 2005b) and permeability (Rosenberger et al., 2005). The hydrophilic nature of carbohydrate may explain the apparently higher fouling propensity of SMPc over that of SMPp, given that proteins are more generally hydrophobic than carbohydrates. Strong interaction between the hydrophilic membrane generally used in MBRs and hydrophilic organic compounds may be the cause of the initial fouling observed in MBR systems. However, the nature and fouling propensity of SMPc has been observed to change during unsteady MBR operation (Drews, Vocks, Iversen, Lesjean, & Kraume, 2006) and, in this study, it was not possible to correlate SMPc to fouling. Subsequent studies from Drews, Vocks, Bracklow, Iversen, and Kraume (2008) have demonstrated that no direct link exists between the concentration of carbohydrate in the supernatant and the extent of fouling in MBR systems (Section 2.3.7.2). However, the contribution of SMP, and not exclusively the carbohydrate fraction, to the formation of the fouling layer is undeniable.

The correlation of MBR membrane fouling with SMP protein has been less widely reported although, since a significant amount of protein is retained by the membrane – from 15%, according to Evenblij and van der Graaf (2004), to 90% (Drews et al., 2006) – it must be presumed that such materials have a role in fouling. The specific resistance apparently increased by a factor of 10 when the SMPp increased from 30 to 100 mg/L (Hernandez Rojas, Van Kaam, Schetrite, & Albasi, 2005). Against this, analysis of the fouling layer has revealed higher levels of carbohydrate and lower protein concentrations compared to those in the mixed liquor (Chu & Li, 2005; Zhang, Chua, Zhou, & Fane, 2006a), tending to reinforce the notion that SMPc is more significant than SMPp in MBR membrane fouling. Humic matter, on the other hand, may not significantly contribute to fouling due to the generally lower MW of these materials (Drews et al., 2006).

Many research studies have been based on synthetic/analogue wastewaters. Those analogues comprising the most basic constituents, such as glucose, are very biodegradable and, as such, would be expected to yield rather lower SMP levels than those arising in real systems. Since it may be assumed that there are almost no substrate residuals from glucose in the supernatant, the less biodegradable SMP induced by cell lysis or cell release would account for most of the supernatant EPS measured in such analogue-based studies and may explain the reduced influence of SMP compared with that of EPS reported in some of these studies (Cho, Song, Lee, & Ahn, 2005b). SUVA measurements carried out on MBR mixed liquor supernatant have confirmed the presence of organic matter originating from the decayed biomass and of larger MW and greater aromaticity and hydrophobicity in this fraction than in the analogue wastewater feed (Shin & Kang, 2003). This would seem to confirm that fouling materials are generated by biological action and arise as SMP, though once again the chemical nature of these products is obviously affected by that of the feed.

In another important study based on synthetic wastewater, Lee, Ahn, and Lee (2001a) revealed that levels of soluble organic matter in isolation cannot be used to predict MBR fouling. By comparing filterabilities of attached and suspended growth micro-organisms, Lee et al. observed the rate of membrane fouling of the attached growth system (0.1 g/L MLSS and 2 g/L attached biomass) to be about seven times higher than that of a conventional suspended growth MBR at 3 g/L MLSS. With similar soluble fraction characteristics in both reactors, it was concluded that the discrepancy arose from the formation of a protective dynamic membrane created by suspended solids in the suspended growth system, a conclusion subsequently corroborated by the work of Ng et al. (2005).

As expected, many operating parameters affect SMP levels in MBRs, and it is very unlikely that the level of carbohydrate or protein in the SMP could, alone, predict fouling propensity. As for EPS, SMP levels decrease with increasing SRT (Brookes et al., 2003b). For SRTs ranging from 4 to 22 days, SMPp and SMPc levels have been reported to decrease by factors of 3 and 6, respectively (Grelier Rosenberger, & Tazi-Pain, 2005).

The lack of a direct relationship between the biological parameters measured in the reactor and the extent of MBR fouling is also due to the preferential deposition of materials onto the membrane surface. Recent characterization of the fouling layers has established that the composition of the fouling layer differs significantly from that of the bulk activated sludge or supernatant (Metzger, Le-Clech, Stuetz, Frimmel, & Chen, 2007; Wang, Wu, Yin, & Tian, 2008; Al-Halbouni et al., 2008), the relative concentrations of protein and carbohydrate in particular being larger on the membrane surface. The more detailed characterization of three cleaning solutions obtained after rinsing, backwashing and chemical cleaning clearly highlighted this preferential deposition (Wu, Cui, & Xu, 2008). Non-uniformity of the cake layer is also manifested across the module as a whole, where regions of both static sludge cake (not removed by aeration) and thinner sludge film (readily removed by the passage of bubbles) have been reported (Chu & Li, 2005).

2.3.6.6. Anaerobic Systems

Anaerobic vs Aerobic Sludge Characteristics

Differences in characteristics between aerobic and anaerobic sludges are most readily attributed to the different mechanisms involved in the biological process. Aerobic biological suspensions mainly comprise micro-organisms and decay products which result from the high specific growth rates and biomass yields, which then impact on the sludge physicochemistry (Section 2.3.6.5) and are determined both by the feedwater quality and bioreactor operational parameters such as SRT and *F*/*M* ratio (Section 2.3.7.2). These changes relate to substrate bioavailability, which affects the excretion of extracellular polymeric substances (EPS) and which then enhance microbial aggregation when the substrate is growth-limiting. For anaerobic biomass, with low hydrolysis rates and biomass yield, the reactor solid inventory is considered to be mainly constituted by influent particulates (Lant & Hartley, 2007) that are of reduced particle size and density. Some physical characteristics such as particle charge, which affects colloidal interactions, have been reported to remain unchanged after digestion (Elmitwalli et al., 2001). It has been reported by various studies (Wilén, Keiding, & Nielsen, 2000) that aerobic sludge deflocculates under anaerobic conditions, due to the release of EPS from the biological matrix, leading to an increased supernatant turbidity and reduced filterability (Rasmussen, Bruus, Keiding, & Nielsen, 1994).

Mixed Liquor Suspended Solids Concentration and Particle Size Impacts

Studies have generally shown membrane permeability to decline with MLSS for a number of different anaerobic matrices, including synthetic sewage (Stuckey & Hu, 2003), digested sludge (Saw, Anderson, James, & Le, 1986) and distillery wastewater (Kitamura, Maekawa, Tagawa, Hayashi, & Farrell-Poe, 1996). Although fouling by sludge flocs is not considered as the main fouling mechanism under low flux operation, higher mixed liquor suspended solid (MLSS) concentrations have been shown to decrease permeability in anMBRs at TMPs above 2 bar in crossflow systems (Ghyoot & Verstraete, 1997). This was attributed to a noted almost 10-fold increase in the colloidal COD (8 μm filtered) concentration between an MLSS concentration of 6 and 25 g/L. In a previous crossflow study permeability at TMPs below 0.1 bar was shown to decrease with increasing MLSS between 0.4 g/L to 2.5 g/L, remaining constant thereafter (Beaubien, Baty, Jeannot, Francoeur, & Manem, 1996). Since in this case permeability appeared to be independent of CFV at Reynolds numbers between 2000 and 15,000, this trend was attributed to increased pore plugging (presumably by colloidal particles) rather than viscosity and concentration polarization. More recent studies on immersed low-pressure submerged systems (aniMBRs) have also reported high biomass concentrations to be detrimental to filtration performance. Jeison and van Lier (2006) found decreasing biomass concentrations by 50% to suppressed cake layer formation more significantly than doubling of the gas sparging intensity at MLSS concentrations of 25 and 50 g/L and membrane gas sparging rates of 1.2 and 2.4 $m^3 m^{-2} h^{-1}$. Similarly Stuckey and Hu (2003) reported a permeability decrease on increasing MLSS concentration from 7 to 35 g/L. The bacteria community has also been shown to impact on fouling through the colloidal content, with thermophilic bacteria apparently generating fivefold increase in cake layer resistance, associated with higher levels of supernatant colloidal/soluble COD (1−15 μm size range), over those of mesophilic bacteria (Lin et al., 2009).

Analysis of particle size distributions in pumped anMBRs (Elmaleh & Abdelmoumni, 1997; Choo & Lee, 1998; Cho & Fane, 2002; Bailey, Hansford, & Dold, 1994) indicates that sidestream systems yield average particle sizes between 3 (Cho & Fane, 2002) and 16 μm (Bailey, Hansford, & Dold, 1994). A similar examination of reported data for immersed systems using gas sparging (Jeison & van Lier, 2006; Hu & Stuckey, 2006; Lin et al., 2009) reveals the mean floc size to range between 65 μm (Hu & Stuckey, 2006) and 90 μm (Jeison & van Lier, 2006) − similar to aerobic MBRs. The reduction in particle size for sMBRs, generally attributed to the floc breakage during pumping in crossflow operation, is presumed to contribute to the decreased permeability of these systems but may also be responsible for the decreased biomass activity reported in some studies (Brockmann & Seyfried, 1996).

Organic Fouling by Extracellular Polymeric Substances

Comparison of aerobic and anaerobic systems by Liao, Allen, Droppo, Leppard, and Liss (2001) and Jia, Furumai, and Fang (1996) revealed more EPS to be generated under anaerobic conditions but the ratio of proteins to carbohydrates to be higher in aerobic systems. However, analysis of literature data (Table 2.11) suggests that both the total EPS and the protein:carbohydrate ratio are higher in aerobic systems, though data are highly scattered. Expressed as a percentage of total VSS, the range of the EPS content of aerobic and anaerobic sludges is 2.5−13.3% and 2−5.7%, respectively. Comparison of anaerobic and aerobic MBRs operated with complete retention of solids and fed with settled sewage (Baek & Pagilla, 2006) showed EPS levels to decrease continuously to a concentration of 27 and 33 mg/g_{VSS}, respectively. Another study of anMBRs (Lee et al., 2008) attributed the fast fouling rate observed following stable operation for 28 days to the sudden increase in extracted EPS from 30 to 235 mg_{TOC}/L. A similar trend was reported by Fawehinmi (2006), who observed an increase in specific resistance to filtration as the EPS content of crushed granular sludge increased from 20 to 130 mg/g_{VSS}. These EPS concentrations are among the highest found in anMBRs, probably resulting from the rupture of the granules. Unusually low EPS levels in aniMBRs have been reported for an expanded granular sludge bed (EGSB), in which granules provided 4.4−6.6 and 0.6−1.6 mg/g_{VSS} of EPSc and EPSp, respectively (Chu, Yang, & Zhang, 2005). Reported levels of extractable EPS from the membrane

TABLE 2.11 Concentration and Composition of Extracted EPS from Aerobic and Anaerobic MBR Sludge

System Configuration	EPS tot (mg/g_{VSS})	Ratio EPSp:EPSc	MLSS (g_{MLSS}/L)	K (LMH/bar)	Ref.
An/Granular	5–7.7	0.1–0.36	14-21	10	Chu et al. (2005)
An/Flocculant	52	0.86	8	72	Lin et al. (2009)
An/Flocculant	57	1.28	8	9.6	
An/Flocculant	24.9	3.01	9.4	25–125	Lee et al. (2008)
An/Flocculant	75*	8	6.5	–	Fawehinmi (2006)
Ae/Flocculant	63–70	1	2.8–5.5	–	Lee et al. (2003)
Ae/Flocculant	81–115	2.4	10–16	40–90	Trussell et al. (2007)
Ae/Flocculant	133**	1.6	18	–	Teck, Loong, Sun, and Leckie (2009)
Ae/Flocculant	40–70	2–4	1.9–6	–	Massé et al. (2006)

*Measured as COD.
**Sum of proteins and carbohydrates.

surface deposit were twice those found in the granules and the ratio of proteins to carbohydrate 2.5 times higher. For dispersed/flocculant aniMBR systems the EPS levels appear to be closer to those of the aerobic systems, though slightly lower. Regardless of the major fouling component, differences between biomass and cake layer EPS composition suggest that soluble or colloidal compounds are as responsible for the increase in membrane resistance in anaerobic as in aerobic MBRs, but that they are more onerous to permeation.

It has been recognized for some time that, as with EPS, the concentration of SMP normalized against influent COD is higher in aerobic (3.1%) than in anaerobic systems (0.2–2.5%), as reported by Barker and Stuckey (2001) based on earlier reported work (Noguera, Araki, & Rittmann, 1994; Kuo, Sneve, & Parkin, 1996). This is due to the lower biomass uptake and decay rates of anaerobic micro-organisms compared to aerobic biomass. Comparative studies of MBR systems with conventional reactors, however, have revealed the concentration of SMP to be higher in MBRs than for conventional biotreatment for both aerobic (Massé, Spérandio, & Cabassud, 2006) and anaerobic processes (Aquino, Hu, Akram, & Stuckey, 2006). This arises both because the high-molecular-weight organic fraction is retained by the membrane (Massé et al., 2006) and because higher SMP production arises by endogenous decay and cell lysis as a consequence of long sludge age operation and high loadings. Conversely to trends reported for conventional reactors, anaerobic MBRs have slightly higher SMP levels than aerobic systems. Analysis of literature data (Table 2.12) shows SMP levels normalized against influent COD to range from 10 to 22% and 10−50% for aerobic and anaerobic MBRs, respectively. Direct comparison between aerobic and anaerobic MBR systems operated in parallel (Baek & Pagilla, 2006) revealed residual COD concentrations in mixed liquor supernatant to be higher in anaerobic MBRs. Available data (Table 2.12) indicate that proteins generally dominate over the carbohydrate content in the SMP fraction in anMBRs.

Of greatest significance, however, is the colloidal matter. An order of magnitude difference in colloid concentration between aerobic and anaerobic systems has been reported for only an 80% difference in soluble COD concentration (Van Voorthuizen, Zwijnenburg, van der Meer, & Temmink, 2008), supporting earlier observations by Choo and Lee (1996b, 1998). As with all membrane systems, colloidal matter is transported more slowly back into the bulk solution than coarser particulate materials due to the lower diffusion rates (Choo & Lee, 1998); they thus tend to accumulate at the membrane surface to form a low-permeability fouling layer. They are also of a size which can plug the membrane pores, particularly for the larger pores of MF membranes, if able to migrate into the membrane. It is this that may account for the markedly diminished permeability of anaerobic MBRs as compared to the aerobic processes, notwithstanding the lower SMP organic carbon concentrations. The high colloid concentration is thought to reflect differences in biodegradation, which is much slower for the anaerobic process and involves several sequential steps such as hydrolysis, acidogenesis, acetogenesis and methanogenesis. Hydrolysis is thought to be a surface-based reaction taking place on influent solids which are converted to simple monomers by extracellular enzymes excreted by hydrolytic and fermentative bacteria. As a result, and due to the low hydrolysis rates and biomass yield of anaerobic bacteria, the reactor solid inventory is considered to be mainly constituted by influent particulates (Lant & Hartley, 2007) that are of reduced particle size (Elmitwalli et al., 2001). AnMBR solids properties, unlike those of aerobic processes, are thus more dependent on influent characteristics than on bioreactor operational parameters. Some physical characteristics such as particle charge, which affects colloidal interactions, have been reported to remain unchanged after digestion (Elmitwalli et al., 2001), and various studies (Wilén, Keiding, & Nielsen, 2000) have reported aerobic sludge to deflocculate under anaerobic conditions, due to the release of EPS from the biological matrix, increasing the supernatant turbidity and reducing filterability.

While the higher colloidal content of the anMBR sludge may reflect higher levels of free bacteria in the mixed liquor, the higher soluble organic concentration may result from lower biodegradation rates or SMP biodegradability under anaerobic conditions (Ince, Ince, Sallis, & Anderson, 2000). There is also

TABLE 2.12 Concentration and Composition of SMP in Aerobic and Anaerobic MBRs

Influent Type	SRT (d)	HRT (h)	Temp (°C)	SMP*$_{COD}$ (mg/L)	SMPc (mg/L)	SMPp (mg/L)	Ref.
Anaerobic							
Ethanol	–	19.7	35	250 (0.0025)			Lin et al. (2009)
	–	77	55	850 (0.0085)			
Synthetic	∞	15	35	1200 (0.24)	80	400	Harada et al. (1994)
Black water	–	12	35	327 (0.28)	81	70	Van Voorthuizen et al. (2008)
	–	12	35	269 (0.23)	45	69	
Synthetic	∞		15	150 (0.3)			Ho and Sung (2010)
	∞		25	50 (0.1)			
Synthetic	150	2.6	35	116 (0.26)	19	52	Aquino et al., 2006
Synthetic	>250	15	35	1789 (0.45)			Akram and Stuckey (2008)
Settled sewage	∞	12	30	51 (0.24)	–	–	Baek and Pagilla (2006)
Glucose	30	12	25	39 (0.07)	–	–	Huang et al. (2008)
	60	12	25	56 (0.1)	–	–	
Settled sewage	∞	8	12	180 (0.5)	8	59	Fawehinmi (2006)
Aerobic							
	20–60			36–42 (0.12–0.14)			Lee et al. (2003)
	10–30			37–82 (0.1–0.21)	12–26	10–79	Trussell et al. (2007)
	10–53			45–80 (0.12–0.22)	18–25	8–15	Massé et al. (2006)
					0.68–0.4	0–3.79	Brookes et al. (2003b)
	300				198	318	Teck et al. (2009)

**Values in brackets correspond to normalized SMP$_{COD}$ with respect to influent COD* evidence of high MW polymeric matter (up to 1000 kDa) being retained by the cake layer in anMBRs, supported by observed changes in SMP composition as a result of permeation (Stuckey, 2003). The apparently increased rejection capability accounts for the relatively high SMP levels found in the mixed liquor supernatant and the low MW (<1.5 kDa) of the permeate organic matter (Harada, Momonoi, Yamazaki, & Takizawa, 1994). Analysis of SMP concentrations from anMBRs (Table 2.12) seems to corroborate trends reported from conventional anaerobic chemostats, which indicate higher levels of SMP produced at higher wastewater strengths, sludge retention times (Noguera et al., 1994; Kuo et al., 1996) and lower temperatures. For instance, SMP concentrations of 150 mg L^{-1} (Hu & Stuckey, 2006; Chu, Yang, & Zhang, 2005) have been reported at SRTs of 145–150 days, whilst at lower SRTs of 30 and 60 days only 39 and 56 mg_{COD} L^{-1} were found, respectively (Huang, Ong, & Ng, 2008). Decreasing temperature from 25 °C by 10–14 °C has been shown to decrease COD removal efficiency by up to 16% (Wen, Huang, & Quian, 1999; Ho & Sung, 2010).

The use of supplementary dosing with PAC to ameliorate fouling has been extensively studied in membrane filtration of potable water and in aerobic MBRs, and such studies have also been conducted on anMBRs (Park, Choo, & Lee, 1999). It has been suggested (Choo & Lee, 1996b) that the addition of an adsorbent or a coagulant can enhance the permeate flux by agglomerating colloids to form larger particles of lower fouling propensity. The coarser and more rigid particles additionally improve scouring of the membrane surface. Dosing of anMBRs with ion-exchange resin has also been studied (Imasaka, Kanekuni, So, & Yoshino, 1989), with beneficial effects noted only at very high concentrations of 5 wt%.

2.3.7. Operation

2.3.7.1. Membrane Scouring

Aerobic Systems

Aeration is arguably the most important parameter in the design and operation of an aerobic MBR. As already stated, aeration is required for biotreatment (Section 2.2.5), floc agitation and membrane scouring (Dufresne, Lebrun, & Lavallee, 1997) and it is not necessarily essential or desirable to employ the same aerator for both duties. Ostensibly, air is used to lift the mixed liquor through the membrane module channels. However, the gas bubbles additionally enhance membrane permeation (Cui, Chang, & Fane, 2003) by inducing liquid flow fluctuations and local tangential shear transients, the shear rate γ (/s) being given by:

$$\gamma = \frac{\kappa U_L}{\delta}, \tag{2.35}$$

where U_L is the liquid CFV (m/s), δ is the separation (m) and κ is a membrane geometry-dependent constant. The effect of shear is to increase back transport, discouraging large particle deposition on the membrane surface and promoting mass transfer of liquid through the membrane (Section 2.1.4.4). Since it is proportional to the cube of particle diameter, lateral migration velocity for smaller particles is much less, leading to more severe membrane fouling by fine materials (Choo & Lee, 1998). Aeration also ameliorates clogging: one of the few formal studies of the accumulation of solids within a fibre bundle, albeit on a bench scale, suggests that SAD_p of 10 to be sufficient to effectively remove the solids from a fibre bundle, but that 'dead zones' can arise within the core of the bundle (Lebegue, Heran, & Grasmick, 2009).

It has long been recognized through studies of model systems (Cui & Wright, 1994; Cabassud, Laborie, & Lainé, 1997; Mercier-Bonin, Fonade, & Lafforgue-Delmorme, 1997; Ghosh & Cui, 1999) as well as MBRs themselves (Le-Clech, Alvarez-Vazquez, Jefferson, & Judd et al., 2003a; Le-Clech, Jefferson, Chang, & Judd, 2003b) that gas bubbles (or 'slugs') passing up through a tubular membrane are able to enhance the flux over that attainable from liquid crossflow at the same velocity. This type of two-phase air–liquid flow is termed 'slug flow' (Fig. 2.33) and represents the most effective type of air–liquid flow for promoting flux. Much work has been conducted, principally by Cui and his various co-workers, to model membrane aeration in channel flow. Thus far, models have been produced which describe the spatial variation of shear with time for rising bubbles as a function of bubble (or slug) size, channel dimension and geometry for Newtonian fluids. It is also possible, within certain boundary conditions, to relate γ to the flux, J, from first principles, provided assumptions can be made about the particle size and concentration, the system hydrodynamics and the fluid and membrane homogeneity. Such assumptions, however, are not pertinent to an iMBR where three-phase intermediate flow prevails in a highly heterogeneous non-Newtonian fluid containing solutes, colloids and particulates. Moreover, the system becomes yet more complicated when the geometry deviates from well-defined channels, as provided by FS or tubular configurations, to HF modules.

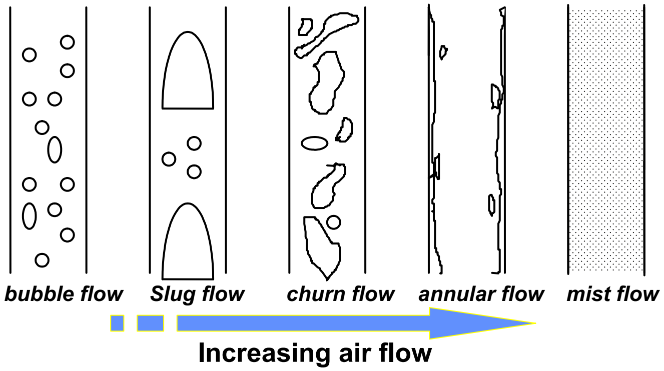

FIG. 2.33 Air–liquid flow regimes in a cylindrical channel *(Judd, Le-Clech, Cui, & Taha, 2001)*.

Aeration also affects HF iMBR performance by causing fibre lateral movement (or sway) (Côté, Buisson, & Praderie, 1998; Wicaksana, Fane, & Chen, 2006), which imparts shear at the membrane surface through the relative motion of the membrane and the surrounding liquid. In the case of HFs, effective distribution of air over the whole element cross-section and length becomes particularly challenging. For MT membrane modules in particular, provided an air bubble of diameter greater than that of the tube diameter is introduced into the tube, then air scouring of the entire membrane surface is assured. This is not necessarily the case for the FS and HF configurations, and HF systems additionally provide no fixed channel for the air bubble to travel up; this may contribute to the lower operating permeability of HF membranes. On the other hand, experimental studies and heuristic data reveal FS systems to demand generally higher aeration rates than HF systems to sustain higher membrane permeabilities, as reflected in aeration demand data from pilot-scale studies and full-scale operating plant (Section 3.2.2.1). Some HF systems are operated with intermittent aeration, lowering the aeration demand further, and the aeration demand of FS systems may be lowered by stacking the membrane modules such that the same volume of air is passed over twice the membrane area.

A number of authors (Ueda, Hata, Kikuoka, & Seino, 1997; Le-Clech et al., 2003c; Liu, Huang, Sun, & Qian, 2003; Psoch & Schiewer, 2005; Guglielmi, Chiarani, Judd, & Andreottola, 2007) have demonstrated that flux increases roughly linearly with aeration rate up to a threshold value beyond which no further increase in permeability takes place. It follows that operation is sub-optimal if the aeration rate, and specifically the approach velocity (Verrecht, Judd, Guglielmi, Mulder, & Brepols, 2008), exceeds this threshold value. Intense aeration has also been reported to damage the floc structure, reducing floc size and releasing SMP in the bioreactor (Rochex, Godon, Bernet, & Escudié, 2008; Menniti & Morgenroth, 2010) in the same way as has been reported for CFV in sMBRs. Given that aeration lifts the sludge through the module, a relationship must exist between gas and liquid velocity (U_G and U_L), respectively. Determination of U_L induced by aeration can be challenging; techniques such as electromagnetic flow velocimetry (Sofia, Ng, & Ong, 2004), particle image velocimetry (Yeo, Law, A.W.K. and Fane, 2007), electrochemical shear probe (Chan, Bérubé, & Hall, 2007) and constant temperature anemometry (Le-Clech, Chen, & Fane, 2006b) have all been used for liquid velocity estimation in iMBRs. Based on short-term critical flux tests, a direct comparison between immersed and sMBRs has shown similar fouling behaviour when the two configurations were, respectively, operated at a superficial gas velocity (U_G) of 0.07−0.11 m/s and CFV of 0.25−0.55 m/s (Le-Clech et al., 2005b). An increase of U_G in the iMBR was also found to have more effect in fouling removal than a similar rise of CFV in the sidestream configuration; tests on full-scale modules have demonstrated a roughly exponential decline in fouling rate with membrane aeration rate at a fixed flux (Monclùs et al., 2010).

Whilst a few studies have indicated improved performance with smaller bubbles (Fane et al., 2005), the greater majority of the literature acknowledges that large bubbles create more turbulence and so better fouling amelioration (Prieske, Prieske, Drews, & Kraume, 2008).

In practice, much development of commercial systems has been focused on reducing aeration whilst maintaining membrane permeability, since membrane aeration contributes significantly to the overall energy demand. A key parameter is thus the specific aeration demand (SAD), either with respect to membrane area (SAD_m in Nm^3 air/(hm^2)) or permeate volume (SAD_p Nm^3 air/m^3 permeate). The latter is a useful unitless indicator of aeration efficiency, and values for this parameter, which generally range between 5 and 50, are now often quoted by the membrane suppliers. Analysis of six large-scale aerobic iMBRs by Verrecht et al. (2008) indicated values of SAD_m between 0.21 and 0.88 m h^{-1} for a selection of plants operating under optimal conditions, with fluxes sustained between 24 and 31 LMH. Further discussion of specific aeration demand is provided in Chapter 3 and values from case studies are included in Chapter 5.

2.3.7.2. *Solid Retention Time (SRT)*

SRT impacts on fouling propensity through MLSS concentration, which increases with increasing SRT, and in doing so reduces the *F*/*M* ratio (Equation (2.23)) and so alters the biomass characteristics. Extremely low SRTs of ~2 days have been shown to increase the fouling rate almost 10 times over that measured at 10 days, with the *F*/*M* ratio correspondingly increasing from 0.5 to 2.4 g COD/(gVSS day) and the MLSS increasing only slightly from 1.5 to 1.2 g/L (Jang et al., 2005b). In practice, the *F*/M ratio is generally maintained below 0.2/day, although there is a trend towards decreasing the SRT to suppress MLSS concentration.

Operation at long SRTs minimizes excess sludge production but the increase in MLSS level which inevitably takes place presents problems of clogging of membrane channels – particularly by inert matter such as hair, lint and cellulosic matter (Le-Clech et al., 2005a), membrane fouling and reduced aeration efficiency, as manifested in the α-factor (Fig. 2.20). Even after increasing membrane aeration by 67%, fouling of an HF sMBR has been reported to almost double on increasing the SRT from 30 to 100 days, producing a corresponding increase in MLSS levels from 7 to 18 g/L and a decrease in *F*/*M* ratio from 0.15 to 0.05 kg COD/(kg MLSS day) (Han, Bae, Jang, & Tak, 2005). At infinite SRT, most of the substrate is consumed to ensure the maintenance needs and the synthesis of storage products. The very low apparent net biomass generation observed can also explain the low fouling propensity observed for high SRT operation (Orantes, Wisniewski, Heran, & Grasmick, 2004). In such cases sludge production is close to zero.

Rosenberger et al. (2006) found the linear correlation between fouling and SMP only at 8 d SRT and not for 15 d. Grelier, Rosenberger, and Tazi-Pain (2006) observed the contribution from both SMP and colloidal matter to fouling to decrease with increasing SRT, corroborated by the results of Trussell, Merlo, Hermanowicz, and Jenkins (2007) based on permeability decrease data over 10−14 d of stable operation. Ahmed, Cho, Lim, Song, and Ahn (2007), based on an FS MF membrane and operation at SRTs up to 100 d under sub-critical conditions, showed specific cake resistance to decrease with SRT with no correlation with SMP but with bound EPS. This work suggested that pore size might play a greater role than physical cleaning mode for this type of fouling. Grelier et al. (2006) identified a correlation between fouling rate and SRT, but not between carbohydrate concentration and resistance or fouling rate for measurements on a full-scale plant. Only in a separate small scale (stirred cell) test was filterability found to be correlated with carbohydrate concentration, possibly because these tests were performed at a constant pressure of 0.5 bar and initial fluxes well above the critical flux which would then be expected to accelerate colloidal fouling. Geilvoet, Remy, Evenblij, Temmink, and van der Graaf (2006) determined the filterability of sludge from a hollow fibre MBR in an 8 mm tubular membrane at a crossflow velocity of 1 m/s and a flux of 60 LMH. Despite the high flux, they also found no clear correlation with SMP concentration. However, since both these test cell experiments were carried out without membrane air scouring, fouling mechanisms may not have been comparable to plant conditions (Schaller, Drews, & Kraume, 2006). Drews et al. (2008), whose results were obtained at rather high SRT (20−30 d), also confirmed that as SRT increases, the relevance of SMP for filtration resistance and fouling decreases. Results indicate that SMP influences fouling only under certain conditions such as larger pore size and low sludge age.

Scientific studies indicate that SRT is a key parameter in determining fouling propensity through MLSS and EPS fraction concentrations. On this basis, an optimum SRT can be envisaged where foulant concentrations, in particular in the SMP fraction, are minimized whilst oxygen transfer efficiency remains sufficiently high and membrane clogging stays at a controllable level (Meng et al., 2009). In practice, SRT tends not to be rigorously controlled. Moreover, SRT probably has less of an impact on fouling than feedwater quality and fluctuations therein. It is nonetheless generally accepted that high SRT tends to lead to a decrease in EPS and SMP (Ahmed et al., 2007), although a peak of biopolymeric materials is generally observed when the SRT is changed during unsteady operation (see below).

2.3.7.3. *Unsteady-state Operation*

Unsteady-state operation can arise from such things as variations in feedwater quality (and so organic load), permeate flow rate (and hence hydraulic load) and aeration rate, which are all known to impact on MBR membrane fouling propensity, along with other dynamic effects (Table 2.13). In an experiment carried out with a large pilot-scale MBR in which the effects of unstable flow

TABLE 2.13 Examples of Dynamic Effects in MBR Operation (see also Table 3.38)

Determinants	Parameters affected
Flow rate	Ultimate flux and rate of change
Feedwater quality	Ultimate composition and rate of change
MLSS dilution	Dilution factor and rate of concentration change
(Partial) aeration loss	Percentage and period of reduction
Backflush/cleaning loss	Period of loss
Hydraulic shock	Rate and level of flow increase
Saline intrusion	Ultimate concentration factor and rate of concentration change

and sludge wastage were assessed (Drews et al., 2006), it was established that the level of carbohydrate in the supernatant before and after each sludge withdrawal increased. Whilst the increase following wastage was thought to be due to the sudden stress experienced by cells due to biomass dilution (which in extreme cases is known to lead to foaming in full-scale plant), increase before sludge withdrawal was attributed to the high MLSS concentration and the resulting low DO level in the bioreactor. It was concluded that unsteady-state operation changed the nature and/or structure (and fouling propensity) of the carbohydrate rather than the overall EPS formation. These findings corroborated results previously reported on effects of transient conditions in feeding patterns: the addition of a pulse of acetate in the feedwater has been shown to decrease significantly the MBR biomass filterability due to the increase in SMP levels produced (Evenblij, Verrecht, van der Graaf, & Van der Bruggen, 2005b). More detailed characterization of the impact of a wide range of unsteady-state conditions on the EPS present in activated sludge has recently been presented (Yang & Li, 2009). Along with changes in DO level, variation in the ratio of monovalent and polyvalent cations present in the feedwater can result in sludge deflocculation, usually leading to increased supernatant SMP levels. In the experiments reported by Van Den Broeck et al. (2010), high monovalent/polyvalent ratios resulted in significant deflocculation and decline in hydraulic performance.

The effects of starvation conditions on the biological suspension have been assessed by incorporating different substrate impulses in batch tests (Lobos, Wisniewski, Heran, & Grasmick, 2005). Exogenous phases were followed by starvation periods, both characterized by the *S*/*X* (substrate to biomass concentration ratio) where high ratios led to multiplication of bacteria cells, whilst at low ratios MLVSS decreased, SMPp production was absent and bacteria lysis ceased. S/X closely relates to F/M ratio (Equation (2.23)), and the low F/M values generally used in MBRs are thus theoretically close to starvation conditions which are in turn likely to be beneficial to MBR operation on the basis of the reduced SMPp production and correspondingly reduced fouling.

The principal period of unsteady-state operation is during start-up when the system is acclimatizing. Cho, Song, Lee, & Ahn (2005b) reported temporal changes in the bound EPS levels when the MBR was acclimatized at three different SRTs (8, 20, 80 days). As expected from general trends described in Section 2.3.6.5, the EPS concentration was lower at the longer SRT (83 vs 26 mgTOC/gSS for SRTs of 8 and 80 days, respectively). An initial latent phase was observed in which EPS concentration did not vary significantly. However, EPS levels increased exponentially after 40 days of operation at an SRT of 8 days, and after 70 days when the MBR was operated at 20 days SRT. No change in EPS levels was observed during the 80 days of operation at 80 days SRT. For another MBR operated at infinite SRT, no significant changes in SMP concentration during 100 days of operation were observed, over which time period the MLSS increased from 1.8 to 4.5 g/L (Jinhua, Fukushi, & Yamamoto, 2006). In a further study, following a latent phase of 30 days, MLSS and SMP levels started to increase significantly and stabilized after 140 days of operation at infinite SRT, whereas EPS levels increased continuously from the start but also stabilized after 140 days (Gao, Yang, Li, Wang, & Pan, 2004a). Nagaoka and Nemoto (2005) observed an increase in MLSS concentration from 4 to 14 g/L over 100 days along with a steady increase in EPS (from 50 to 250 mgTOC/L). There therefore appears to be no distinct pattern regarding foulant species generation and start-up, other than a general trend of more stable foulant levels at longer SRTs.

The generation of MBR foulants arising from changes in salinity has been studied by Reid, Liu, and Judd (2006), and the literature on the CASP effects date back to the 1960s (Ludzack & Noran, 1965; Tokuz & Eckenfelder, 1979). Reports indicate changes in salinity to have a greater impact on biotreatment efficacy, as manifested in the outlet organic carbon concentration, than high salinity levels per se. According to Reid et al., SMP and EPS turbidity, EPSp and SMPc all increased when a shock load of sodium chloride was administered to an MBR in a way designed to mimic saline intrusion in coastal MBRs. As with other studies (Section 2.3.6.5), permeability decline correlated with SMPc.

Finally, seasonal variations of the environment are also expected to affect MBR performances. A long-term study revealed the buffering effect of long SRT on fouling behaviour. Although the fouling reversibility was observed to vary at short SRT of 13 days (i.e. greater fraction of irreversible fouling at high temperature), the impact of temperature variations on fouling was not observed for an SRT of 50 days (Miyoshi et al., 2009).

2.3.7.4. *Anaerobic and Anoxic Systems*

Specific Gas Demand, aniMBRs

Gas sparging to maintain high membrane permeabilities, as used in immersed aerobic systems, is more problematic in anMBRs since air cannot be used routinely. Air sparging has been shown to be effective in anMBRs (Lee, Jung, & Chung, 2001c) but the duration of air sparging was necessarily brief in this study (5 s every 10 min), providing little permeability promotion overall. The use of extended intermittent aeration has been reported for nitrification/denitrification MBR systems (Yeom, Nah, & Ahn, 1999; Nagaoka & Nemoto, 2005). In this less common scenario, a single tank was used for both anoxic and aerobic biological degradation. Filtration was carried out in only the aerobic phase to take advantage of the anti-fouling properties of the air scouring, since severe fouling has been reported when aeration ceases (Jiang et al., 2005; Psoch & Schiewer, 2005). Sparging with headspace gas in aniMBRs has been shown to be effective for immersed polymeric (Fawehinmi et al., 2004; Stuckey & Hu, 2003) and for sidestream ceramic membrane anMBRs (Kayawake, Narukami, & Yamagata, 1991). As with the aerobic MBRs, a maximum permeability arises at a certain gas flow rate (Imasaka, Kanekuni, So, & Yoshino, 1989; Stuckey & Hu, 2003).

Research into aniMBRs (Table 2.14) suggests that while the flux ranges between 2.4 (Lin et al., 2009) and 12 LMH (Hu & Stuckey, 2006), the applied SGD_m varies more widely between effectively no gas sparging for UASB systems (Wen, Huang, & Quian, 1999; Chu, Yang, & Zhang, 2005) and 3 $Nm^3/(m^2\ h)$ for flocculant ones (Imasaka, So, Matsushita, Furukawa, & Kanekuni, 1993; Hu & Stuckey, 2006; Lee et al., 2008). A study of the impact of upflow velocity on fouling rates of an HF membrane-based aniMBR operated under intermittent cycles of 3 min filtration and 1.5 min relaxation showed that increasing SGD_m from 3 to 8 $Nm^3/(m^2\ h)$ produced insufficient shear to sustain the flux (Chu et al., 2005). Hu and Stuckey (2006) employed a similar SGD_m of 3 $Nm^3/(m^2\ h)$ at a flux of 8 LMH and maintained a stable TMP of 0.4 bar during 90 days of operation. A stable permeability of 40 LMH/bar after 30 days of operation was also reported by Imasaka et al. (1993) for the same SGD_m value but with a liquid crossflow velocity of 0.2 m s^{-1} and 30 s of backwashing every half-hour; a stable TMP of 1 bar was maintained for 20 days in a mesophilic aniMBR operated at a flux of 7.2 LMH with a gas sparging intensity of 1.5 $Nm^3/(m^2\ h)$. Permeabilities of around 10 LMH/bar were reported for another aniMBR by Chu et al. (2005) and Wen et al. (1999) after two weeks of intermittent filtration of 3−4 min on/1.5 min off operation, the net flux being greatly reduced by this level of relaxation. Overall the permeability is much lower than for aerobic MBRs which for full-scale domestic wastewater treatment plants are generally between 150 and 250 LMH bar^{-1} (Section 3.2.2.1) and the SGD_p values at least an order of magnitude higher. The latter arises from specific scouring rates being up to eight times higher (for the FS configuration) and fluxes 2−3 times lower.

TABLE 2.14 Membrane Performance of Submerged Anaerobic MBRs

Reactor Type/sludge	Membrane Material Pore Size	Geometry	SGD (m/h)	Flux (LMH)	TMP (kPa)	Fouling Rate (kPa/h)	t_{op} (h)	Filtration Cycle (min) Filtration/relaxation	Ref.
UASB	PVDF/0.22	FS	1.8	25		0.33–2.52			Wu, An, Li, and Wong (2009)
UASB	PET	HF		5	<30	0.04–0.08			An, Wang, Wu, Yang, and Zhou (2009)
UASB	PE/0.1	HF	0	10.4	<100		480	3/1.5	Chu et al. (2005)
UASB	PE/0.03	HF	0	5	< 70	0.2	336	4/1	Wen et al. (1999)
UASB	PE/0.03	HF	0	10	<70	0.5	120		
CSTR	PE/250	HF	3	8–12	45	0	>2160	Continuous	Hu and Stuckey (2006)
CSTR	PE/0.22	FS	3	8–12		0	>2160		
Granular	PVDF/250	HF	0.27–0.54	10				8/1, plus 1 min backwash	Van Voorthuizen et al. (2008)
CSTR	PVDF/250	HF	1.35	8					
CSTR	PE/0.45	FS	3	5	2–10	0.0083	720	7/3	Lee et al. (2008)
UASB (Meso)	PVDF/70	FS	1.5	7.2	10	0	450		Lin et al. (2009)
UASB (Thermo)	PVDF/70	FS	1.5	2.4	25	0	250		

Crossflow Velocity, ansMBRs

As with all membrane processes (Section 2.1.4.5), increasing crossflow increases flux in sidestream systems (ansMBRs, Table 2.15) by suppressing the fouling layer concentration polarization (Grethlein, 1978; Saw, Anderson, James, & Le, 1986; Imasaka et al., 1989). As with most membrane separation processes, membrane resistance determines flux at low TMPs, with no reported impact of crossflow or MLSS concentration above a value of 2.5 g/L for the latter (Beaubien et al., 1996). At higher TMPs, crossflow (and thus surface shear) becomes important (Beaubien et al., 1996; Zhang et al., 2004), the flux increasing linearly with CFV (Beaubien et al., 1996) with the slope decreasing with increasing MLSS partly due to viscosity effects. However, a plateau has been reported at Reynolds numbers beyond ~2000 (i.e. with significant turbulence) where no further increase in permeability takes place (Elmaleh & Abdelmoumni, 1997, 1998; Choo & Lee, 1998; Choo et al., 2000). At very high TMPs, permeate flux has been shown to decrease with increasing TMP due to compaction of the fouling layer (Elmaleh & Abdelmoumni, 1997). However, this effect appears to depend on the membrane; Saw et al. (1986), filtering anerobic sludge, observed that at very high TMPs the permeate flux decreased with TMP for an MF membrane but was constant for an 8−20 kDa MWCO UF membrane. The authors suggested that this was due to the impact of the membrane substrate on the fouling layer structure, but a more likely explanation is migration of fines through the cake at higher TMPs into the more porous MF membrane, causing pore plugging (Beaubien et al., 1996).

For ceramic membranes, where the fouling layer is minimal, high crossflows have been reported as having a detrimental effect because the thinning cake layer offers less protection against internal fouling (Kang, 1996; Choo & Lee, 1998; Choo et al., 2000). Elmaleh and Abdelmoumni (1997) have reported close to zero fouling for crossflows above 3 m/s in a polymeric MT membrane ansMBR, with flux increasing linearly with shear stress up to this point. Baffles were shown by these authors to increase flux, the effect being greatest in the transition region between laminar and turbulent flow. However, the increase in flux attained by these measures is normally at the expense of a punitive increase in energy demand (Bourgeous, Darby, & Tchobanoglous, 2001) and non-uniform, and thus sub-optimal, TMP distribution (Lee, Burt, Rusoti, & Buckland, 1999). High-shear operation might also be expected to impact negatively on floc size and biomass bioactivity (Brockman & Seyfried, 1996; Ghyoot & Verstraete, 1997; Choo & Lee, 1998) with, at highest shears, cell lysis taking place, though such effects are apparently less severe for anaerobic than aerobic biomass (Elmaleh & Abdelmoumni, 1997).

2.3.8. Fouling Mechanisms in MBRs

Various reviews of MBR fouling encompassing fouling mechanisms have been presented in the literature (Chang et al., 2002a,b; Le-Clech et al., 2006b; Meng

TABLE 2.15 Membrane Performance of Sidestream Anaerobic MBRs

Reactor Configuration	Membrane Material Pore Size (μm)	MLSS (g/L)	CFV (m/s)	TMP (kPa)	Flux (LMH)	Fouling Rate (LMH/d)* (kPa/h)**	t_{op} (h)	Filtration cycle	Ref.
CSTR/ Suspended	Ceramic/0.22		2	40–50	15–18	0*	>1680	Continuous	Cadi, Huyard, Manem, and Moletta (1994)
CSTR/ Suspended	–/100		3	100	9	0*	3360	Continuous	Saddoud, Ellouze, Dhouib, and Sayadi (2006)
UASB	PVDF/0.22		0.93	2.5–15	30	<0.05**	360	Continuous	Cho and Fane (2002)
				2.5–40	40	0.38**	100		
				2.5–10	50	0.75**	10		
CSTR	Ceramic/0.2	1.6	2	35	65	0	>120	Continuous	Beaubien et al. (1996)
		7.2			47	0	>120	Continuous	
		22			27	0	>120	Continuous	
	PES/50#		1–1.1	115	10				He et al. (2005)
CSTR	PS/3000#	15	0.8	49	21–42	1	168–240	Continuous	Harada et al. (1994)
	–/20#	0.3†	0.57	200	10	0.5–0.7	720	Relaxation Depressurization	Choo and Lee (1996b)

**TMP rise rate.*
***Flux diminution rate.*
#Daltons.
†VSS.

et al., 2009). MBRs are routinely operated under notionally constant flux conditions with convection of foulant towards the membrane surface therefore maintained at a constant rate. Since fouling rate increases roughly exponentially with flux (Le-Clech et al., 2003b; McAdam et al., 2010a,b; Monclùs et al., 2010), sustainable operation dictates that MBRs should be operated at modest fluxes and preferably below the so-called critical flux (Section 2.1.4.6). As noted previously, even sub-critical flux operation can lead to fouling according to a two-stage pattern (Ognier et al., 2001; Wen et al., 2004; Brookes, Jefferson, Guglielmi, & Judd, 2006): a low TMP increase over an initial period followed by a rapid increase after some critical time period. Pollice, Brookes, Jefferson, & Judd (2005) reviewed the sub-critical fouling phenomenon, introducing the parameters t_{crit} and dTMP/dt to represent the critical time over which low-fouling operation at a rate of dTMP/dt is maintained. Prior to these two filtration stages, a conditioning period is generally observed (Zhang et al., 2006a,b). The three-stage process, wherein various mechanisms prevail, is summarized in Fig. 2.34.

Stage 1: Conditioning fouling

The initial conditioning stage arises when strong interactions take place between the membrane surface and the EPS/SMP present in the mixed liquor. Ognier, Wisniewski, & Grasmick (2002a) described rapid irreversible fouling in this initial stage, and passive adsorption of colloids and organics have been observed even for zero-flux operation and prior to

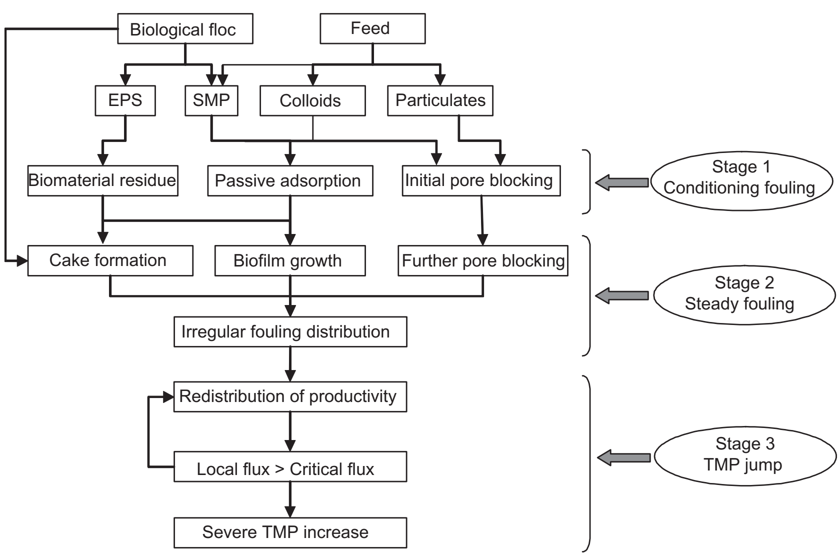

FIG. 2.34 MBR fouling mechanisms for operation at constant flux *(adapted from Zhang et al., 2006a–d)*.

particle deposition (Zhang et al., 2006a). Another detailed study based on passive adsorption revealed the hydraulic resistance from this process to be almost independent of tangential shear, and the initial adsorption to account for 20−2000% of the clean membrane resistance depending on the membrane pore size (Ognier et al., 2002b). In a more recent study, the contribution of conditioning fouling to overall resistance was found to become negligible once filtration takes place (Choi et al., 2005a). By applying a vacuum pump (rather than suction) coupled with air backflushing, Ma, Li, Du, Chen, and Shen (2005) were able to reduce colloidal adsorption onto the membrane. These studies suggest that colloid adsorption onto new or cleaned membranes coupled with initial pore blocking may be expected in MBRs (Jiang et al., 2005). The intensity of this effect depends on membrane pore size distribution, surface chemistry and especially hydrophobicity (Ognier et al., 2002a). In a test cell equipped with direct observation through a membrane operating with crossflow and zero flux, flocculant material was visually observed to deposit temporarily on the membrane (Zhang et al., 2006a). This was defined as a random interaction process rather than a conventional cake formation phenomenon. While some flocs were seen to roll and slide across the membrane, biological aggregates typically detached and left a residual footprint of smaller flocs or EPS material. Biomass approaching the membrane surface was then able to attach more easily to the membrane surface to colonize it and contribute to Stage 2.

Stage 2: Slow/steady fouling

Even when operated below the critical flux for the biomass, temporary attachment of the floc can contribute to the second fouling stage. After Stage 1, the membrane surface is expected to be mostly covered by SMP, promoting attachment of particulate and colloidal biomass material. Because of the low critical flux measured for SMP solutions, further adsorption and deposition of organics on the membrane surface may also occur during Stage 2. Since adsorption can take place across the whole surface and not just on the membrane pore, biological flocs may initiate cake formation without directly affecting flux in this initial stage. Over time, however, complete or partial pore blocking takes place. The rate of EPS deposition, and resulting TMP rise, would then be expected to increase with flux leading to a shorter Stage 2. Such fouling would prevail even under favourable hydrodynamic conditions providing adequate surface shear over the membrane surface. However, since uneven distribution of air and liquid flow is to be expected in iMBRs, correspondingly inhomogeneous fouling must take place.

Stage 3: TMP jump

With regions of the membrane more fouled than others, permeability is significantly less in those specific locations. As a result, permeation is promoted in less fouled areas of the membrane, exceeding a critical flux in these localities. Under such conditions, the fouling rate rapidly increases, roughly exponentially with flux. The sudden rise in TMP or 'jump' is a consequence of constant flux operation, and several mechanisms can be postulated for the rapid increase in TMP under a given condition. As with classical filtration mechanisms (Fig. 2.10), it is likely that more than one mechanism will apply when an MBR reaches the TMP jump condition, and a number of models can be considered:

i. *Inhomogeneous fouling (area loss) model*: This model was proposed to explain the observed TMP profiles in nominally sub-critical filtration of upflow anaerobic sludge (Cho & Fane, 2002). The TMP jump appeared to coincide with a measured loss of local permeability at different positions along the membrane, due to slow fouling by EPS. It was argued that the flux redistribution (to maintain the constant average flux) resulted in regions of sub-critical flux and consequently in rapid fouling and TMP rise.

ii. *Inhomogeneous fouling (pore loss) model*: Similar TMP transients have been observed for the crossflow MF of a model biopolymer (alginate) (Ye, Le-Clech, Chen, & Fane, 2005). These trends revealed the TMP transient to occur with relatively simple feeds. The data obtained have been explained by a model that involves flux redistribution among open pores. Local pore velocities eventually exceed the critical flux of alginate aggregates that rapidly block the pores. This idea was also the base of the model proposed by Ognier et al. (2004). While the 'area loss' model considers macroscopic redistribution of flux, the 'pore loss' model focuses on microscopic scale. In MBR systems, it is expected that both mechanisms occur simultaneously.

iii. *Critical suction pressure model*: The two-stage pattern of a gradual TMP rise followed by a more rapid increase has been observed from studies conducted based on dead-end filtration of a fine colloid by an immersed HF. At a critical suction pressure it is suggested coagulation or collapse occurs at the base of the cake, based on membrane autopsy evaluations supplemented with modelling (Chang, Fane, & Waite, 2005). A very thin dense layer close to the membrane surface, as observed in the study, would account for the rapid increase in resistance leading to the TMP jump. Although this work was based on dead-end rather than crossflow operation, the mechanism could apply to any membrane system where fouling continues until the critical suction pressure is reached, whereupon the depositing compound(s) coalesce or collapse to produce a more impermeable fouling layer.

iv. *Percolation theory*: According to percolation theory, the porosity of the fouling layer gradually decreases due to the continuous filtration and material deposition within the deposit layer. At a critical condition, the fouling cake loses connectivity and resistance, resulting in a rapid increase in TMP. This model has been proposed for MBRs (Hermanowicz, 2004), but indicates a very rapid change (within minutes), which is not always observed in practice. However, the combination of percolation theory with the inhomogeneous fouling (area loss) model could satisfy the more typically gradual inclines observed for TMP transients. Similarly, fractal theory was successfully applied to describe cake microstructure and properties and to explain the cake compression observed during MBR operation.

v. *Inhomogeneous fibre bundle model*: Another manifestation of the TMP transient has been observed for a model fibre bundle where the flow from individual fibres was monitored (Yeo & Fane, 2005). The bundle was operated under suction at constant permeate flow, giving constant average flux, and the flow was initially evenly distributed among the fibres. However, over time the flows became less evenly distributed so that the standard deviation of the fluxes of individual fibres started to increase from the initial range of 0.1−0.15 up to 0.4. Consequently, the TMP rose to maintain the average flux across the fibre bundle, mirroring the increase in the standard deviation of the fluxes. At some point, both TMP and standard deviation rose rapidly. This is believed to be due to flow maldistribution within the bundle leading to local pore and flow channel occlusion. It was possible to obtain steadier TMP and standard deviation profiles when the flow regime around the fibres was more rigorously controlled by applying higher liquid and/or airflows.

More recently, the TMP jump has also been explained by poor oxygen transfer existing within the fouling layer. As a result of transfer limitation, bacteria present in the biofilm layer can die, releasing extra levels of SMP. Experimental data have shown an increase in SMP concentration at the bottom of the fouling layer when the level of DO declines (Hwang et al., 2008).

2.3.9. Fouling Control and Amelioration in MBRs

Whilst an understanding of fouling phenomena and mechanisms may be enlightening, control of fouling and clogging in practice is generally limited to five main strategies:

1. applying appropriate pretreatment to the feedwater,
2. employing appropriate physical or chemical cleaning protocols,
3. reducing the flux,
4. increasing the aeration and
5. chemically or biochemically modifying the mixed liquor.

All of the above strategies are viable for full-scale operating MBRs, and each is considered in turn below.

2.3.9.1. Feed Pretreatment

It is generally recognized that the successful retrofitting of an ASP or SBR with an MBR is contingent on upgrading the pretreatment and, specifically, the screening. Whilst an MBR can effectively displace primary sedimentation, biotreatment and secondary solid–liquid separation, as well as tertiary effluent polishing, classical screens of around 6 mm rating are insufficient for an MBR. Such relatively coarse screens increase the risk of clogging of the membrane module retentate flow channels, especially by hairs in municipal wastewaters, which aggregate and clog both the membrane interstices and aeration ports (Section 3.6.2). HF membranes have a tendency to form aggregates of hair and other debris, which collect at the top of the membrane element. Hairs may then become entwined with the membrane filaments and are not significantly removed by backflushing or cleaning. FS membrane clogging occurs when inhomogeneous fouling takes place, causing localized dewatering or sludging in the membrane channels (Section 3.6.2), as well as at the channel inlet. If the aeration fails to remove this solids build-up, sludge accumulates above the blockage, increasing the affected excluded area. Fibres collecting in the aeration system can change the flow pattern and volume of air to the membranes, reducing the scouring and so promoting membrane fouling and clogging. Aerators are thus normally designed to resist clogging and/or allow periodic flushing with water.

Screening for clogging amelioration has been discussed by Frechen, Schier, and Linden (2007) for 19 European MBR plants, This report revealed most of the 10 German plants considered, to be fitted with horizontal or vertical slit screens of 0.5 or 1 mm diameter, most of these being protected by 3–6 mm slit screens, regardless of the membrane configuration (Table 2.16). The same report also revealed marked differences in the performance of identically rated screens with differently shaped apertures. A 0.75-mm mesh screen was found to remove 66% more solids and 2.5 times more COD than an identically rated slit screen.

2.3.9.2. Physical and Chemical Cleaning Protocols

Cleaning strategies have been outlined in Section 2.1.4.3, and protocols applied in practice are detailed in Chapters 4 (for comparative pilot plant studies) and 5 (full-scale reference sites) with a summary of these data presented in Section 3.2.3. The classification of the fouling types is often based on the method used to recover the initial permeability. With the wide range of strategies employed to remove fouling from the membrane surface, it is no surprise that different (and sometimes confusing and contradictive) definitions have been introduced. A practical definition of the various fouling types is given in Table 2.17. This is based on the rate at which the fouling is expected to form and the time interval between cleaning strategies applied to remove them (Kraume, Wedi, Schaller, Iversen, & Drews, 2009), and includes the term 'residual fouling' to differentiate between fouling removed by maintenance cleaning and that by recovery cleaning (see below).

Physical Cleaning

Key general cleaning parameters are duration and frequency, since these determine process downtime and if backflushing is used a further key parameter is the backflush flux. Less frequent, longer backflushing (600 s filtration/45 s

TABLE 2.16 Pre-treatment of Selected European MBR Plants (Frechen et al., 2007)

	WwTW	p.e.	Date	Membrane Type	Gap Geometry/size (mm) Stage 1	Stage 2
GERMANY	Rödingen	3000	1999	HF	slit,h 3.00	slit,h 0.50
	Markranstädt	12,000	2000	HF	slit,v 3.00	mesh 0.75
	Monheim	9700	2003	HF	slit,h 1.00	–
	Kaarst-Nordkanal	80,000	2004	HF	slit,v 5.00	mesh 1.00
	Waldmössingen	2600	2004	HF	slit,v 5.00	slit,h 0.50
	Seelscheid	11,500	2004	FS	slit,v 3.00	–
	Eitorf	7500	2005	FS	existent	slit,v 1.00
	Woffelsbach	6200	2005	FS	slit,v 3.00	slit,v 0.50
	Konzen	9700	2006	FS	slit,v 3.00	slit,v 0.50
	Bergheim-Glessen	9000	2007	HF	slit,v 6.00	mesh 1.00
EUROPE	Porlock/UK	3000	1998	FS	hole 3.00	–
	Swanage/UK	23,000	2000	FS	hole 6.00	hole 2.00
	Brescia/Italy	46,000	2002	HF	slit,h 3.00	hole 2.00
	Schilde/Belgium	10,000	2003	HF	slit,v 2.00	mesh 1.00
	Guéthary/France	10,000	2003	HF	– ,1.00	–
	Varsseveld/ Netherlands	23,000	2005	HF	slit,v 6.00	hole 0.80
	Rietliau/Czech Rep	22,000	2005	HF	slit,v 6.00	mesh 0.75
	Heenvliet/ Netherlands	3330	2006	FS	slit,v 6.00	hole 3.00
	Arenas de Iguna/Spain	20,000	2006	FS	hole, 3.00	–

p.e.: Population equivalent; slit,h/v: horizontal/vertical.

TABLE 2.17 Classification of Fouling (Adapted from Kraume et al., 2009)

Definition (with *Preferred Term*)	Fouling Rate (mbar/min)	Time Interval	Cleaning Method Applied
Cake, *reversible* or removable fouling	0.1–1	10 min	Physical cleaning (e.g. relaxation, backflush)
Residual fouling	0.01–0.1	1–2 weeks	Maintenance cleaning (e.g. chemically enhanced backflush)
Irreversible fouling	0.001–0.01	6–12 months	Chemical cleaning
Permanent, long-term or *irrecoverable* fouling	0.0001–0.001	Several years	Cannot be removed

backflushing) has been found to be more efficient than more frequent but shorter backflushing (200 s filtration/15 s backflush) (Jiang et al., 2005). In another study based on factorial design, backflush frequency (between 8 and 16 min) was found to have more effect on fouling removal than either aeration intensity (0.3–0.9 m^3/h per m^2 membrane area) or backflush duration (25–45 s) for an HF iMBR (Schoeberl, Brik, Bertoni, Braun, & Fuchs, 2005), with backflush strength having an intermediate impact (Wu, Cui, & Xu, 2008). Whilst more effective cleaning would generally be expected for more frequent, stronger and longer backflushing, possible permutations need exploring to minimize energy demand. This has been achieved through the design of a generic control system which automatically optimized backflush duration according to the monitored, TMP value (Smith, Vigneswaran, Ngo, Ben-Aim, & Nguyen, 2005). However, many such studies have not always taken account of the loss of productivity which results from the use of permeate during the backwashing.

Air can also be used to effect backflushing (Sun, Huang, Chen, & Wen, 2004) or to enhance backflushing with water. Up to 400% increase in the flux over that attained from continuous operation has been recorded using an air backflush, although in this case 15 min of air backflush were required every 15 min of filtration (Visvanathan, Yang, Muttamara, & Maythanukhraw, 1997). Whilst air backflushing is undoubtedly effective, anecdotal evidence suggests that it can lead to partial drying out of some membranes, which can then produce embrittlement and so problems of membrane integrity.

Membrane relaxation encourages diffusive back transport of foulants away from the membrane surface under a concentration gradient, which is further enhanced by the shear created by air scouring (Chua, Arnot, & Howell, 2002; Hong et al., 2002). Detailed study of the TMP behaviour during this type of operation has revealed that, although the fouling rate is generally higher than for continuous filtration, membrane relaxation allows filtration to be maintained for longer periods before the need for chemical cleaning arises (Ng et al., 2005). Relaxation is almost ubiquitous in modern full-scale iMBRs (Chapter 5), and studies assessing maintenance protocols have tended to combine relaxation with backflushing for optimum results (Vallero, Lettinga, & Lens, 2005; Zhang et al., 2005). A more systematic comparison of backflushing and relaxation operating conditions was proposed during short-term filtration periods of 24 h (Wu et al., 2008). Although the overall degree of fouling (in terms of TMP increase) was similar in the various operating conditions, tests revealed the nature of the incipient membrane fouling varied significantly with filtration mode.

In practice, physical cleaning protocols tend to follow those recommended by the suppliers. Relaxation is typically applied for 1–2 min every 8–15 min of operation, both for FS and for HF systems (Section 3.2.2.3). For HF systems, backflushing, if employed, is usually applied at fluxes of 1–3 times the operating flux and often supplements rather than displaces relaxation. It is likely that operation without backflushing, whilst notionally increasing the risk of slow accumulation of foulants on or within the membrane, conversely largely preserves the biofilm on the membrane, which affords a measure of protection. This fouling layer is substantially less permeable and more selective than the membrane itself, and thus can be beneficial to the process provided the total resistance it offers does not become excessive.

Chemical Cleaning

Physical cleaning is supplemented with chemical cleaning to remove residual and irreversible fouling (Fig. 2.35), with this type of cleaning tending to comprise some combination of:

- maintenance cleaning at moderate chemical concentrations on a twice weekly to monthly basis, designed to remove residual fouling; and
- intensive (or recovery) chemical cleaning (once or twice a year), used to remove the so-called irreversible fouling.

Maintenance cleaning is designed to maintain membrane permeability and so reduce the frequency of intensive cleaning. It is performed either with the membrane in situ, a normal CIP, or in the case of an immersed system sometimes with the membrane tank drained (referred to as 'cleaning in air', CIA). Intensive, or recovery, cleaning is either conducted ex situ or in the drained membrane tank to allow the membranes to be soaked in cleaning reagent. Intensive cleaning is generally carried out when further filtration is no longer sustainable because of a diminished permeability. Recovery chemical cleaning methods recommended by suppliers are all based on a combination of hypochlorite, generally at 0.1–0.5 wt%, for removing organic matter, and organic acid (either citric or oxalic, possibly supplemented with mineral acid to achieve a target pH of ~3) for removing inorganic scalants. Whilst some scientific studies of the impacts of chemical cleaning on the MBR system, such as the

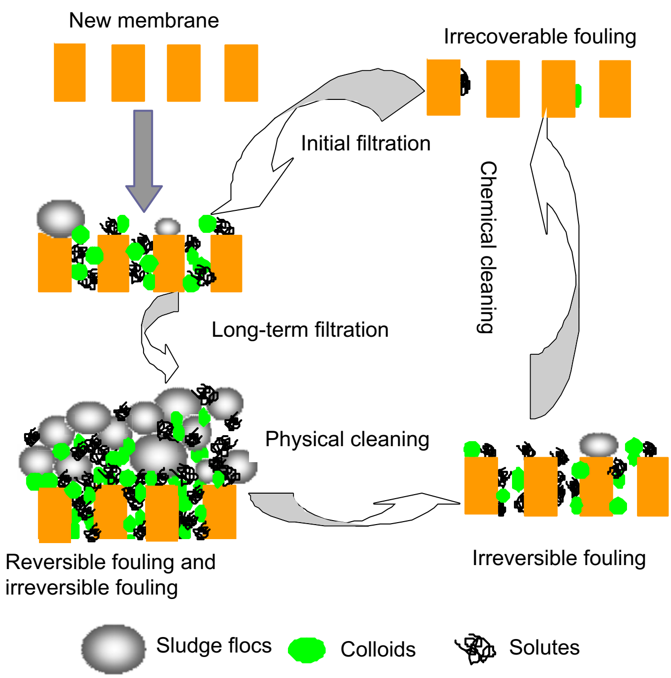

FIG. 2.35 Fouling and cleaning *(adapted from Meng et al., 2009)*.

microbial community (Lim, Ahn, Song, & Ji-nWoo, 2004), have been conducted, there has been no systematic study comparing the efficacy of a range of cleaning reagents or cleaning conditions on MBR permeability recovery. Some experiments with augmented cleaning, such as sonically enhanced processes (Lim & Bai, 2003; Fang & Shi, 2005), have been conducted. Whilst ultrasonic cleaning can undoubtedly enhance flux recovery, tests conducted in potable water suggest that it can result in adverse impacts on membrane integrity (Masselin et al., 2001).

Maintenance cleaning, usually taking 30−120 min for a complete cycle, is normally carried out every 3−7 days at moderate reagent concentrations of 200−500 mg/L NaOCl for classical aerobic MBRs. Recovery cleaning employs rather higher reagent concentrations of 0.2−0.3 wt% NaOCl, coupled with 0.2−0.3 wt% citric acid or 0.5−1 wt% oxalic acid (Section 3.2.2.3). Membrane cleaning studies on anaerobic systems have generally indicated that a combination of caustic and acid washes is required to remove organic and inorganic (namely, struvite) foulants from organic anMBR membranes (Choo et al., 2000; Lee et al., 2001c; Kang et al., 2002). For inorganic membranes, acid washing has been found to be less effective, and this has been attributed to surface charge effects (Kang et al., 2002). However, the membrane ageing and fouling history, variations in feedwater and biomass characteristics and differing operation and maintenance protocols make systematic optimization of cleaning protocols challenging. Such optimization is generally only achieved heuristically, and a thorough testing of various cleaning agents in a large pilot plant has enabled recovery cleaning to be delayed by ~2 years in one case (Brepols, Drensla, Janot, Trimborn, & Engelhardt, 2008).

Feedback Control Systems

Given the constant variations in the biomass nature and the temporal development of the fouling layer on the membrane surface, for any MBR system a pre-determined operating mode is likely to be sub-optimal for at least some of the time. Since 2003, feedback control systems have been proposed to optimize the use of anti-fouling strategies in MBRs. Based on a simple polynomial model calibrated by consecutive cycles (Busch & Marquardt, 2009), or simply based on permeability drop (Smith, Vigneswaran, Ngo, Nguyen, & Ben-Aim, 2006; Ginzburg, Peeters, & Pawloski, 2008), control systems developed have resulted in a reduction of backflush duration up to 25% or up to 50% in membrane aeration. Another relatively simple on-line method involved the combined monitoring of permeate flow rate, TMP and temperature to determine permeability and optimize the maintenance process on this basis (Joss, Boehler, Wedi, & Siegrist, 2009). More complex systems, taking into account the impact of growth of biofilm, concentration polarization phenomena and pressure drop in the permeation line, have been successfully designed, although they still require extensive calibration (Busch & Marquardt, 2009; Yeon et al., 2009). The successful application of control devices is possible only if appropriate inputs and outputs are properly defined and the integrity of the data is assured. System outputs can include control of the permeate pump (on/off or speed), the relaxation frequency, duration and membrane aeration rate, the backflush frequency, duration and flux, and the filtration membrane aeration rate, although adjustment of the aeration rate is also possible only if variable-speed blowers have been installed.

Another strategy has been developed recently to attempt to better predict high fouling rate, and involves the use of a small dedicated filtration apparatus to assess the filterability of the biomass at a given time. The Delft filtration characterization method (DFCm) comprises a sidestream membrane system, in which 30 L of sampled biomass is filtered following a standardized protocol (Evenblij et al., 2005b). To avoid biomass handling and to obtain a faster response, two other filtration systems have been developed to be directly submerged into the MBR. The VITO fouling measurement (VFM) uses a single tubular membrane and the Berlin filtration method (BFM) is based on flat sheet configuration (Huyskens et al., 2008; De La Torre, Iversen, Moreau, & Stuber, 2009, respectively). Whilst these systems all have the advantage of employing a standard method for sludge characterization, it is uncertain as to whether they offer a significant advance on feedback control based on permeability measurement of the actual process membranes.

2.3.9.3. Flux Reduction

Reducing the flux always reduces fouling but obviously then impacts directly on capital cost through membrane area demand. A distinction must be made between operating (i.e. gross) flux and net flux (the flux based on throughput over a complete cleaning cycle), as well as peak and average flux. Historically there appear to have been two modes of operation of an MBR regarding operating flux, which then determine the cleaning requirements and thus net flux:

- *Sustainable permeability operation*: In this instance, the conditions are chosen so as to maintain stable operation (little or negligible increase in TMP at constant flux) over an extended period of time (i.e. several weeks or months) with only moderate remedial measures (namely relaxation), if any. Most immersed FS and all sidestream systems have traditionally operated under these conditions, with sMBRs operating continuously (i.e. without relaxation) between chemical cleans.
- *Intermittent operation*: In this mode of operation, the operational flux is above that which can be sustained by the filtration cycle operating conditions and, as a result, intermittent remedial measures are employed. These comprise relaxation supplemented with backflushing and, usually, some kind of maintenance chemical cleaning procedure. All immersed HF systems operate in this manner.

Modern practice appears to favour operation at net fluxes of 18−25 LMH for iMBRs challenged with municipal wastewater, incorporating physical cleaning, regardless of membrane configuration. Maintenance cleaning, if employed, adds insignificantly to downtime. The greatest impact on operating vs net flux is therefore peak loading, often from storm waters if no flow balancing is provided. It is these increased hydraulic loads, coupled with feedwater quality fluctuations, which represent one of the major challenges to MBR design and operation. Most of the MBR suppliers allow their system to be operated at high flux (up to twice the normal value) to cope with potential peak loadings. However, these periods of high permeation are generally limited to a maximum of 1−2 h, and are sometimes coupled with increased aeration requirement and followed by extended relaxation periods (at lower flux) to allow the fouling accumulated during the peak flow operation to be removed physically.

2.3.9.4. Aeration Increase

Whilst increasing aeration rate invariably increases the critical flux up to some threshold value, increasing membrane aeration intensity is normally prohibitively expensive. As already stated, much attention has been focused on commercial development of efficient and effective aeration systems to reduce the specific aeration demand, with possibly the most important publications arising in the patent literature (Miyashita et al., 2000; Côté, 2002) and including cyclic aeration (Rabie, Côté, Singh, & Janson, 2003) and jet aeration (Fufang & Jordan, 2001, 2002). The use of uniformly distributed fine air bubbles from 0.5 mm ports has been shown to provide greater uplift and lower resistance compared to a coarse aerator having 2 mm ports at similar aeration rates (Sofia, Ng & Ong, 2004). In the same study, a bi-chamber (a riser and down-comer) in an FS MBR has been shown to play a significant role in inducing high CFVs. The use of a variable aeration rate to increase the flux during peak loads has been reported for short-term tests (Howell, Chua, & Arnot, 2004) and on full-scale plants (Stone & Livingston, 2008; Ginzburg, Peeters, & Pawloski, 2008). There have additionally been a number of studies where flux has been correlated with aeration (Ueda, Hata, Kikuoka, & Seino, 1997; Monclùs et al., 2010), but it is generally recognized that increasing aeration beyond some threshold value has no impact on the membrane permeability (Section 2.3.7.1) and, as such, the value of increasing aeration during the filtration cycle to control fouling is questionable. On the other hand, effective uniform distribution of aeration to suppress clogging is of paramount importance.

2.3.9.5. Chemical/biochemical Mixed Liquor Modification

The biomass quality can be controlled biochemically, through adjustment of the SRT (Section 2.3.7.2), or chemically. In practice, SRT is rarely chosen on the basis of foulant concentration control. Instead a target value is almost invariably based on target water quality (for nitrification in particular), sludge production rate, membrane module clogging propensity and/or biomass aeration efficiency. However, studies have shown that a modicum of fouling control can be attained through the addition of chemicals.

Coagulant Coagulation

Ferric chloride and aluminium sulphate (alum) have both been assessed for membrane fouling amelioration, most extensively for potable systems but also for MBRs (Zhang, Sun, Zhaoa, & Gao, 2008c). In MBR-based trials, addition of alum to the reactor led to a significant decrease in SMPc concentration, along with an improvement in membrane hydraulic performances (Holbrook et al., 2004). Small biological colloids (from 0.1 to 2 μm) have been observed to coagulate and form larger aggregate when alum is added to MBR activated sludge (Lee et al., 2001b). Although more costly, dosing with ferric chloride was found to be more effective than alum. Ferric dosing of MBRs has been used for enhancing the production of iron-oxidizing bacteria responsible for the degradation of gaseous H_2S (Park, Lee, & Park, 2005). In this study, specific ferric precipitates like ferric phosphate and K-jarosite ($K\text{-}Fe_3(SO_4)_2(OH)_6$) were observed to foul the membrane. Pre-treatment of the effluent by pre-coagulation/sedimentation has been shown to provide some fouling reduction (Adham, DeCarolis, & Pearce, 2004), and pre-clarification is employed at some sewage treatment works. In another example, the ferric dosing was shown to control both irreversible fouling and suspension viscosity (Itonaga, Kimura, & Watanabe, 2004). Pre-coating of MBR membranes with ferric hydroxide has also been studied as a means of increasing permeability and improving permeate quality (Zhang et al., 2004). In this study, additional ferric chloride was added to remove non-biodegradable organics which accumulated in the bioreactor.

Adsorbent Agents

Addition of adsorbents into biological treatment systems decreases the level of organic compounds. Dosing with PAC produces biologically activated carbon (BAC) which adsorbs and degrades soluble organics and has been shown to be effective in reducing SMP and EPS levels in a comparative study of a sidestream and immersed hybrid PAC–MBR (Kim & Lee, 2003). Decreased membrane fouling has also been demonstrated in studies of the effects of dosing MBR supernatant with up to 1 g/L PAC (Lesage, Sperandio, & Cabassud, 2005) and dosing activated sludge itself (Li, He, Liu, Yang, & Zhang, 2005c), for which an optimum PAC concentration of 1.2 g/L was recorded. In the latter study, floc size distribution and apparent biomass viscosity were identified as being the main parameters influenced, resulting in a reduced cake resistance, when PAC was dosed into the bioreactor. Conversely, no significant improvement in performance was recorded when a concentration of 5 g/L of PAC was maintained in the bioreactor without sludge wastage (Ng et al., 2005). It was postulated that, under these conditions, the PAC was rapidly saturated with organic pollutants and that fouling suppression by PAC relies on its regular addition brought about by lower SRTs.

Experiments conducted with different system configurations based on immersed HF membranes allowed direct comparison of hydraulic performances for pre-flocculation and PAC addition. Under the operating conditions employed, pre-flocculation provided higher fouling mitigation than that of PAC addition (Guo, Vigneswaran, & Ngo, 2004). However, the use of both strategies simultaneously provided the greatest permeability enhancement (Guo et al., 2004; Cao, Zhu, Lu, & Xu, 2005).

A detailed mathematical model has been proposed for predicting performances for hybrid PAC–MBR systems (Tsai, Ravindran, Williams, & Pirbazari, 2004). The model encompasses sub-processes such as biological reaction in bulk liquid solution, film transfer from bulk liquid phase to the biofilm, diffusion with biological reaction inside the biofilm, adsorption equilibria at the biofilm–adsorbent interface and diffusion within the PAC particles. Numerous other studies in which the use of PAC has been reported for fouling amelioration have generally been limited in scope and have not addressed the cost implications of reagent usage and sludge disposal. Tests have been performed using zeolite (Lee et al., 2001b) and aerobic granular sludge, with an average size around 1 mm (Li et al., 2005b) to create granular flocs of lower specific resistance. Granular sludge was found to increase membrane permeability by 50% but also lower the permeability recovery from cleaning by 12%, which would be likely to lead to unsustainable operation. There have additionally been studies on the use of granular aerobic sludge in aerobic MBRs (Tay, Yang, Zhuang, Tay, & Pan, 2007; Zhou et al., 2007), as well as anMBRs (Chu, Yang, & Zhang, 2005).

Proprietary and Other Reagents

Other types of additives, based on cationic polymer-based compounds, have been recently developed to enhance membrane performance. The first product to appear on the market was *MPE50*, developed by Nalco for use in iMBRs., which has been tested by a number of authors (Yoon et al., 2005; Guo, Vigneswaran, Ngo, Kandasamy, & Yoon, 2008; Lee, Li, Noike, & Cha, 2008). The addition of 1 g/L of the reagent directly to the bioreactor has been shown to reduce the SMPc level from 41 to 21 mg/L (Yoon et al., 2005). The interaction between the polymer and the soluble organics in general, and SMPc in particular, was identified as being the main mechanism responsible for the performance enhancement. In another example, an MBR operated at an MLSS level as high as 45 g/L yielded a lower fouling propensity when 2.2 g/L of polymer was dosed into the bioreactor. A number of other 'anti-fouling' products have since become available, including MPL30 (Nalco), KD452 (Adipap), as well as generic chemicals such as chitosan and starch. A recent study has comprehensively compared a wide range of these flocculants, adsorbents and additives (Iversen et al., 2009a), and revealed high SMP removal by KD452 and an increased critical flux by MPE50, KD452 and starch. Details of pilot testing of these three compounds were also reported by the authors. Biological side effects, clearly observed during overdosing of the compounds, were observed for the use of $FeCl_3$ and chitosan (Iversen et al., 2009b).

2.3.10. Permeate Water Quality

The efficacy of membrane bioreactors (MBRs) with respect to permeate quality is generally governed by the key universally regulated water quality determinants of biochemical (and sometimes chemical) oxygen demand, suspended solids and ammonia. Nutrient concentration is of increasing significance and in some circumstances the bacteriological content may be regulated. It is generally accepted that MBRs provide excellent treated water quality, achieving generally four-to sixfold removal of pathogenic bacteria, almost complete removal of suspended solids, and often reducing ammonia or TKN levels to less than 1 mg/L (Chapter 5). It therefore stands to reason that only (a) onerous particles significantly smaller than the effective membrane pore size and (b) non-biodegradable dissolved materials present a challenge to the process. The latter constraint applies to all biotreatment and clarification processes; removal of such materials by the process is dependent entirely on (a) their affinity for the sludge solids and (b) the perm-selectivity of the membrane.

2.3.10.1. Suspended Matter

Since the maximum nominal membrane pore size is around 0.4 μm, somewhat smaller than most bacteria, the permeate product is substantially disinfected, provided the membrane remains intact. Moreover, whilst viruses are somewhat smaller than bacteria, evidence suggests that they are also retained in the bioreactor (Shang, Wong, & Chen, 2005). This arises because viruses require a host to remain viable, and are thus associated with solid matter which is also substantially larger in size than the membrane pores. This is corroborated by noted differences between rejection figures for 'native' and 'seeded' phage (Hirani, DeCarolis, Adham, & Jacangelo, 2010), the latter being artificially introduced and therefore remaining partially unassociated. The same study by Hirani et al., performed at pilot-scale using a number of commercial membranes (Section 3.2.1.2), revealed little correlation between membrane pore size and rejection of the seeded coliphage (Fig. 2.36), and consistently between 3 and 3.5 log removal (i.e. 99.9–99.97% rejection) for the indigenous species for five of the six membrane products tested.

All other available evidence suggests that the only solid material capable of permeating the membrane is colloidal matter. Such material is onerous only to downstream dense membrane processes, and reverse osmosis in particular. However, the levels of colloidal matter encountered in MBR permeate, as determined by the silt density index (SDI, Kennedy et al., 2008), is generally low enough to cause few problems to RO operation. MBR permeate quality with respect to suspended solids is only significantly diminished when the membrane integrity is compromised (Section 3.6.6), or when biofilms developing in the permeate side of the membrane slough off. It is the latter which are probably the main cause of turbidity spikes encountered from permeate quality monitoring, rather than a loss of membrane integrity.

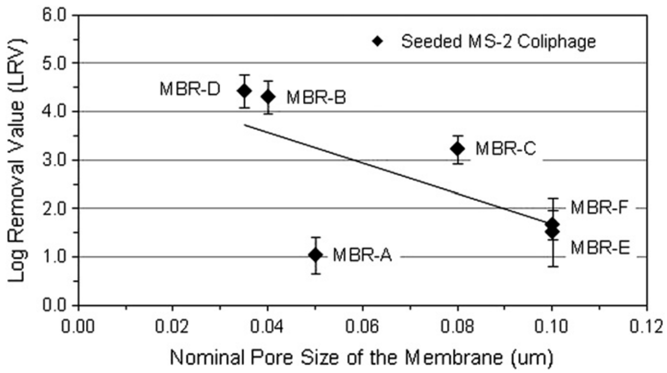

FIG. 2.36 Removal of seeded *MS-2 Coliphage* by the MBRs with different membrane pore size *(Hirani et al., 2010)*.

2.3.10.2. Soluble Matter

The uncoupling of HRT and SRT to allow higher solids concentrations may be expected to permit greater retention of recalcitrant dissolved materials by virtue of adsorption onto the commensurately higher solids surface area. Biodegradation of these species, however, relies on the development of a bacterial community at these long SRTs – ostensibly slow-growing ones – which are capable of breaking down the organics. There is conflicting evidence of the efficacy of extending the SRT in MBR operation. There is little evidence of any improvement in removal of pathogenic micro-organisms, including viruses (Zhang & Farahbakhsh, 2007; DeCarolis, Hirani, & Adham, 2009; Hirani et al., 2010). This is unsurprising given that particle rejection is similarly unaffected by its concentration in the reactor. However, there is also contradictory information concerning soluble organic and inorganic micropollutants.

Removal of the soluble species which are not directly rejected by the membrane since they are orders of magnitude smaller than the membrane pore size, can be either through degradation or phase change. A change of phase may occur through volatilization (assisted by aeration, i.e. sparging), adsorption onto solids or precipitation. Three important groups of key soluble contaminants can be identified:

- metals, present either as free ions or complexed with organic matter;
- organic micropollutants, such as pharmaceuticals and personal care products (PPCPs) and endocrine disrupting compounds (EDCs) generally; and
- nutrients, and specifically nitrate and phosphate, and their related compounds.

Metals

Metals cannot be biodegraded, but only assimilated into biomass or precipitated to expedite their removal. A recent review (Santos & Judd, 2010) of available information from municipal wastewater treatment trials suggests that no significant removal of dissolved metals takes place in a bioreactor, such that the improved performance of an MBR over that of the CASP is associated entirely with rejection of metals present as solids. It has been widely reported that complexation of some metals with organic ligands derived from extracellular polymeric substances is significant in determining their solubility (Guibaud, Comte, Bordas, Dupuy, & Baudu, 2005; Nakhla, Holakoo, Yanful, & Bassi, 2008). Notwithstanding this, there appears to be little evidence from pilot or full-scale plant to support the notion that MBR sludge provides significantly improved retention of dissolved metals. It appears that on average an MBR can be expected to reduce the residual metals concentration by around a factor of 2 over that of an ASP, this factor representing the separation of the two lines in Fig. 2.37. This improvement arises ostensibly through improved retention of the insoluble or adsorbed fraction. Moreover, whilst some authors

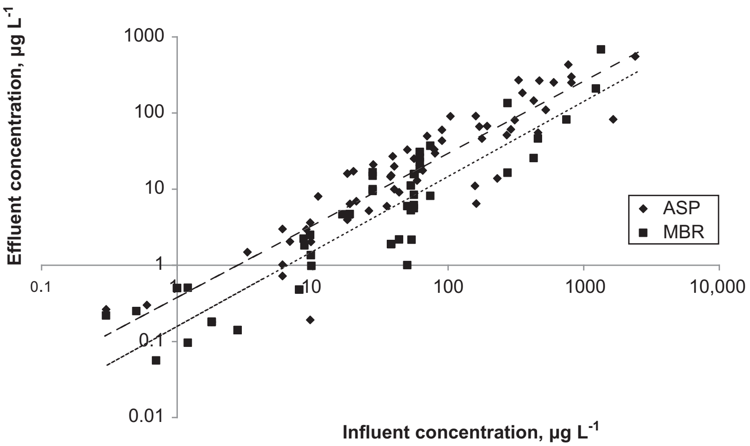

FIG. 2.37 Effluent vs influent metals concentration, ASP vs MBR, municipal wastewater treatment; data for seven transitional metals (Cd, Cr, Cu, Hg, Pb, Ni, Zn).

report an increase in removal with SRT, the increase is generally small and varies between studies (Santos & Judd, 2010).

Organics

Since organic micropollutants can potentially be biodegraded, some impact from retention time may be expected for these species. A recent review of available literature for PPCPs (Sipma et al., 2010) has revealed that:

- readily removed pharmaceuticals (acetaminophen, ibuprofen and paroxetine) are equally well removed in ASPs and MBRs;
- for a few moderately removed species (sotalol and hydrochlorothiazide) and highly intractable species such as carbamazepine and some macrolides (Bernhardt, Müller, & Knepper, 2006; Göbel, McArdell, Joss, Siegrist, & Giger, 2007), which are actually promoted by biotreatment through various physico-biochemical processes, there is no appreciable difference in performance between the ASP and the MBR; and
- for many other PPCPs removal tends to be greater for the MBR, but not significantly so.

Some authors have reported little appreciable difference in the performance of an MBR over that of the ASP for removal of many pharmaceutical products (Clara, Strenn, Ausserleitner, & Kreuzinger, 2004; Cirja, Ivashechkin, Schäffer, & Corvini, 2008). However, significant improvements have been reported for poorly biodegradable persistent polar pollutants, such as diclofenac, mecoprop and sulfophenylcarboxylates (Bernard et al., 2006), chemically complex compounds such as ketoprofen and naproxen (Kimura, Hara, & Watanabe, 2005) and others such as the anti-depressant fluoxetine (Radjenovic, Petrovic, & Barceló, 2007, 2009). This has been primarily attributed to the longer SRTs attainable by the MBR (Strenn, Clara, Gans, & Kreuzinger, 2004; Lesjean et al., 2005a; Clara, Kreuzinger, Strenn, Gans, & Kroiss, 2005), and studies conducted at very long SRTs, and commensurately low *F*/*M* ratios, have yielded better removals for some species (Göbel et al., 2007). Studies have shown similar removals for CASPs and MBRs for some species when both processes have been operated at the same SRT (Clara, Strenn, Ausserleitner, & Kreuzinger, 2004). However, several investigators have found no clear correlation between SRT and biodegradation of pharmaceuticals (Lishman et al., 2006; Vieno, Tuhkanen, & Kronberg, 2007; Zhang, Geißen, & Gal, 2008b), since for some compounds such as diclofenac and 17a-ethinylestradiole (Clara, 2005) other operating parameters appear to be important; little additional removal is attained at SRTs greater than 30 days (Suarez, Carballa, Omil, & Lema, 2008).

Other parameters have also been studied. Whilst both hydraulic residence time (HRT) and operating temperature may be expected to impact on removal, there appears to be little evidence of this in practice. Even with the associated widely fluctuating loadings, the performance of decentralized plants has been shown to be similar to those of larger centralized systems (Abegglen et al., 2009). Evidence suggests that degradation of some EDCs present at ng/L concentrations follows pseudo-first-order kinetics (Joss et al., 2006), such that their per cent removal is independent of concentration but highly dependent on residence time if the reactor configuration is plug flow. This appears to have been corroborated to some extent by studies of several acidic pharmaceuticals at a decreased pH (Urase, Kagawa, & Kikuta, 2005). pH is thought to influence removal according to the micropollutant pK_a value, the acid dissociation constant, since this would then affect its affinity for the largely hydrophobic sludge solids (Cirja, Ivashechkin, Schäffer, & Corvini, 2008). An interesting correlation has been reported between pharmaceuticals removal and nitrification (Perez, Eichhorn, & Aga, 2005; Batt, Kim, & Aga, 2006); nitrifying bacteria, and enzymes such as ammonium monooxygenase specifically (Berthe-Corti & Fetzner, 2002), have been postulated as being capable of co-metabolizing a wide range of refractory organic micro-pollutants. There is also evidence of impacts of the presence or depletion of C and N (Drillia et al., 2005) and oxygen (Zwiener & Frimmel, 2003), with significantly greater degradation of diclofenac demonstrated under anoxic compared with aerobic conditions.

Nutrients

Removal of nutrients (N and P) by both MBRs and CASPs is largely determined by the bioreactor design and operating conditions (HRT, SRT and redox), and the influent characteristics (i.e. COD/TKN ratio, alkalinity and temperature), as detailed in Section 2.2.4. Removal of both the nitrate formed through nitrification and phosphate present in the sewage then demands configuring the process to produce zones which are depleted (anoxic or anaerobic) and/or enriched (aerobic) with dissolved oxygen. Biological removal of over 90% for both N and P is generally achievable for an appropriately designed process, but other constraints imposed on the process design and by the nature of the carbon source generally limit removal to between 70 and 90% for many biological nutrient removal (BNR) plants. Residual P removal may be effected by chemical dosing to precipitate the phosphate salt. As with metals, membrane separation offers only a modest improvement in water quality with respect to phosphate over that attained by the classical BNR process based on gravitational separation.

2.3.10.3. Prognosis

There is little to suggest that existing MBR designs can routinely produce significantly higher permeate water quality with respect to the key pollutants identified above. Moreover, the high energy demand of MBRs makes it difficult to justify employing the technology on this basis. On the other hand, evidence from conventional anaerobic systems supports the notion that breakdown of alkylphenolic carboxylates is encouraged under anaerobic conditions (Minamiyama, Ochi, & Suzuki, 2006). Moreover, the addition of supplements for targeting micropollutants specifically is relatively unexplored, though such an approach is not thermodynamically favoured at the low concentrations concerned.

2.4. SUMMARY

Membrane separation processes applied to MBRs have conventionally been limited to MF and UF for separation of the permeate product from the bioreactor MLSS. Other processes, in which the membrane is used to support a biomass and facilitate gas transfer into the biofilm (Sections 2.3.2−2.3.3) or extract ions (Sections 2.3.3.1 and 2.3.3.2) or water, through transmembrane osmotic or vapour pressure difference (Section 2.3.3.5), have not reached the commercial stage of development. Membrane module configurations employed for biomass separation MBRs are limited to FS and HF for immersed processes (where the membrane is placed in the tank), and mainly MT (where it is placed outside the biotank). The latter provide shear mainly through pumping, although more recent processes also employ air-lift, whereas immersed processes rely entirely upon aeration to provide air-lift and shear. Shear enhancement is critical in promoting permeate flux through the membrane and suppressing membrane fouling and clogging, but generating shear also demands energy.

A considerable amount of research has been devoted to the study of membrane fouling phenomena in MBRs, and the trends of those studies published in the mid-noughties have been summarized by Meng et al. (2009) (Table 2.18). There is a general consensus that fouling constituents originate from the clarified biomass (Section 3.3.6.2). Many authors who have employed standard chemical analysis on this fraction have identified the carbohydrate fraction of

TABLE 2.18 Membrane Fouling Trends against Sludge and O&M Factors (Adapted from Meng et al., 2009)

Property	Effect on Membrane Fouling	Ref.
Sludge		
MLSS	MLSS↑ → normalized permeability↓	Trussell et al. (2007)
	MLSS↑ → fouling potential↑	Psoch and Schiewer (2006)
	MLSS↑ → Cake resistance↑, specific cake resistance↓	Chang et al. (2005)
Viscosity	MLSS/viscosity↑ → membrane permeability↓	Trussell et al. (2007)
	Viscosity↑ → membrane resistance↑	Chae, Ahn, Kang, and Shin (2006)
F/M	*F/M*↑ → fouling rates↑	Trussell, Merlo, Hermanowicz, and Jenkins (2006)
	MLSS (2–3 g/L): *F/M*↑ → irreversible fouling↑	Watanabe, Kimura, and Itonaga (2006)
	MLSS (8–12 g/L): *F/M*↑ → reversible fouling↑	Watanabe et al. (2006)
	F/M↑ → Protein in foulants↑	Kimura et al. (2005, 2008)
EPS	Polysaccharide↑ → fouling rate↑	Drews et al. 2006
	Bound EPS↑ → specific cake resistance↑	Cho et al. 2005a
	Polysaccharide↑ → fouling rate↑	Lesjean et al. (2005a,b)
	Bound EPS↑ → membrane resistance↑	Chae et al. (2006)
	(Loosely) bound EPS↑ → membrane fouling↑	Ramesh et al. (2006); Ramesh, Lee, and Lai (2007)
SMP	SMP is more responsible for fouling than MLSS	Zhang et al. (2006a)
	Colloidal TOC relates with permeate flux	Fan et al. (2006)
	Filtration resistance is determined by SMP	Jeong, Cha, Yoo, and Kim (2007)
	SMP is probably responsible for fouling	Spérandio et al. (2005)

TABLE 2.18 Membrane Fouling Trends against Sludge and O&M Factors (Adapted from Meng et al., 2009)—cont'd

Property	Effect on Membrane Fouling	Ref.
	Polysaccharide is a possible indicator of fouling	Le-Clech et al. (2005b)
	SMP↓ → fouling index↓	Jang et al. (2006)
	Fouling rates correlate with SMP	Trussell et al. (2006)
	Impact of polysaccharide is SRT dependent	Drews et al. (2008)
Filamentous bacteria	Filamentous bacteria↑ → sludge viscosity↑	Meng, Shi, Yang, and Zhang (2007)
	Bulking sludge can cause a severe fouling	Sun et al. (2008)
	Filamentous bacteria↓ → cake resistance↓	Kim and Jang (2006)
O&M		
SRT	SRT decrease from 100 to 20 d → TMP↑	Ahmed et al. (2007)
	SRT decrease from 30 to 10 d → fouling↑	Zhang et al. (2006b)
	SRT↑ → fouling potential of SMP↑	Liang, Liu, and Song (2007)
	SRT decrease from 5 to 3 d → fouling↑	Ng et al. (2006)
HRT	HRT↓ → membrane fouling↑	Meng et al. 2007
	HRT↓ → membrane fouling↑	Chae et al. (2006)
	HRT↓ → membrane fouling↑	Cho et al. (2005a)
Aeration	Aeration intensity↑ → permeability↑	Trussell et al. (2007)
	Air sparging improves membrane flux	Psoch and Schiewer (2006)
	Larger bubbles for fouling control are preferable	Phattaranawik, Fane, Pasquier, and Bing (2007)
	Air backwashing for fouling control is preferable	Chae et al. (2006)

(*Continued*)

TABLE 2.18 Membrane Fouling Trends against Sludge and O&M Factors (Adapted from Meng et al., 2009)—cont'd

Property	Effect on Membrane Fouling	Ref.
	Bubble-induced shear reduces fouling	Wicaksana et al. (2006)
	Air scouring can prolong membrane operation	Sofia et al. (2004)
Flux	Sub-critical flux mitigates irreversible fouling	Lebegue, Heran, and Grasmick (2008)
	Sub-critical flux mitigates fouling	Guo, Vigneswaran, Ngo, and Xing (2007)

the SMPs (SMPc) arising from the bacterial cells as being mainly responsible for fouling, rather than suspended solid materials. On the other hand, there is evidence to suggest – particularly from anMBRs – that fouling can be attributed to colloidal materials per se, regardless of their chemistry (Sections 2.3.6.6 and 2.3.7.4). In any event, attempts to predict fouling rates by EPS/SMP levels have not translated well across different plants or studies since biomass characteristics vary significantly from one plant to another. Moreover, achieving a consensus on the relative contributions of candidate foulants to membrane fouling is constrained by the different analytical methodologies and instruments employed.

The greater proportion of the research into the fundamental mechanisms of MBR fouling was conducted between 1995 and 2005. Although a good understanding of the effect of single biological/operating parameters on the hydraulic performances of the system has been obtained, the complex interactions between those concepts make simple modelling and prediction of MBR fouling challenging. Since 2005, the majority of the studies have tended to focus on the optimization of the operating conditions (e.g. use of additives), and improved characterization of the fouling layer and of the biomass, and not just the SMP. It has been now recognized that ex situ measurements made on the SMP may be of limited value in characterizing fouling, and that the use of in situ filtration methods may provide better information on fouling propensity. Given that the fouling layer is not solely composed of SMP species, further characterization of its nature (and that of the irrecoverable fraction in particular) may help to improve long-term MBR performance and membrane life.

There are also cross-disciplinary issues in the area of membrane fouling. There appears to be little interconnection between foulant analysis in wastewater and potable applications, and between membrane cleaning in the industrial process and municipal water and wastewater sectors. Studies in the potable area tend to point to colloidal materials and Ca-organic carboxylate complexation as being the two key foulant types, and this may apply as much to wastewater as potable water membrane applications. Within the municipal sector, the number of studies devoted to characterization of foulants vastly exceeds that for optimizing chemical cleaning (Section 1.5), notwithstanding the fact that it is the latter which controls irrecoverable fouling and so, ultimately, membrane life. Membrane cleaning in industrial process water applications, however, is rather more advanced — dating back to the 1980s — with protocols arguably developed on a more scientific basis than those in the municipal sector.

Over the past 5 years or so, the issue of fouling in municipal applications in particular has diminished in relative importance from the practitioner's perspective (Section 3.6.1). This is in part due to the better understanding of the fouling phenomena developed by the large number of research studies based on the topic, but primarily due to the ever-increasing amount of heuristic information from practitioners regarding operation and maintenance (O&M). Specifically, in practice MBR technology is constrained mainly by the macroscopic constituents of design and operation, for example the homogeneous distribution of membrane air and appropriate chemical cleaning protocol respectively, rather than microscopic aspects such as foulant chemical speciation. Nonetheless, there remains an ever diminishing but still significant energy penalty associated with MBR operation — ostensibly relating to fouling — which engages practitioners and academics.

With regard to O&M of full-scale plant, it is the dynamic effect which exerts the greatest influence on MBR performance, ultimately leading to equipment and/or consent (i.e. target product water quantity/quality) failures. Specifications for full-scale MBR installations are generally based on conservative estimates of hydraulic and organic (and/or ammoniacal) loading. However, in reality, these parameters fluctuate significantly. Moreover, even more significant and potentially catastrophic deterioration in performance can arise through equipment malfunction and operator error, leading to:

- decreases in the MLSS concentration (either through loss of solids by foaming or by dilution with feedwater),
- foaming problems, sometimes associated with the above,
- loss of aeration (through control equipment malfunction or aerator port clogging) and
- loss of permeability (through misapplication of backflush and cleaning protocol, hydraulic shocks or contamination of the feed with some unexpected component, or clogging generally).

Clogging can arise both in the membrane channels and in the aerator ports, in both cases impacting deleteriously on flux distribution and thus fouling rates. In this respect, developing methods of ensuring homogeneity of air distribution can advance both fouling and clogging control. Once again, though, the number of MBR papers published which have been devoted specifically to aerator design are limited.

It is perhaps unsurprising that the areas of irrecoverable fouling and clogging have attracted limited attention by the academic community. These phenomena can only be studied over extended time periods, which is not conducive to academic research. Research into fouling characterization is likely to continue for some time and new membranes and systems are being developed from research programmes globally, primarily devoted to decreasing costs or carbon footprint generally and energy demand in particular. However, much practical information can be obtained from the examination of pilot- and full-scale plant data (Chapters 3 and 5), and it is also instructive and expedient to consider the attributes of existing individual commercial technologies (Chapter 4).

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Chapter 3

Design, Operation and Maintenance

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3.1. MBR SYSTEM OPERATING PARAMETERS

There are essentially three main operations of a membrane bioreactor (MBR) contributing most significantly to operating expenditure (OPEX). These are the following:

(a) membrane permeability maintenance,
(b) microbiology maintenance and

(c) liquid and sludge transfer.

Of these, maintaining membrane permeability is the most significant, and impacts on OPEX through:

i. scouring and/or agitation by aeration (for immersed systems) and/or liquid crossflow (for sidestream systems);
ii. cleaning, both physical (relaxation and/or backflushing) and chemical (maintenance and/or recovery); and
iii. membrane replacement, should irreparable damage be sustained or otherwise recovery cleaning prove ineffective.

Since microbiology is also maintained by aeration − both for suspending the biomass and maintaining dissolved oxygen levels for sustaining microbiological activity − it follows that aeration energy is the most significant contributor to OPEX for immersed systems. Design of an iMBR therefore demands knowledge both of the feedwater quality, which principally determines the oxygen demand for biotreatment, and the aeration demand for fouling control, which relates to a number of system characteristics as summarized in Fig. 2.27. Whereas existing established biochemical/biokinetic expressions (Section 2.2.4) can be used to design the biological component of an MBR, aided by available tools for this purpose, the key interrelationships between flux, permeability and scouring (by aeration or liquid crossflow) for the membrane component can currently only be arrived at through heuristic data or empirical study. Similarly, the frequency of cleaning, which incurs downtime and thus decreases the overall flux, can only be determined in the same way.

The various governing principles of MBR membrane filtration have been detailed in Section 2.3.6.3, and key membrane design and operating parameters are summarized and defined in Table 3.1. The basic expressions can be temperature corrected. Correction of flux for permeate temperature T is classically based on the viscosity correction factor $1.024^{(T\text{-}20)}$, and correction of air flow rate for both air temperature (T_a in °C) and blower inlet pressure ($P_{A,1}$ in bar) by the general gas equation correction factor $(293/(273 + T_a))(P_{a,in}/1.01)$. However, both of these are classical expressions. Whilst correction of air flow for temperature and pressure may be reasonable, the temperature cannot necessarily be assumed to be that of the ambient air temperature. In the case of normalization of the flux, evidence suggests that simple viscosity correction is insufficiently rigorous (Section 2.3.6.3).

3.2. IMMERSED MBR MEMBRANE REFERENCE DATA

3.2.1. Pilot Plant Trials

Since MBR performance is highly dependent on feedwater quality, true comparison of the performance of different MBR technologies can only be

TABLE 3.1 Key Membrane Design and Operating Parameters

Raw Data		Normalized/derived data	
Permeate, feed flow (m^3/h)	Q_p, Q_f	Flux, J ($L\,m^{-2}\,h^{-1}$, LMH)	Q_p/A_m
Membrane area, m^2	A_m	Permeability, K (LMH/bar)	$J/\Delta P_m$
Mean transmembrane pressure, TMP (bar)	ΔP_m	Membrane aeration demand per unit membrane area, SAD_m ($Nm^3/(m^2\,h)$)	Q_A/A_m
Membrane aeration rate (m^3/h)	$Q_{A,m}$	Membrane aeration demand per unit permeate flow, SAD_p	$1000 \times SAD_m/J$ or $Q_{A,m}/Q_p$
Physical cleaning (backflush) interval (h)	t_p	Number of physical cleaning cycles per chemical clean, n	$\frac{t_c}{t_p+\tau_p}$
Physical cleaning (backflush) duration (h)	τ_p	Net flux, J_{net} (LMH)	$n\frac{(Jt_b-J_b\tau_b)}{t_c+\tau_c}$
Backflush flux, LMH	J_b		
Chemical cleaning interval (h)	t_c		
Chemical cleaning duration (h)	τ_c		

achieved when they are tested against the same feedwater matrix. Two options exist for such comparisons:

(a) the use of analogues of precisely controlled composition; and
(b) the simultaneous trialling of technologies challenged with the same feedwater.

The dichotomy over the respective benefits of analogue and real matrices for research and development (R&D) purposes is not limited to MBRs, and applies across many sectors. The use of analogues for feedwaters in water and wastewater treatment R&D allows total control of feed quality as well as permitting the convenience of testing sequentially or simultaneously without detracting from the performance comparison. On the other hand, it is widely recognized that analogues can never satisfactorily represent real sewage, particularly so in the case of such a crucially important parameter as fouling propensity, and can be extremely expensive to produce. For pilot trials of MBR technologies of a reasonable scale (i.e. based on a small number of full-scale membrane modules), conducting trials based on real feedwaters is always preferred. A number of such trials have been carried out since around the turn of the millennium which permit a useful technology comparison (Table 3.2), albeit with certain caveats. The studies identified in the table have all employed at least one full-scale membrane module per bioreactor and at least three

TABLE 3.2 Comparative Pilot Plant Trials

Technology Tested	Reference				
	Point Loma, MWH*	Beverwijk, DHV**	Bedok/Ulu Pandan PUB#	Pietramurata, Univ Trento†	Kloten/ Opfikon Eawag
Zenon	X	X	X	X	X
Kubota	X	X	X	X	X
MRE‡	X	X	X		X
Norit	X	X			
Huber	X	X			
Memcor	X			X	
Toray	X	X			
Asahi Kasei	X		X		
Puron	X				

*Adham et al. (2004); DeCarolis et al. (2009); Hirani et al. (2010).
**van der Roest et al. (2002); Lawrence et al. (2005).
#Tao et al. (2005); Oda et al. (2009); Qin et al. (2009).
†Guglielmi et al. (2007); Verrecht et al. (2008).
‡Mitsubishi Rayon Engineering.

different technologies. A number of pilot trials of individual technologies have also been conducted. An exhaustive appraisal of these is not possible, and these have generally not been comparative.

In the subsequent sections the data from each of these trials are tabulated, with derived (i.e. calculated) parameters indicated.

3.2.1.1. Beverwijk WwTP, the Netherlands

A pioneering comparative pilot trial was carried out at Beverwijk–Zaanstreek wastewater treatment works between 2000 and 2004, with a substantial body of work published in 2002 (van der Roest, Lawrence, & van Bentem, 2002). The work represents one of the earliest large-scale comparative pilot trials and was conducted by DHV in collaboration with the Dutch Foundation of Applied Water Research (STOWA). The ultimate goal of the work was the implementation of a number of full-scale plants of 60–240 megalitres per day (MLD) capacity in the Netherlands. Results from trials on four MBR systems of 24–120 m^3/day capacity were published in the 2002 report, with the four technologies originally tested being Kubota, Norit X-Flow (air-lift sidestream), Mitsubishi Rayon and Zenon. Subsequent reports by this group (Schyns, Petri, van Bentem, & Kox, 2003; Lawrence, Ruiken, Piron, Kiestra, & Schemen, 2005) have not contained the same level of technical detail regarding operation and maintenance.

The reported trials (van der Roest et al., 2002) were conducted in four phases:

I. Primary clarification with ferric dosing prior to screening;
II. Primary clarification with simultaneous ferric dosing (i.e. downstream of screening);
III. Raw wastewater with simultaneous precipitation; and
IV. Raw wastewater with bio-P removal.

All MBRs were configured with a denitrification zone, with an anaerobic tank added for Phase IV. The Kubota system was initially fitted with a single deck; a double deck was fitted mid-way through Phase II. Data presented in the following tables refer only to the first three phases. Water quality data (Table 3.3) relate to the range of mean values (i.e. not the total range of values recorded) reported for the four technologies tested. The process design and pre-treatment (i.e. screening) were all as specified by the supplier. A noticeable impact on α-factor was recorded independent of the membrane type, with the value decreasing from 0.78 to 0.79 for the conventional activated sludge process (CASP) in operation at the works to 0.43−0.54 across all the MBR technologies.

Design (Table 3.4) and operation and maintenance (Table 3.5) data are reported below. Comparison of process performance is somewhat difficult from the reported data, since the process operating conditions were changed frequently throughout the trials. Variation in hydraulic retention time (HRT) meant that the organic loading rate varied significantly between the different technologies tested, from 0.043 to 0.059 kg biological oxygen demand (BOD)/m^3 for the Mitsubishi Rayon module to 0.075−0.11 for the Zenon. Ferric dosing was applied at different concentrations for the different technologies.

The protocol adopted was for operation under flow conditions set by the flow to the sewage treatment works; a fixed proportion of the flow was directed

TABLE 3.3 Feedwater Quality, Beverwijk Trials

Parameter, mg/L	Raw Sewage	Advanced Primary Treated Sewage
COD	548−621	297−422
TKN	57−61	39−67
Total phosphate-P	9.3−12.1	7.1−8.3

TABLE 3.4 Design Information, Beverwijk Trials

Parameter	Kubota	MRE SUR	Zenon	Norit X-Flow
Flow (m^3/day)				
Peak	190–240	154–230		30–43
Normal (design)	48–72	32–48	44–74	9
Screening	2.0 mm basket filter, SD: 2.5–1 mm rotary drum filter, DD	0.75 mm parabolic sieve	0.75 mm static half drum with brush	0.5 mm rotating drum
Tank sizes (m^3)				
Denitrification	12.0	15.6	12.0	3.6
Nitrification	–	7.8	7.7	2.1
Membrane	18.8	10.8	3.9	~0.1 (a-ls)
Recycle ratio	~8	~12	~8	~5
Membrane				
Type	*510*	*SUR 234*	*ZW500, a* or *c*	*F-4385*
Configuration	150 panels per deck, SD/DD	Triple deck	4 modules	8 modules, 2 × 4
Total membrane area, m^2	240	315	184	240

SD = Single deck; DD = double deck; and a-ls = air-lift sidestream (no membrane tank).

to the MBR pilot plants. This led to rather conservative average fluxes. Operating (or gross) fluxes ranged between 10 and 20 LMH for most of the time, leading to even lower net fluxes of 8–15 LMH. As a result, high permeabilities were generally recorded, particularly for the Kubota membrane and especially following the introduction of cleaning strategies such as relaxation. Commensurately high SAD numbers ($SAD_p = 60–90$, m^3 air/m^3 permeate in the case of Kubota) resulted. Peak flow tests conducted on each of the technologies always produced a significant drop in permeability.

For each of the trials operational problems were encountered which led to irreversible, and sometimes irrecoverable, fouling. This included partial aerator blockage, membrane tube blockage and partial dewatering (Section 3.6.2). Repeated chemical cleans were thus employed over the course of the trials, for both maintenance and recovery of membrane permeability, and relaxation was only introduced mid-way through the trials. A strong temperature dependence

TABLE 3.5 Operation and Maintenance (O&M) Data, Beverwijk Trials

Parameter	Kubota	MRE SUR	Zenon	Norit X-Flow
Membrane aeration capacity, Nm^3/h	90–180	75–120	100 (cycled)	140
Cycle	8 min on 2, r*	20 h on/ 4 h[#]	backflushed[¶]	20 h on/ 4 h[††]
Net flux, LMH				
Normal	8.3–12.5	5–8	20[‡]	15–20, 37[‡‡]
Peak	32.5–42	20.3–30.6	35[‡]	50
Biological aeration capacity, Nm^3/h	160	160	100	140
F/M ratio	0.04–0.18	0.02–0.14	0.04–0.18	0.04–0.12
HRT, h	10.2–15.4	15–22	7.6–12.3	15.2
SRT, d	27–70	31–87	26–51	42–66
MLSS, g/L	10.5–12	8.9–11.6	10.4–11.2	
Chemical cleaning reagents	NaOCl, 0.5% Oxalic acid, 1%	NaOCl, 0.5%, followed by acid	NaOCl, 1%, followed by 0.3% citric acid	NaOCl, 0.5%, followed by *Ultrasil*[$]
Derived data				
SAD_m*, $Nm^3/(m^2\ h)$	0.75	0.28–0.38	0.54[§]	0.33–0.6[††]
SAD_p, m^3 air/m^3 permeate	60–90 normal; 18–23 peak	48–56 normal; 12–14 peak	27; 15	30–40, 16 opt; 12–16 peak
Mean permeability *K*, LMH/bar	200–250 w/o. r* 500–800 w. r* 350 peak w. r*	200 normal[†] 140–150 peak[†]	200–250 320–350 after clean	250 normal 75–200 peak
Permeability decline, $\Delta K/\Delta t$, LMH/(bar h)	1.5**	0.39**	20**	

Relaxation (r) introduced mid-way through Phase I; permeability data refer to without (w/o.) and with (w.) relaxation.
**Refers to peak flux operation: for the Zenon membrane this was 60 LMH.*
[#]*Night-time relaxation introduced during Phase III, along with backflushing at 20 LMH.*
[†]*Assumed to be with relaxation.*
[‡]*Refers to 500c module.*
[¶]*Authors state 'ratio of net to gross flux was 83–85%'.*
[§]*Intermittent operation.*
[††]*Night-time relaxation introduced during Phase II.*
[‡‡]*With weekly maintenance clean.*
[$]*Combination of sulphuric and phosphoric acid.*

of permeability was noted for all membranes tested at low temperature, with the sustainable permeability decreasing markedly at temperatures below 10 °C.

It was concluded from this trial that the Kubota plant could not be operated routinely at net fluxes of 38.5 LMH, although the fouling rate was decreased when relaxation was introduced. Peak flux operation of both the Zenon and Kubota systems at 41–43 LMH for 100 h led to a manageable decline in permeability however, with permeability recovering to normal levels on reverting to low flux operation. Under optimum conditions of regular maintenance cleaning and backflushing at 20.8 LMH, the permeability of the Mitsubishi Rayon Engineering (MRE) membrane was maintained at between 200 and 300 LMH/bar at some unspecified flux. A net flux of 20.8 LMH was achieved for a period of 5 days, but peak fluxes of 28 LMH or more led to a rapid decline in permeability. Enhanced cleaning at higher reagent concentrations (1 wt% NaOCl) and temperatures (40 °C) may have led to the noted deterioration of the membranes. The Norit X-Flow membrane was eventually successfully operated at 37 LMH with weekly maintenance cleaning and night-time relaxation (4 h every 24 h). However, there were problems maintaining this flux due to tube blockage. The Zenon plant, fitted with the 500c module, appears to have been operated under conditions whereby a reasonable net flux of around 20 LMH with an accompanying permeability of 200–250 LMH was sustained through maintenance cleaning twice weekly over an extended time period. It is possible that the other technologies could also have been sustained at higher fluxes if operated under more optimal conditions.

3.2.1.2. Point Loma WwTP, San Diego

This trial resulted from an award made to the City of San Diego and Montgomery Watson Harza (MWH) from the Bureau of Reclamation to evaluate MBR + reverse osmosis (RO) technology for wastewater reclamation (Adham, DeCarolis, & Pearce, 2004). Originally a 16-month study was conducted at the Point Loma Wastewater Treatment Plant (PLWTP) at San Diego, CA on four MBR membrane products and technologies: Kubota, US Filter (now Siemens), Mitsubishi Rayon and Zenon (now GE). A subsequent trial (DeCarolis, Hirani, & Adham, 2009; Hirani, DeCarolis, Adham, & Jacangelo, 2010) was conducted based on four more membranes: Huber, KMS Puron, Norit (configured as an air-lift sidestream) and Toray. The technology providers for some of these products were Enviroquip (now Ovivo) for Kubota, Kruger for Toray and Dynatec for the air-lift multitube system based on Norit membranes (*Dyna-Lift*).

In the first trial (Adham et al., 2004) the technologies were each challenged first with raw wastewater (Part 1) and then with advanced primary treated effluent containing polymer and coagulant residual (Part 2), arising from clarification with ferric chloride (27 mg/L, average dose) and a long chain, high-molecular-weight anionic polymer. Details of feedwater quality

(Table 3.6), MBR technology design (Table 3.7) and operation (Table 3.8) are given below.

In this trial only the advanced primary treated effluent feedwater was tested on all four technologies. Results from untreated primary sewage were inconclusive since the feed evidently contained unclarified ferric coagulant. All four technologies performed satisfactorily with respect to nitrification, disinfection and BOD removal. Only the Kubota was operated with denitrification and consequently demanded less oxygen from fine bubble aeration than the other technologies (Table 3.8). This being said, the data for specific aeration demand appear to demonstrate the following, for this trial:

(a) The MRE system had a significantly higher specific aeration demand with respect to membrane aeration (SAD_m and SAD_p), reflecting the somewhat lower permeability of the membrane material.

(b) There was little difference in the figures for the total specific aeration demand ($TSAD_p$) of all those technologies operating without denitrification under optimum conditions, the figures ranging from 51 to 62.

The latter point is interesting in that it suggests one of two things: (a) that the air from fine bubble aeration somehow ameliorated membrane fouling or (b) that the coarse bubble aeration contributed substantially to the dissolved oxygen (DO).

The permeability data also provide interesting information on system operation and maintenance. The Zenon system was the only one operated with regular (three times a week) maintenance cleans by backpulsing with chlorine (250 ppm) or citric acid (2%) four times with a 30-s soak time between each cycle. This incurred downtime but allowed the system to operate at a relatively high net flux (37.2 LMH, significantly higher than the more commonly applied

TABLE 3.6 Feedwater Quality, Point Loma Trials

	Adham et al. (2004)		Hirani et al. (2010)
Parameter, mg/L	Raw Sewage	Pre-treated Sewage	Raw Sewage
Ammonia-N	27.3	26.6	23
BOD_5*	213	97	155
COD	463	216	376
TOC	40	44	58
TKN	42.9	44.8	–
Orthophosphate-P	0.61	0.46	0.076

*5-Day BOD.

TABLE 3.7 Design information, First Trial (Adapted from Adham et al., 2004)

Component	Kubota	US Filter#	Zenon	MRE
Screening				
	3.2-mm travelling band screen	1.0-mm wedgewire slotted rotary screen.	0.75-mm perforated rotary drum screen	0.75-mm perforated rotary drum screen
Tank sizes, m^3				
Denitrification	6.42	3.79*	–	–
Nitrification	12.5	5.7	4.92	6.06
Membrane	–	0.34	0.7	0.95
Recycle ratio	4 (303 L/min)	–	–	–
Membrane				
Type	*510*	*B10 R*	*ZW500d*	*Sterapore SUR*
Configuration	100 panels per deck, double deck	4 modules	3 modules	50 modules per bank, 2 banks
Total membrane area, m^2	160	37	69	100

**Not used in trial.*
#Now Siemens.

value of 25 LMH). The projected cleaning cycle time, calculated from the permeability decline value and transmembrane pressure (TMP) boundary values of 0.1 and 0.5 bar, was similar for the three HF technologies. The cleaning cycle time of the Kubota technology, on the other hand, could not be predicted since, under the conditions employed, there was no noticeable permeability decline. It is possible that the inclusion of denitrification for this process treatment scheme may have ameliorated fouling in some way. Cleaning protocols employed are summarized in Table 3.9.

The MBR effluent was very low in turbidity (<0.1 NTU on average, cf. 36–210 NTU for the feedwater) across all technologies tested, and average effluent BOD, total organic carbon (TOC) and chemical oxygen demand (COD) concentrations were, respectively, <2 mg/L, <9 mg/L and <31 mg/L. Silt density index (SDI) values measured for the Kubota MBR effluent during Phase I ranged from 0.9 to 1.1. Effluent from the Kubota MBR was treated downstream using Saehan BL and Hydranautics LFC3 reverse osmosis membranes, which were operated with minimal fouling when challenged with MBR permeate from either raw wastewater or advanced primary effluent. The

TABLE 3.8 O&M Data, First Trial (Adapted from Adham et al., 2004)

Parameter	Kubota	US Filter *Memjet*	GE Zenon	MRE *SUR*
Membrane aeration rate, Nm^3/h	96	14.4	36[#]	90–114[‡]
Cycle, min	9 on/1 relax	12 on/0.75 relax/ 0.25 backflush	10 on/0.5 relax	12 on/2 relax
Net flux, LMH	25	19.2–24.2, pt 1 24.2[‡]–40, pt 2	37.2[†]	20–25[‡]
Fine bubble aeration rate, Nm^3/h	<18	42, pt 1 78, pt 2	96	17–27[‡]
Target (actual) DO, mg/L	2 (3–5)	>1 (2–4)	1 (0.5–1.5)	>1 (1–2)
HRT, h	5.1	6–11.5, pt 1 6, pt 2	2–5	3–4
SRT, d	10–40	12–25	15–20	25–40
MLSS, g/L	10–14, pt 1 9–12, pt 2	8–10, pt 1 6–8, pt 2	8–10	8–14
Derived data				
SAD_m*, $Nm^3/(h\,m^2)$	0.60	0.39	0.52	0.72[‡]
SAD_p^*, m^3 air/m^3 permeate	24	16–20, pt 1 9.8[‡]–16, pt 2	14[‡]	45
$TSAD_p$, m^3 tot. air/m^3 permeate	<29	63–79, pt 1 62[‡]–103, pt 2	51	52–66[‡]
Mean permeability[‡] K, LMH/bar	250	150	270	140
Permeability decline[‡], $\Delta K/\Delta t$, LMH/(bar h)	0	0.12	0.21	0.17
Cleaning cycle time**, d	–	67	59	49

pt 1 = Part 1: raw sewage containing ferric matter; pt 2 = part 2: advanced primary treated (clarified) sewage.
Specific aeration demand.
***Based on a cycle between 0.1 and 0.5 bar at the given flux J and permeability decline $\Delta K/\Delta t$, hence $t_c = 8J/(\Delta K/\Delta t)$.*
[#]*Intermittent: 10 s on, 10 s off.*
[†]*Sustained by maintenance cleaning by backpulsing w. 250 ppm NaOCl applied three times a week.*
[‡]*Data refer to that recorded for the latter part of the trial: advanced primary feed, optimal conditions.*

TABLE 3.9 Cleaning Protocols for the Four MBR Technologies, First Trial (Adapted from Adham et al., 2004)

Clean Type	Kubota	MRE SUR	Zenon	USF Memjet
Recovery				
Reagents used	NaOCl, 0.5%; oxalic acid, 1%	NaOCl, 0.3%, citric acid, 0.2%	NaOCl, 0.2%, citric acid, 0.2–0.3% (pH 2–3)	NaOCl, 100 mg/L as Cl_2
Protocol	CIP by soaking for 2 h; reagents applied consecutively	CIP by flushing through for 2 h and soaking for 2 more hours	Tank drained and membranes hosed down with water; membranes backpulsed with reagent until flux stable	Tank drained and membranes rinsed with water for 10 min; membranes flushed through with reagent then soaked for 4 min
Maintenance	Reagent		Concentration	Duration
Zenon	NaOCl	250 mg/L	10–15 s × 3 backpulses at 55 LMH, 30 s relaxation between pulses	

average net operating pressure of the Saehan 4040 BL (low pressure) RO membranes measured during testing was 3.1 bar, and that of the Hydranautics LFC3 (fouling resistant) RO membranes measured during testing was 8.3 bar. A 1–2 mg/L dose of chloramine in the RO feed was effective in mitigating membrane biofouling.

In the subsequent trial the sewage was of similar quality, though lower in P concentrations (Table 3.6), and encompassed four more technologies (Table 3.10). This second trial (DeCarolis et al., 2009; Hirani et al., 2010) included (a) a peak loading test and (b) a micro-organism rejection study, as well as steady-state studies of flux sustainability at the recommended aeration rates (Table 3.11) and chemical cleaning protocols (Table 3.12). The peak loading test involved increasing the flux by between 50 and 219% for three 2-h periods over a diurnal cycle across the range of technologies studied (Table 3.13). For some technologies, the peak loading was accompanied by ameliorative measures in the form of increased membrane aeration or decreased filtration cycle time. The rejection study encompassed both indigenous and seeded MS-2 coliphage.

Results for steady-state operation indicated SAD_m values of 0.34–0.74 with accompanying SAD_p values of 7.6–27, the lowest arising for the side-stream air-lift configured technology for which supplementary sludge pumping was employed. The peak loading trials produced a 22–32% decrease in permeability for the immersed technologies and no change for the air-lift sidestream, which was subjected to the lowest peak loading factor of 50%

TABLE 3.10 **Design Information, Second Trial (Adapted from Hirani et al., 2010)**

Component	Puron	Huber	Neosep/ Toray	Dyna-Lift/Norit
Screening	0.8 mm perforated rotary drum screen from Waste-Tech Inc.			
Tank Sizes, m^3				
Anoxic	2.19	NA	4.9	4.73
Aerobic	1.54	14	11.4	5.3
Membrane	0.7	12	7.19	–
Membrane recycle ratio	4	8.2	5	10.8
Membrane				
Type	*PSH 500C2*	*VRM*	*K100*	*38 PRV*
Configuration	Immersed hollow fibre	Immersed flat sheet	Immersed flat sheet	Sidestream tubular
Total membrane area, m^2	30	108	140	29

NA, Not applicable.

compared to 100–219% for the immersed systems (Table 3.13). In all cases the permeability was recovered following the 2-h peak loading period. The micro-organism rejection study revealed more variable performance against seeded coliphage than the indigenous species, with the latter consistently removed according to a log rejection value (LRV) of 3.2–3.4 with no correlation with either membrane pore size or flux. This result reflects the association of indigenous coliphage with particles (Section 2.3.10). Faecal and total coliforms were similarly consistently removed by 5.5–6.0 and 5.8–6.9 logs, respectively.

3.2.1.3. Bedok Water Reclamation Plant, Singapore

This trial arose from the Singapore NEWater project under the auspices of the Centre for Advanced Water Technology of the Singapore Utilities International Private Limited, a wholly owned subsidiary of Singapore Public Utilities Board. The work supports the country's NEWater programme, i.e. the production of high-grade water of potable quality from municipal effluent to ensure a diversified and sustainable water supply. The project began in 2000 with the installation of ultrafiltration/microfiltration (UF/MF) plants for polishing secondary sewage prior to RO treatment. The MBR/RO option has been explored with trials of three MBR pilot plants operating simultaneously. The

TABLE 3.11 O&M Data, Second Trial (Adapted from Hirani et al., 2010)

Parameter	Puron	Huber	Neosep/ Toray	Dyna-Lift/ Norit
Membrane aeration rate, Nm^3/h	10.3	51.6	102.9	10.12
Cycle, min	6 on/0.33 backflush	9 on/1 relax	9 on/1 relax	10 on/1 backflush
Net flux, LMH	22.6	25	27	45.7
HRT, h	4–11	8–15	5–7	7–11
Median SRT, d	13	15	20	33
MLSS, g/L	9–12	8–14	9–12	8–12
Cleaning cycle time, d	191	207	>920	332
Derived data				
Max permeability *K*, LMH/bar	340	250	390	420
SAD_m, $Nm^3/(m^2\ h)$	0.34	0.48	0.74	0.35*
SAD_p, m^3 air/m^3 permeate	15	19	27	7.6*

**Supplemented by sludge pumping at ~11× permeate flow rate.*

three technologies are unspecified in the report produced (Tao et al., 2005) but have membrane properties similar to those of the Kubota, Zenon and MRE products. The plants were commissioned in March 2003 and fed throughout with primary settled sewage having mean COD, BOD_5, ammonia, TKN and TP levels of 265, 99, 33, 33 and 9 mg/L, respectively.

The MBR plants are described in Table 3.14. The plants were all of around 75 m^3 total tank and all were submerged MBRs, two with a separate membrane tank. The baseline HRT and solids retention time (SRT) values were, respectively, 6 h (hence 300 m^3/day flow) and 21 days for all systems, and the target biotank MLSS was 10 g/L. An ex situ clean was only carried out once on MBR A and only due to failure of the aeration system. Following operation for 2–3 months under baseline conditions the flux was increased for an unspecified period and following this the aeration reduced for another unspecified period. It is not clear whether the conditions identified were optimal. According to the analysis conducted, MBR B (assumed to be MRE) was lower in energy demand than the other two MBRs. However, MBR B was also the only one configured without a separate aeration tank. Contrary to some previous reported data

TABLE 3.12 Cleaning protocols for the Four MBR Technologies, Second Trial (Adapted from Hirani et al., 2010)

Clean	Puron	Huber	Neosep/Toray	Dyna-Lift/Norit
Maintenance				
Reagents used	0.4% NaOCl, 0.2% citric acid	0.2% NaOCl, 0.2% citric acid	0.05% NaOCl, 0.05% citric acid	No maintenance cleaning
Protocol	Backflush with cleaning solution and soak, total duration ~30 min; reagents applied consecutively	Soak with NaOCl solution for ~2 h, if necessary soak with citric acid solution for 2 h	Soak with NaOCl solution for about 2 h, soak with citric acid solution for 2 h	
Recovery				
Reagents used	0.1% NaOCl, 0.25% citric acid	0.02% NaOCl, 0.02% citric acid	NA	0.02% NaOCl
Protocol	Drain membrane tank; soak membrane in cleaning solution for 4–6 h	Drain membrane tank; soak membrane in cleaning solution for 4–6 h	NA	Drain membrane modules; soak membranes in cleaning solution for 50 min, followed by citric acid if required

NA, not applicable.

(Fig. 2.20), these authors reported a linear relationship between viscosity and MLSS between 4 and 14 g/L for all three systems.

The mean product water quality from each of the MBRs tested was found to be broadly similar at <0.2 NTU, <2 mg/L TKN, <1 mg/L NH_4-N and <5 mg/L TOC, and slightly higher than product water from UF polishing of secondary effluent (7 mg/L TOC, 4.5 mg/L TKN and 3 mg/L NH_4-N). However, for MBRs A and B the permeate TOC was found to rise significantly – to as high as 100 mg/L – following chemical cleaning. As the authors suggested, this was likely to be due to the removal of the protective gel layer by chemical cleaning, this layer taking around 36 h of filtration to be reformed. TOC in the permeate may also arise from chemical cleaning with citric or oxalic acid.

Pilot testing has shown the MBR-RO option to produce a slightly superior quality product water than the conventional approach of secondary treatment followed by UF/MF + RO, specifically with respect to TOC, nitrate and ammonia (Qin et al., 2006), and also tends to be lower in cost. The work

TABLE 3.13 **Operational Changes during Peak Loading (Adapted from Hirani et al., 2010)**

Operating Parameter	Puron	Huber	Neosep/ Toray	Dyna-Lift/ Norit
From average to peak flux				
% Increase in flux	219	115	100	50
% Decrease in permeability	32	22	28	0
% Increase in SAD_m	63	0	0	0
% Increase in backflush flux	38	NA	NA	0
% Decrease in filtration cycle time	0	56	0	0
MLSS concentration and recirculation ratios (x Q) during peak flux				
Measured aeration tank MLSS concentration, mg/L	10,200	8200	10,000	11,500
Recirculation ratio, average flux operation	4.0	8.2	5.0	10.8
Recirculation ratio, peak flux operation	3.1	3.8	4.0	10.0
Membrane tank MLSS concentration	15,291	11,120	13,392	12,800

NA, not applicable.

preceded the construction of the 23-MLD MBR demonstration plant at Ulu Pandan based on GE Zenon technology, where membrane aeration demand has been further optimized (Section 5.3.1.7). A 68-MLD MBR plant is planned at Jurong Water Reclamation Plant and is expected to be commissioned in 2011. The MBR permeate provided from this plant will be for industrial use.

Results from two other individual studies of MBR technologies have been reported. A trial of the MRE SADF membrane has been carried out at Bedok (Oda, Itonaga, Kawashima, Chidambara-Raj, & Bartels, 2009), as well as one conducted at the Ulu Pandan site on the Asahi Kasei membrane (Qin, Oo, Tao, Kekre, & Hashimoto, 2009).

The study of the 375 m^2 Mitsubishi Rayon SADF pilot plant at Bedok ran for more than six months upstream of an RO plant. For an MLSS concentration of ~7 g/L, a flux of 33 LMH was maintained for 43 days with the TMP steady at 0.12–0.13 bar at an unspecified membrane aeration rate and an operation cycle of 7 min filtration/1 min relaxation. The membrane was maintenance cleaned weekly for 30 min by soaking in 300 mg/L sodium hypochlorite solution. This was supplemented by cleaning in place with 3 g/L sodium hypochlorite and 1% citric acid when the TMP rose by more than 0.15 bar over the course of the cycle.

TABLE 3.14 Design and O&M Data, Bedok Trials

Parameter	MBR A	MBR B	MBR C
Membrane	0.4 μm FS, 0.8 m^2 panel area	0.4 μm HF, 280 m^2 element area	0.035 μm HF, 31.5 m^2 element area
Probable technology	Kubota, double deck	MRE	Zenon (*500d*)
Membrane area, m^2	480	1120	1008
Tank sizes			
Anoxic tank volume, m^2	30.8	37.5	25.2
Aerobic tank volume, m^2	11.4	–	27.9
Memb. tank volume, m^2	32.8	37.5	21.8
O&M			
MLSS g/L	6–12	6–14	4–13
Net flux, LMH	13–28.4 (26)	16–24 (24)	6.2–29.3 (12.4)
Initial TMP, bar	0.04	0.17	0.1
Cycle, min	9 on/1 relax	13 on/2 relax	12 on/0.5 backflush + relax
Cleaning cycle time, d	90	120	3.5*
Chemical cleaning reagents	0.6% NaOCl, 1% oxalic acid	0.3% NaOCl, 2% citric acid	NaOCl, citric acid*
Derived data			
SAD_p, m^3 air/m^3 permeate	28–50 (50)	16–24 (24)	20–30 (30)
Init. permeability *K*, LMH/bar	650	66	124
SAD_p, m^3air/m^3permeate (Energy demand, kWh/m^3)			
Baseline	50 (1.4)	24 (1.3)	30 (1.7)
High flux	34 (1.2)	21 (1.0)	25 (1.3)
Low aeration	28 (1.0)	16 (0.8)	20 (1.1)

**Maintenance cleaning employed; citric acid cleaning suspended after 11 months.*

The four-month study of the Asahi Kasei membrane was based on a 50 m^3/d pilot plant. The membrane tank MLSS was in the range of 6840−9540 mg/L. The membranes were operated on a cycle of 9 min filtration/1 min backflush and at an SAD_m of 0.24−0.30 Nm^3/h per m^2 membrane area, with maintenance cleaning using 2000 mg/L NaOCl applied twice weekly. Over a 35-day period a net flux of 29 LMH was sustained with only a marginal increase in TMP from 0.24 to 0.32 bar, before a process disruption led to membrane module clogging which was not successfully cleared throughout the remainder of the trial. At this flux the corresponding mean permeability and SAD_p values were 109 LMH/bar and 9.3, respectively. Permeate COD, TOC, T-N and NH_4-N concentrations were <30 mg/L, 5−7 mg/L, <13 mg/L and <0.1 mg/L, respectively.

3.2.1.4. University of Trento Pilot Plant, Pietramurata

An extensive comparative pilot trial was conducted by the University of Trento at Pietramurata WwTP in Italy to evaluate different technologies for the upgrading of existing plants. The plants tested were originally Zenon and Kubota, with a Memcor plant subsequently studied. The first plant, installed in June 2001, consisted of two separate MBRs whose biotreatment tanks were provided by dividing a single rectangular stainless steel tank into two separated treatment lines for a Zenon and Kubota MBR, both configured with pre-denitrification. All experimental activities on site were performed between March and December, since the whole system was outdoors and the ambient temperature too low to operate during the winter months.

The biological volume was calculated based on a flux value of 15 LMH resulting in volumes of 5.14 and 4.23 m^3 for the aerobic tanks, and 2.76 and 2.27 m^3 for the anoxic tanks for the respective Zenon and Kubota processes. The permeate was removed by suction, originally using progressing cavity (*Mono*) pumps but a self-priming pump was installed in 2003. A PC pump was also used for recirculation from the aerobic to the anoxic tank at recycle ratios between 3 and 5. Aeration for the biological process was supplied to both aerobic tanks from the main aeration pipe of the full-scale plant and was measured by specific air flow meters. The Zenon/Memcor and Kubota systems were operated under different SRTs, imposed by daily sludge wasting. All usual operating parameters were monitored (TMP, DO, etc.), and both critical flux analyses and cleaning optimization conducted (Guglielmi, Chiarani, Judd, & Andreottola, 2007; Guglielmi, Chiarani, Saroj, & Andreottola, 2008).

Membrane modules originally installed comprised a Zenon 500c (67 m^2 membrane surface area) and a Kubota E50 (40 m^2 of membrane surface area). In November 2003 a new 40 m^2 HF Memcor (now Siemens) module was installed in a separate 0.3 m^3 tank and fed from the Zenon tank, returning to it under gravity. Due to this special combination, the permeate flow suctioned from both HF lines was often reduced to avoid excessive organic loading of the biological system. Simultaneous short- and long-term flux-step tests on both systems were likewise avoided. In 2005 the Memcor module was operated for a few weeks on biomass previously acclimatized using the Zenon module. The aerator depth was 1.9–2.0 m for the Zenon and Memcor modules and 2.4 m for the Kubota stack (Table 3.15).

During 2002 the Zenon module was cleaned monthly using 300 mg/L NaOCl as Cl_2. From 2004 onwards the reagent concentration for the monthly clean was reduced to 200 mg/L. From 2002 to 2003 the Kubota module was cleaned twice yearly with 3000 mg/L (i.e. 0.3 wt%) NaOCl as Cl_2 and once a year with a 10,000 mg/L (1 wt%) solution of oxalic acid. From 2004 the module was cleaned monthly with a 200 mg/L solution of NaOCl with no supplementary acid cleaning. The Memcor module was cleaned monthly using 300 mg/L NaOCl. For all three modules the membranes were additionally cleaned using 100 mg/L NaOCl immediately before each flux-step test. The mean feed COD and N-NH_4^+ levels were 575–988 and 23–33 mg/L, respectively, with the BOD/COD ratio generally being <0.5. The corresponding outlet concentrations were <22 and <5, respectively, with the total organic nitrogen (TON) being below 11 mg/L at all times.

A further 12-month trial has been conducted by this research group at the wastewater works at Lavis on the Huber VRM MBR technology. The membrane aeration rate was 0.35 $Nm^3/(m^2\,h)$, which maintained a flux of 13–22 LMH and a permeability of 150–540 LMH/bar on a cycle of 9 min filtration/1 min relaxation. The SAD_p thus ranged from 16 to 25 for this trial.

TABLE 3.15 O&M data: Optimum and Mean Values Trento Trials

	Zenon		Kubota		Memcor	
Parameter	**Opt***	**Ave**	**Opt***	**Ave**	**Opt***	**Ave**
HRT, h	14	10	6	17	11	10
SRT, d	20	21	9	17	15	14
Cycle, min	9 on/1 off		–**		9 on/1 off	
Flux#, LMH	30	16	31	15	22	21
Derived data						
Permeability# K, LMH/bar	150	120	620	261	270	182
SAD_m, $Nm^3/(m^2\,h)$	0.25	0.33	0.88	0.98	0.20	0.20
SAD_p, m^3 air/m^3 permeate	8.2	27	28	79	22	17

Conditions employed under which the highest fluxes and lowest aeration rates were sustained.
***9 on/1 off routinely, continuous operation for final trial.*
#Temperature-corrected to 20°C for viscosity.

3.2.1.5. EAWAG Pilot Plant, Kloten/Opfikon, Switzerland

The Engineering Department of EAWAG (The Swiss Federal Institute of Aquatic Science and Technology), the Swiss national water research organization, set up a pilot-scale MBR which was operated continuously for four years to gain hands-on experience of running such plants. The test protocol incorporated the study of the impact of membrane maintenance protocols, SRT and chemical flocculants on performance with reference to three different MBR technologies. The pilot plant was also used for studying the fate of micropollutants in an MBR compared with conventional technologies, the work for which has been extensively published (Göbel et al., 2005; Joss et al., 2005, 2006; Göbel, McArdell, Joss, Siegrist and Giger, 2007).

The pilot plant was located on the municipal WwTW at Kloten/Opfikon, a conventional activated sludge plant with an average dry weather flow of 17 ± 4.3 MLD (55,000 population equivalent (p.e.); maximum flow 645 L/s). This plant receives sewage from Zürich airport ($\sim$50% of the influent flow) and from the nearby towns of Kloten and Opfikon. The pilot plant was fed with primary effluent (Table 3.16) from the full-scale plant following primary treatment, comprising screening at 10 mm, a sand/oil trap and primary clarification at 2 h HRT. The flow rate was proportional ($\sim$0.2%) to that of the full-scale facility (similar to the Beverwijk trial, Section 3.2.1.1). The volume of wastewater treated was 30 ± 12 m^3/day, with a maximum daily flow of 74 m^3/day.

The pilot comprised a cascade of 2–6 tanks, each of 2 m^3 volume, providing anaerobic and anoxic treatment at recycle flow rates of 80 ± 30 and 50 ± 23 m^3/day, respectively. Aerobic treatment at mean MLSS concentrations

TABLE 3.16 Design Information, Kloten/Opflikon (Value ± Standard Deviation)

	Kubota	GE Zenon	MRE
Tank size, m^3	2.6	1.6	1.9
Sludge recycle ratio	1.3 ± 0.5	1.2 ± 0.5	1.3 ± 0.5
Sludge content, gSS/L	8.3 ± 3.4	8.2 ± 3.5	7.0 ± 2.9
Membrane Type	0.8 m^2 panels	*ZW500a*	*SUR*
Configuration	50 panels	1 module	80×1 m^2
Total membrane area, m^2	40	46	80
Net permeate prod., m^3/d	9.1 ± 3.9	11 ± 4.5	9.4 ± 3.8
Max. permeate prod., m^3/d	25.4	28.7	24.2

of between 7 and 8.3 g/L was coupled with membrane separation with membrane areas of 40–80 m^2 across the three streams (Table 3.17), all modules being single deck. Four phases of work were undertaken:

I. First year (day 1–385): 17 ± 2 days sludge age, 8 m^3 anaerobic volume.
II. Second year (day 410–745): 32 ± 3 days sludge age, 6 m^3 anaerobic volume.
III. Third year (day 770–1090): 58 ± 10 days sludge age, 6 m^3 anaerobic volume.
IV. Fourth year (day 1090–1510): 56 ± 11 days sludge age, no anaerobic volume, ferric addition.

Data for the three technologies (Table 3.17) indicate considerable changes in permeability, particularly for the Kubota module. Permeability fluctuations for

TABLE 3.17 O&M Data, Kloten/Opflikon (Value ± Standard Deviation)

	Kubota	GE Zenon	MRE
Membrane aeration rate, Nm^3/h	60 ± 4	50 ± 5, air cycling 10 s on/10 s off	29 ± 4
Cycle, min	8 on/2 relax	5 on/0.5 backflush	8 on/0.5 backflush + 1.5 relax
Ave. net flux, LMH	9.5 ± 4	10 ± 4	4.8 ± 2
Gross flux, LMH	17 ± 4	19.5 ± 3	10 ± 2
Fine bubble aeration rate, Nm^3/h	0–10	0–10	0–10
HRT, h*	3 ± 1.3	1.9 ± 0.8	1.9 ± 1.1
SRT, d	15–67	15–67	15–67
MLSS, g/L	8.3 ± 3.4	8.2 ± 3.5	7 ± 2.9
Derived data			
SAD_m, $Nm^3/(m^2\ h)$	1.5	0.54	0.37
SAD_p, m^3 air/m^3 permeate	88	28	38
$TSAD_p$, m^3 tot. air/m^3 perm.	88–90	28–38	38–48
Mean permeability K, LMH/bar	200 ± 160	200 ± 83	90 ± 70
Cleaning cycle time[2], d	Variable	Variable, 14 d	Variable

**HRT in the membrane compartment only.*

this membrane did not appear to correlate with operational changes such as cleaning protocols. For the Zenon module, however, permeability correlated roughly with MLSS concentration in the membrane compartment for much of the trial. The permeability increased appreciably – doubling over a 30-month period – with the introduction of regular maintenance cleaning (Table 3.18) every 14 days using 150 ppm NaOCl followed by a citric acid clean at pH 2.5.

The MRE unit provided a low permeability throughout the study due to clogging of the hollow fibres by dewatered sludge. Several measures taken to ameliorate this problem, such as overnight relaxation and intensive or regular chemical cleaning, were unsuccessful. Regular backflushing for 30 s during each permeate production cycle, as applied to the Zenon module, provided a stable low permeability. No difference in efficacy of maintenance cleaning at low hypochlorite concentrations 'in air' (Section 2.3.9.2) rather than in place was noted.

Maintenance cleaning in place (CIP) was conducted through backflushing for 20–30 min using 150 mg/L NaOCl followed by an acid clean at pH 2.5 ± 0.5. For the HF membranes the backflush (1.5 times the maximum operating flux) was applied intermittently with 20–30 s pulses applied at 5 min intervals. Flat sheet (FS) membranes were backflushed continuously under gravity at a maximum hydrostatic head of 0.1 bar (the limit recommended by the supplier is 0.2 bar). Chemical usage per backflush was between 2 L/m^2 for the HF membranes and 5 L/m^2 for the FS ones. Intensive (recovery) cleaning was through soaking the membranes for 3–6 h in a high-strength (0.1–0.5 wt%) NaOCl solution with pulsed aeration combined with suction for 5–10 s every 20 min. Hypochlorite cleaning led to significant foaming due to organic matter. The amount of chemical solution required depended on the packing density of the membrane, but appeared to be in the range of 20–100 L/m^2. Care was taken to drain the membrane of hypochlorite before applying the acid wash of 0.5 mM sulphuric acid together with a 5-mM citric acid buffer, the mineral acid being used to counter the alkalinity and the buffered acid pH being 2.8.

TABLE 3.18 Cleaning Protocols for the Four MBR Technologies, Kloten/Opfliken

	Kubota	Mitsubishi Rayon	Zenon
Reagents used	NaOCl, 150–2000 mg/L; citric acid pH 2.5	NaOCl, 150–1000 mg/L; citric acid pH 2.5	NaOCl, 150–1000 mg/L; citric acid pH 2.5
Established protocol	–*	–*	Maintenance clean, 150 mg/L NaOCl and citric acid clean every 14 d

No protocol tested yielded satisfactorily reproducible results.

3.2.2. Pilot and Full-Scale Plant Data

As discussed in Section 2.1, the key parameters regarding the membrane operation, and thus the maintenance of flow through the plant, are the flux, permeability, cleaning frequency and protocol, membrane aeration (for immersed and air-lift sidestream systems) and, for pumped sidestream systems, liquid crossflow. A summary of the pilot and full-scale data, the latter being taken from the case studies (Chapter 5), is given in Tables 3.19–3.25 and the key trends are presented in Figs 3.1 and 3.2. Each parameter is considered in turn for the immersed and air-lift sidestream configurations.

3.2.2.1. Flux, Permeability and Specific Aeration Demand

Comparison of the pilot study optimum data (Table 3.19) reveals substantial variation in some of the key performance parameters, possibly because the conditions were not uniformly optimized between studies and because of external factors such as temperature. For the three membrane module products featuring in 4–5 different trials, the standard deviation is between 31 and 57% across the four key parameters of flux, permeability, SAD_m and SAD_p. Extending the analysis to available full-scale data from Chapter 5 (Table 3.20) provides a more reasonable basis for an analysis and establishing appropriate operating conditions. Data collated for the range of municipal wastewater

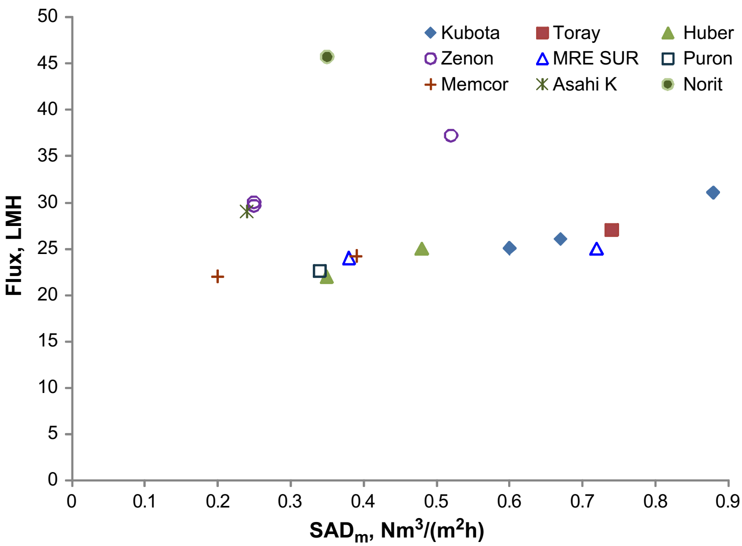

FIG. 3.1 Flux versus SAD_m, optimum values from the MWH, PUB and Trento comparative pilot plant studies.

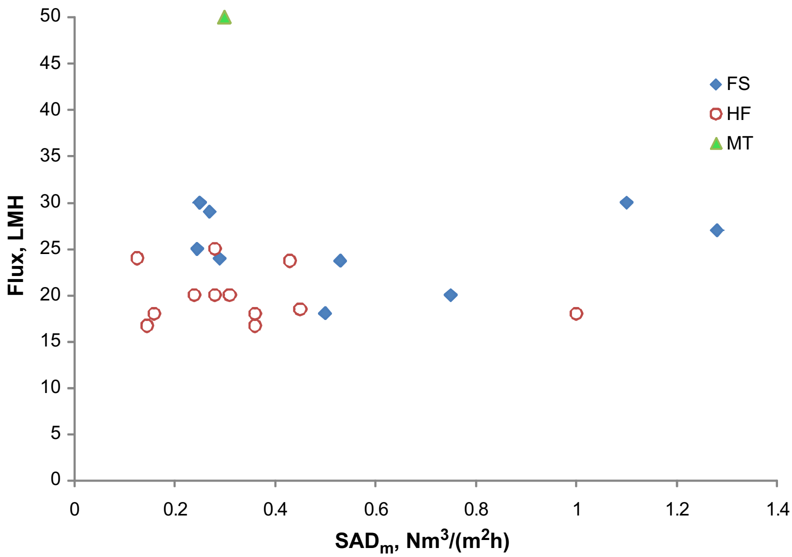

FIG. 3.2 Flux versus SAD$_m$, mean values from available full-scale municipal data.

iMBR plants summarized in Chapter 5 indicate a slightly broader range of percent standard deviation values for these same four parameters than those measured in the pilot plant studies. Values range from 21 to 67% for data collated for all FS and all HF plant, with these two categories each encompassing a range of commercial technologies.

A comparison of the key parameters of flux and SAD$_p$ for the Kubota and Zenon mean pilot plant data for municipal wastewaters reveals similar mean fluxes but with a lower aeration demand for the latter (Table 3.19), a trend repeated for the full-scale data (Table 3.20). However, it is likely that some plants are operating sub-optimally for both configurations, with overly conservative aeration demands in some cases. In the case of the mean permeability data the trend is reversed; the higher scouring rates of the FS systems permit a commensurately lower operational TMP. Correlations between flux and aeration data, on the other hand, reveal no clear trends. Aeration energy data appear not to follow the trend in specific membrane aeration energy demand (SEDA$_m$). The mean SEDA$_m$ for the FS plants is only 20% higher than that of the HF installations, based on the full-scale plant data, though the data set for the FS figure is small. The corresponding data sets for the industrial plants are small and more highly scattered than those for the municipal plant, though generally indicate fluxes 20–25% lower.

Impacts of aeration are not discernible. The general observation from laboratory-scale studies is that the attainable flux increases with increasing aeration rate due to increased scouring. However, fluxes sustained appear to

TABLE 3.19 Summary of Comparative Pilot Data for Municipal Wastewater Treatment, Optimum Values

Technology	J, LMH	K, LMH/bar	SAD_m, $Nm^3/(m^2h)$	SAD_p	Source
Kubota	12.5	500	0.75	60	DHV
	42	350	0.75	60	DHV
	25	250	0.6	24	MWH
	26	650	0.67	28	PUB
	25	500	0.67	28	PUB
	31	620	0.88	28	Trento
	9.5	200	1.5	88	EAWAG
Average	***24***	***439***	***0.83***	***45***	
% Std. dev.	***45%***	***40%***	***37%***	***54%***	
MRE SUR	8	200	0.28	48	DHV
	20	150	0.29	12	DHV
	25	140	0.72	45	MWH
	24	66	0.38	16	PUB
	4.8	90	0.37	38	EAWAG
Average	***16***	***129***	***0.408***	***32***	
% Std. dev.	***57%***	***41%***	***44%***	***53%***	
GE Zenon	20	250	0.54	27	DHV
	35	250	0.54	15	DHV
	37	270	0.52	14	MWH
	30	124	0.25	20	PUB
	30	150	0.25	8.2	Trento
	10	200	0.54	28	EAWAG
Average	***27***	***207***	***0.44***	***19***	
% Std. dev.	***51%***	***46%***	***31%***	***53%***	
Siemens Memcor	24.2	150	0.39	16	MWH
	22	270	0.2	17	Trento

(*Continued*)

TABLE 3.19 Summary of Comparative Pilot Data for Municipal Wastewater Treatment, Optimum Values—cont'd

Technology	J, LMH	K, LMH/bar	SAD_m, $Nm^3/(m^2h)$	SAD_p	Source
Huber	25	250	0.48	19	MWH
	22	540	0.35	16	Trento
Norit	37	250	0.6	16	DHV
	46	420	0.35	7.6	MWH
Koch Puron	23	340	0.34	15	MWH
Toray	27	390	0.74	27	MWH
Asahi Kasei	29	109	0.24	9.3	PUB
MRE SADF	33	264	–	–	PUB

change little with specific aeration demand for both the pilot (Fig. 3.1) and full-scale data (Fig. 3.2) even with the conservative O&M data omitted. A clear outlier is the air-lift MT system datum, where the specific aeration demand is significantly reduced but the requirement for sludge pumping increased to provide the necessary scour to maintain the flux.

3.2.2.2. Pre-Treatment

The widely recognized critical importance of pre-treatment (Section 2.3.9.1) has led to more rigorous screening of municipal wastewaters in recent years, with only 25% of the plants using 'slot' based fine screens such as wedgewire or bar (Table 3.21). There remains a trend for less rigorous screening of FS plants than of HF plants, with 3 mm being most common for the former and 1 mm for the latter. Two to three plants appear to have been fitted with RAS screening. The data would appear to be slightly at odds with those reported by Frechen, Schier and Linden (2007) (Section 2.3.9.1) for 19 European MBR plants, where uniformly conservative screening down to 0.5−1 mm was reported for both FS and HF plants.

3.2.2.3. Physical and Chemical Cleaning

A summary of physical cleaning protocols for the pilot plant study data (Table 3.22) and those from full-scale installations (Table 3.23) reveals that physical cleaning is predominantly by relaxation rather than backflushing. Pilot plant data indicate downtime for physical cleaning to account for between 4 and 20% of the operating time, with no profound difference between the two

TABLE 3.20 Summary of Full-Scale Plant Data for Wastewater Treatment, Averaged Data for iMBRs

		Flux, LMH		Permeability, LMH/bar		SAD_m, $Nm^3/(m^2\ h)$		SAD_p		$SEDA_m$*		$SEDA_t$**	
Config.		mun	ind	mun	ind	mun	ind	mun	ind	mun	ind	mun	ind
FS	**Mean**	**19.4**	**13.4**	**261**	–	**0.57**	**0.80**	**27.5**	**91.9**	**0.34**	**0.64**	**1.26**	–
	%SD	**21**	**71**	**66**	–	**67**	**93**	**56**	**98**	**19**	–	**36**	–
	Data	12	5	8	0	10	5	10	5	4	1	7	0
HF	**Mean**	**19.5**	**15.4**	**104**	**47**	**0.30**	**0.23**	**15.4**	**16.5**	**0.29**	**0.31**	**0.84**	**0.97**
	%SD	**39**	**33**	**65**	**87**	**35**	**36**	**41**	**59**	**24**	**18**	**113**	**88**
	Data	14	9	12	7	11	6	11	6	9	5	12	7

**Specific energy demand for aeration of the membrane, kWh/m^3.*
***Specific energy demand for aeration in total, kWh/m^3.*

TABLE 3.21 Summary of Screening Rating in mm, from a Total Sample of Nine FS Plants and 11 HF Plants, Full-Scale

Config.	Min	Max	Mode	Types
FS	1	3	3	1 Perforated, 5 mesh, 3 wedgewire
HF	0.5	3	1	2 Perforated, 7 mesh, 2 bar

configurations. Full-scale plant data (Table 3.24), on the other hand, reveal longer mean filtration cycles but also longer relaxation periods for the FS plants. On average the downtime for relaxation for both configurations is around 10%. However, there is further downtime and loss of permeate product for backflushing in the case of the HF modules, accounting for an additional 6–9% decrease in conversion when applied.

Maintenance cleaning is routinely employed for HF technologies, though cleaning frequency and reagent strength vary considerably (Table 3.24) and the mean values of the latter are possibly skewed by a few exceptionally high values. Cleaning frequency is from twice a week to twice a year, and both hypochlorite strength and citric acid strength vary by more than an order of magnitude. It has been recognized from pilot trials that increased, and possibly

TABLE 3.22 Summary of Pilot Plant Physical Cleaning Protocols

Study	Kubota	MRE SUR	GE Zenon	Siemens Memcor	Koch Puron	Huber	Toray	Norit A-L
DHV	8 on/2 relax	20 h on/4 h off	Back-flushed*					20 h on/ 4 h off
MWH	9 on/1 relax	12 on/2 relax	10 on/0.5 relax	12 on/0.75 relax/0.25 bf	6 on/ 0.33 bf	9 on/1 relax	9 on/1 relax	10 on/1 bf
PUB	9 on/1 relax	13 on/2 relax	12 on/0.5 bf + relax					
TRENTO	Continuous		9 on/1 relax	9 on/1 relax		9 on/1 relax		
EAWAG	8 on/2 relax	8 on/0.5 bf + 1.5 relax	5 on/0.5 bf					

r – Relaxation; b – backflushing.
Net flux 83–85% of gross flux.

TABLE 3.23 Summary of Full-Scale Municipal Plant Physical Cleaning Protocols

	Filtration Cycle Time, min			Relaxation Duration, s			Backflush Duration, s		
Config.	Min	Max	Mean	Min	Max	Mean	Min	Max	Mean
FS	9	55	22	60	300	133	X	X	X
HF	5.8	15	10	0	180	59	0	60	26

more aggressive, maintenance cleaning is required at higher operating fluxes to maintain permeability by suppressing fouling. However, no such correlation is apparent from the full-scale data; it can be assumed that more conservative maintenance cleaning is employed at some sites than others as a matter of course. Whilst increased concentrations of chemical cleanants may be onerous with respect to procurement costs and waste disposal (Section 3.6.4), the downtime incurred for chemical cleaning is small. Even at the maximum frequency of eight cleans per month at a mean downtime of 2 h per clean, the overall downtime is only around 2.2%.

Data for industrial plant are limited but on average differ little from those of the municipal plant regarding the application of hypochlorite. Citric acid is apparently not employed in the industrial plants reviewed in Chapter 5, though its use would be expected to ameliorate any scaling problems (Section 2.3.9.2).

For both FS and HF systems recovery cleans are generally applied at intervals of 6–18 months depending on the flux and, more importantly, the fouling/clogging propensity of the sludge. Recovery cleans generally employ hypochlorite concentrations from 0.1 to 0.6 wt% NaOCl (Table 3.25),

TABLE 3.24 Summary of Full-Scale HF Municipal Plant Maintenance Chemical Cleaning Protocols

	Concentration in mg/L			Frequency, Per Month			Downtime, h		
Reagent	Min	Max	Mean	Min	Max	Mean	Min	Max	Mean
NaOCl	200	3000	1050	0.167	8	3	0.75	4.5	1.98
Citric acid	450	15,000	6056	0.3	4	2	0.75	3	2
H_2O_2	2000	20,000	2000	–	–	4	–	–	X

TABLE 3.25 Summary of Full-Scale Municipal Plant Recovery Chemical Cleaning Protocols

Config., reagent	Concentration in mg/L			Frequency, per month			Downtime, hours		
	Min	Max	Mean	Min	Max	Mean	Min	Max	Mean
FS, NaOCl	250	6000	4100	0.028	0.333	0.181	3	6	4.5
FS, citric acid	10,000	30,000	20,000	0.333	0.333	0.333	3	3	3
HF, NaOCl	400	4000	2100	0.0417	0.0833	0.079	3	90	30.4
HF, citric acid	450	20,000	9400	0.083	0.167	0.095	4	24	9

sometimes adjusted to a pH of ~12 (though this can be onerous to PVDF membranes). Lower cleaning reagent strengths are used for the Huber rotating membrane, which is close to being chemical-free and operates at a slightly lower flux than other FS systems. Citric acid strengths as high as 3 wt% have been employed. There is no evident significant difference between FS and HF membranes with respect to recovery cleanant strength, but the mean frequency is higher in the case of FS membranes – since they are not generally maintenance cleaned. The percentage downtime is even lower than that of maintenance cleaning: an overnight soak of 16 h every six months incurs a downtime of only 0.36%. However, management of the higher-strength spent reagent demands greater care, though for sidestream modules the quantities of reagent demanded for cleaning are lower and thus more readily quenched prior to discharge.

3.2.3. Summary: Guide Design Values for a Municipal Immersed MBR

Data provided from the pilot trials and full-scale installations suggest a rule-of-thumb mean design net flux of 20 LMH to be reasonable for either an HF or FS iMBR. Whilst the mean SAD_m value for the FS appears to be almost double that of the HF at 28 versus 15 $Nm^3/(m^2\ h)$, there is a cluster of data points which suggest that operation at significantly lower SAD_m and SAD_p values is possible for this configuration.

As far as tank sizes are concerned the sizing of the biotank is determined by bio-stoichiometric and biokinetic considerations (Sections 3.3 and 3.4). The membrane tank size and HRT are normally made as small as possible so as to reduce:

(a) the effect of changing substrate and DO availability conditions in the main aerobic tank from those prevailing in the membrane compartment;

(b) the cost of chemical cleaning, by reducing the cleaning reagent volume for recovery cleaning;
(c) the tank construction costs, by reducing both the footprint and freeboard (the sludge height above the membrane module); and
(d) sludge accumulation, by ensuring that the sludge flow through the tank (the sludge crossflow) is 4−5 times permeate flow.

The membrane tank size is thus primarily determined by the bulk module packing density (Fig. 4.65) and the freeboard.

3.3. iMBR DESIGN METHODOLOGY: STEADY-STATE MODEL

The design for an iMBR is generally based on a combination of empirical/heuristic data (Section 3.2.2) and biokinetics/biochemical stoichiometry (Section 2.2.4). This then demands information and/or appropriate assumptions regarding the interrelationship between aeration and:

(a) permeability and cleaning protocol for the membrane permeation component; and
(b) feedwater quality, flows and biokinetics for the biological component.

Whilst the latter can be approached from a biochemical basis, the former cannot reasonably be calculated from first principles. Three distinct but interrelated design phases thus arise, namely the design of: (1) the membrane process, (2) the biological process (and determination of oxygen demand) and (3) the aeration systems.

An example of a design calculation for a nitrifying/denitrifying MBR (Fig. 3.3), 25 MLD in capacity and based on HF technology, is detailed below. The plant is intended to treat medium strength wastewater (Table 3.26) to a required effluent standard of 0.5 mg/L of NH_4-N (N_e) and 12 mg/L NO_3-N (NO_e). Bio-P removal has not been incorporated in the process design, and more details in this regard are available elsewhere (Tchobanoglous, Burton, & Stensel, 2003; Grady, Daigger, & Love, 2010). The design methodology is schematically illustrated in Fig. 3.4 with the governing biological and aeration design equations listed in Table 3.27.

3.3.1. Membrane System Design and Operation

Table 3.28 details the design and operational parameters for the membrane system, based on average values obtained from 11 HF MBRs for treatment of municipal sewage (Section 3.4.3). Key hydraulic design parameters are the net flux (J_{net}), the maximum allowed flux during a limited time period ($J_{net,peak}$) and the peak influent flow (Q_{peak}). It is assumed that $Q_{peak} = 2Q$ (Table 3.26), while $J_{net,peak} = 140\%$ of J_{net} and can be sustained for the duration of the maximum flow conditions; assumptions will vary according to the anticipated

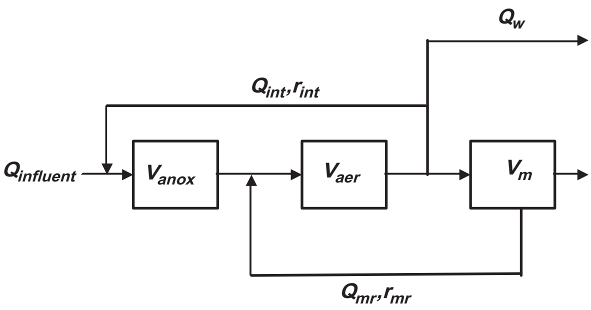

FIG. 3.3 Schematic overview of generic nitrifying/denitrifying MBR.

infiltration and trade (i.e. industrial) discharges. On the basis of the stipulated assumptions, the required membrane area is 43% larger than that needed to treat the average flow. Assuming a conservative value for the membrane packing density in the membrane tank (φ_{tank}) allows the minimum membrane tank volume ($V_{m,min}$ — Eq. (3.21)) to be determined. The larger the discrepancy existing between Q and Q_{peak}, the greater the membrane area demanded and the higher the CAPEX (capital cost) and OPEX (since a larger membrane area will lead to a commensurately greater membrane aeration demand). It may therefore be more economical to install buffering capacity for flow equalization (Section 3.4.5). Coarse bubble membrane aeration contributes to the oxygen required for biological degradation, which requires an adjustment to the biological ('process') aeration.

TABLE 3.26 Feedwater characterization (Adapted from Tchobanoglous et al., 2003)

Parameter	Unit	Value	Parameter	Unit	Value	Eq.
Q	m^3/d	25,000	Alkalinity	g/m^3	200	
Q_{peak}	m^3/d	50,000	f_{bs}	–	0.2*	
COD	g/m^3	430	f_{bp}	–	0.5*	
BOD S	g/m^3	190	f_{us}	–	0.05*	
TSS	g/m^3	210	f_{up}	–	0.25*	
VSS	g/m^3	160	bpCOD	g/m^3	215	(3.22)
TKN	g/m^3	40	pCOD	g/m^3	323	(3.23)
NH_4-N	g/m^3	25	iTSS	g/m^3	50	(3.8)
NO_3-N	g/m^3	0	nbVSS	g/m^3	53.3	(3.7)

Based on typical wastewater fractionation (Ekama et al., 1984)

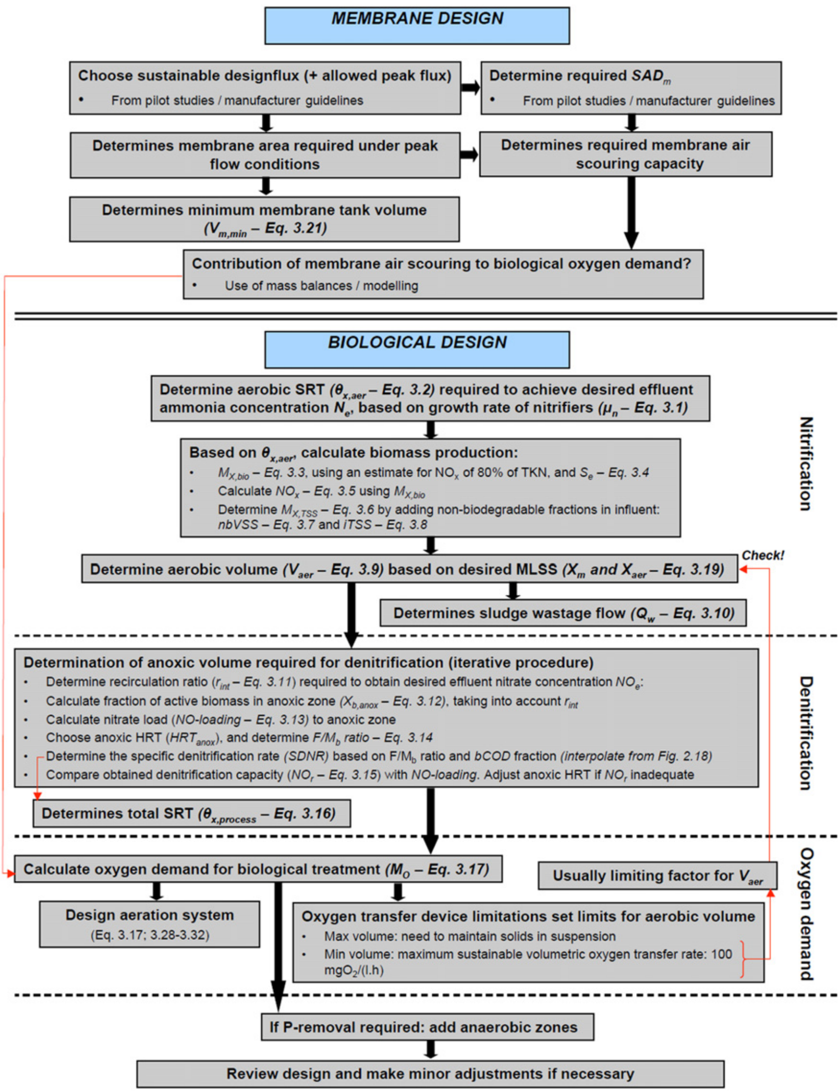

FIG. 3.4 MBR design calculation.

A key constraint of MBRs is the requirement for a relatively short retention time in the membrane tank to limit concentration of the solids and subsequent membrane channel clogging. This demands that the transfer rate between the membrane tank and the biological tank (the return activated sludge or RAS) is in the region of 3–5 times the treated water flow (r_{mr} is assumed to be 4 in Table 3.28) – somewhat higher than the equivalent RAS flow for an ASP. It is also relatively high in dissolved oxygen (DO), which makes the anoxic zone to

TABLE 3.27 Biological and Aeration Operating Parameters and Design Equations

Biological parameters	Equation	
Biological parameters		
Specific growth rate of nitrifying bacteria, μ_n in gVSS/(gVSS d)	$\left(\frac{\mu_{n,m} N_e}{K_n + N_e}\right)\left(\frac{DO}{K_O + DO}\right) - k_{e,n}$	(3.1)
Aerobic SRT, $\theta_{x,aer}$ in d	$SF\frac{1}{\mu_n}$	(3.2)
Sludge yield, $M_{X,bio}$ in g/d	$\frac{QY(S-S_e)}{1+k_e\theta_{x,aer}} + \frac{f_d k_e QY(S-S_e)\theta_{x,aer}}{1+k_e\theta_{x,aer}} + \frac{QY_n(NO_x)}{1+k_{e,n}\theta_{x,aer}}$	(3.3)
Effluent BOD, S_e g/m^3	$\frac{K_s(1+k_e\theta_{x,aer})}{\theta_{x,aer}(\mu_m - k_e) - 1}$	(3.4)
Concentration of TKN oxidized to form nitrate, NO_x in g/m^3	$TKN - N_e - 0.12M_{X,bio}\Big/Q$	(3.5)
Total sludge yield, $M_{X,TSS}$ in g/d	$M_{X,bio}/0.85 + Q(nbVSS + iTSS)$	(3.6)
Influent non-biodegradable VSS, nbVSS in g/m^3	$(1 - bpCOD/pCOD)VSS$	(3.7)
Influent inert TSS, iTSS in g/m^3	$TSS - VSS$	(3.8)
Aerobic tank volume, V_{aer} in m^3	$\frac{M_{X,TSS}\theta_{x,aer}}{X_{aer}}$	(3.9)
Sludge wastage flow, Q_w in m^3/d	$\frac{V_{aer}}{\theta_{x,aer}}$	(3.10)
Nitrate recirculation ratio, r_{int}	$\frac{NO_x}{NO_e} - 1$	(3.11)
Active biomass concentration in anoxic zone, $X_{b,anox}$ in g/m^3	$\left(\frac{Q\theta_{x,aer}}{V}\right)\left(\frac{Y(S-S_e)}{1+k_e\theta_{x,aer}}\right)\left(\frac{r_{int}}{r_{int}+1}\right)$	(3.12)
Nitrate load to anoxic zone, NO-loading in g/d	$Qr_{int}(TKN - N_e - 0.12M_{X,bio}/Q)$	(3.13)

Food to active biomass concentration in anoxic zone, F/M_b in g BOD/(gTSS d)	$\frac{QS}{V_{anox}X_{b,anox}}$	(3.14)
Denitrification capacity, NO_r, in g/d	$V_{anox}\mathrm{SDNR}(X_{b,anox})$	(3.15)
SRT, $\theta_{x,process}$ in d	$\frac{V_{aer}X_{aer}+V_{anox}X_{anox}}{Q_wX_{aer}} = \theta_{x,aer} + \theta_{x,anox}$	(3.16)
Oxygen requirement, M_O in kg/d	$Q(S - S_e) - 1.42M_{X,Bio} + 4.33Q\mathrm{NO}_x - 2.86Q(\mathrm{NO}_x - \mathrm{NO}_e) = M_m + M_b$	(3.17)
MLSS concentration in anoxic zone, X_{anox}, in g/m^3	$X_{aer}\left(\frac{r_{int}}{1+r_{int}}\right)$	(3.18)
MLSS concentration in aerobic zone, X_{aer}, in g/m^3	$X_m\left(\frac{r_{mr}}{1+r_{mr}}\right)$	(3.19)
Anoxic tank as fraction of total volume, f_{anox}	$\frac{V_{anox}}{V_{process}}$	(3.20)
Minimum required membrane tank volume, $V_{m,min}$ in m^3	$\frac{A_m}{\varphi_{tank}}$	(3.21)
Biodegradable particulate COD, *bp*COD in g/m^3	$f_{bp}\mathrm{COD}$	(3.22)
Total particulate COD, *p*COD, in g/m^3	$(f_{bp} + f_{up})\mathrm{COD}$	(3.23)
Sludge waste per unit permeate, Q'_w, m^3/m^3	$\frac{Q_w}{J_{net}A_m}$	(3.24)
Hydraulic retention time for tank *x*, HRT_x, in h	$\frac{V_x}{Q}$	(3.25)
Total process volume, $V_{process}$, in m^3	$V_{anox} + V_{aer}$	(3.26)
Total process HRT, $\mathrm{HRT}_{process}$, in h	$HRT_{anox} + HRT_{aer}$	(3.27)
Aeration parameters		
Suspended solids correction factor, α	$e^{-\omega_x X}$, where x refers to fine or coarse bubble	(3.28)

(*Continued*)

TABLE 3.27 Biological and Aeration Operating Parameters and Design Equations—cont'd

Biological parameters	Equation	
Temperature correction factor, Φ	$1.024^{(T-20)}$	(3.29)
Membrane aeration rate, $Q_{A,m}$ in Nm^3/h	$SAD_m A_m$	(3.30)
O_2 transferred by membrane aeration, M_m in kgO_2/d	$Q_{A,m}\rho_A(SOTE_{coarse}y_{coarse})O_{A,m}\alpha\beta\Phi$	(3.31)
Net air flow for biological requirements, $Q_{A,b}$ in Nm^3/h	$\frac{M_0-M_m}{\rho_A(SOTE_{fine}y_{fine})O_{A,m}\alpha\beta\Phi}$	(3.32)

Symbols:

bCOD	Biodegradable COD in g/m^3
f_{anox}	Anoxic tank as % of aerobic tank size
f_{bp}	Slowly biodegradable COD fraction
f_{bs}	Readily biodegradable COD fraction
f_{up}	Particulate non-biodegradable COD fraction
f_{us}	Soluble non-biodegradable COD fraction
k_e	Heterotrophic endogenous decay coefficient, gVSS/(g VSS d)
$k_{e,n}$	Nitrification endogenous decay coefficient, gVSS/(g VSS d)
K_s	Saturation coefficient (heterotrophic), $gBOD/m^3$
$MLSS$	Design mixed liquor suspended solids, g/m^3

$O_{A,m}$	Mass percentage of oxygen in air, %
Q	Average flow, in m^3/d
M_m	Oxygen transferred by membrane aeration, kg/d
r_{mr}	Membrane recirculation ratio
S	BOD influent, in g/m^3
SDNR	Specific denitrification rate, gNO_3-N/(gMLVSS d)
$SOTE_{fine}$ and $SOTE_{coarse}$	Standard oxygen transfer efficiency, fine and coarse bubble aeration, %/m
TKN	Nitrogen influent, g/m^3
TSS	Total suspended solids influent in g/m^3
VSS	Volatile suspended solids, g/m^3
X_m	Mixed liquor suspended solids level in membrane tank, g/m^3
Y	Heterotrophic yield coefficient, gVSS/(gBOD)
y_{coarse} and y_{fine}	Coarse and fine bubble aerator depth, m
Y_n	Nitrification yield coefficient, gVSS/(gBOD)
β	Salinity correction factor
φ_{tank}	Membrane packing density in membrane tank, m^2/m^3
μ_m	Maximum heterotrophic specific growth rate, gVSS/(gVSS d)
$\mu_{m,n}$	Maximum nitrification specific growth rate, gVSS/(gVSS d)
ρ_A	Air density, kg/m^3
ω_{fine} and ω_{coarse}	ω-Factor for fine and coarse bubble aeration.

TABLE 3.28 HF Membrane Design and Operational Parameters

Parameter	Unit	Value	Ref.	Parameter	Unit	Value	Ref.
J_{net}*	LMH	19.5		SAD_p	Nm^3/m^3	15.4	Table 3.1
$J_{net,peak}$	LMH	27.3		φ_{tank}	m^2/m^3	45	
K	LMH/bar	104		r_{mr}	–	4	
A_m	m^2	76,312	Q/J	X_m	g/m^3	10,000	
SAD_m	$Nm^3/(m^2 h)$	0.30		$V_{m,min}$	m^3	1696	Eq. (3.21)

**The net flux relates to the operating flux and system downtime incurred by cleaning (Table 3.1).*

which it may be returned for denitrification less efficient. It is thus usually returned to the aerobic zone of the biotank.

3.3.2. Biological Process Design and Oxygen Demand

As with other aerobic treatment processes, the biological component design demands knowledge of appropriate values of the biokinetic constants. A comprehensive listing of biokinetic constants, along with references identifying their origins, is provided in Appendix B. Since there is no consensus on appropriate biokinetic values for an MBR process, typical values as applicable to ASPs as reported in Tchobanoglous et al. (2003) may be used: $K_s = 20$, $k_e = 0.12$, $\mu_m = 6$, $Y = 0.4$, $K_n = 0.74$, $k_{e,n} = 0.08$, $\mu_{m,n} = 0.75$ and $Y_n = 0.12$. These are not necessarily the most appropriate for an MBR, but it has been demonstrated (Verrecht, Maere, Nopens, Brepols, & Judd, 2010) that their use appears not to introduce significant error.

The design of the biological system is not greatly different from that of conventional activated sludge (CAS). The simplest procedure employs the stoichiometric correlations introduced in Section 3.4 and summarized in Fig. 3.4 and Table 3.29. However, it is widely recognized (Ekama et al., 1984; Tchobanoglous et al., 2003; Grady et al., 2010) that due to the complex interactions between the different biological processes, a correct process design for nutrient removal can only be obtained through dynamic simulations using the IWA ASM models (Section 3.4.1). A design based on stoichiometry should thus be considered as a useful starting point, which can be refined and optimized by using it as input for the IWA ASM models (Section 3.4).

The design procedure starts by determining the aerobic SRT ($\theta_{x,aer}$ – Eq. (3.2), Table 3.27) based on the growth rate of nitrifiers (μ_n – Eq. (3.1)) required to achieve the desired effluent ammonia concentration N_e. Often, a safety factor SF is applied to ensure a sufficiently long SRT to handle TKN

TABLE 3.29 Biological Operating Parameters and Design Calculations

Aerobic zone				Anoxic zone			
Parameter	Unit	Value	Eq.	Parameter	Unit	Value	Eq.
$\theta_{x,aer}$	d	9.3	(3.2)	$\theta_{x,anox}$	d	3.6	(3.16)
V_{aer}	m^3	5326	(3.9)	V_{anox}	m^3	3333	(3.25)
HRT_{aer}	h	5.1	(3.25)	HRT_{anox}	h	3.2	
μ_n	gVSS/(gVSS d)	0.16	(3.1)	r_{int}	–	1.61	(3.11)
S_e	$gBOD/m^3$	0.79	(3.4)	$X_{b,anox}$	g/m^3	1524	(3.12)
NO_x	gNO_3-N/m^3	31.3	(3.5)	NO-loading	kg/d	1255	(3.13)
X_{aer}	g/m^3	8000	(3.19)	X_{anox}	g/m^3	4930	(3.18)
DO	gO_2/m^3	2		F/M_b	gBOD/(gTSS d)	1.48	(3.14)
SF	–	1.5		SDNR	g/g d	0.25	–*
				NO_r	kg/d	1270	(3.15)
				NO_r/NO-loading	–	1.01	
Overall process parameters				Sludge yield			
Parameter	Unit	Value	Eq.	Parameter	Unit	Value	Eq.
$\theta_{x,process}$	d	12.9	(3.16)	$M_{X,bio}$	kgVSS/d	1714	(3.3)
$V_{process}$	m^3	8659	(3.26)	$M_{X,TSS}$	kgTSS/d	4599	(3.6)
$HRT_{process}$	h	8.3	(3.27)	Q_w	m^3/d	574	(3.10)
f_{anox}	–	0.38	(3.20)	Q'_w	m^3/m^3	0.02	(3.24)
M_O	kgO_2/d	7079	(3.17)				

*Interpolated from Fig. 2.18.

peaks in the influent. Temperature has a large impact on the growth rate of nitrifiers and $\theta_{x,aer}$ should be determined for the lowest anticipated temperatures (Ekama et al., 1984). For simplicity, all calculations in the example design have been performed at 20 °C, such that no temperature corrections are necessary. Based on $\theta_{x,aer}$, the biomass production in terms of VSS ($M_{x,bio}$ – Eq. (3.3)) can be calculated based on the effluent BOD concentration (S_e – Eq. (3.4)) and an estimate for NO_x of 80% of the TKN. By adding the influent non-biodegradable VSS (nbVSS – Eq. (3.7)) and inert TSS (iTSS – Eq. (3.8)) the sludge yield based on total solids is obtained ($M_{x,\mathrm{TSS}}$ – Eq. (3.6)), allowing calculation of the required aerobic volume (V_{aer} – Eq. (3.9)) and the sludge wastage flow (Q_{w} – Eq. (3.10)).

If biological nitrogen removal is required, sufficient anoxic tankage must be added. To determine this volume, an iterative procedure can be followed. The recirculation rate (r_{int} – Eq. (3.11)) can be calculated based on the effluent nitrate limit (NO_{e}), which allows calculation of the active fraction of the anoxic biomass ($X_{\mathrm{b,anox}}$ – Eq. (3.12)). $M_{x,\mathrm{bio}}$ can be used to calculate a better approximation for the concentration of TKN oxidized to form nitrate (NO_x – Eq. (3.5)), which allows determination of the nitrate load to the anoxic zone (*NO*-loading – Eq. (3.13)). Anoxic HRT ($\mathrm{HRT}_{\mathrm{anox}}$) has to be estimated, and based on the corresponding food to active biomass concentration in the anoxic zone (F/M_{b} – Eq. (3.14)), the specific denitrification rate (SDNR) and denitrification capacity (NO_r – Eq. (3.15)) can be determined. If NO_r is significantly higher or lower than NO-loading, $\mathrm{HRT}_{\mathrm{anox}}$ has to be adjusted and the procedure reiterated until an adequate value has been found. The total process SRT ($\theta_{x,\mathrm{process}}$ – Eq. (3.16)) can then be calculated. Finally, the oxygen demand required for biological treatment (M_{O} – Eq. (3.17)) can be determined. Table 3.29 shows the results of this design approach applied to the wastewater characteristics given in Table 3.26 at average influent flow Q. It is clear that the contribution of nbVSS and iTSS cannot be ignored for the typical medium strength wastewater of this example, since the sum of these components makes up a large part (63%) of the total sludge production.

Even though the above discussion shows that Monod kinetics can be applied to determine biological tank sizes for MBRs, there are generally two other considerations significantly affecting bioreactor sizing which relate to the physical constraints of available equipment. Both are associated with the oxygen loading demanded by the wastewater to be treated, and the resulting energy input required to meet this oxygen demand. If a specified oxygen supply is needed to meet the process oxygen requirement, and the oxygen transfer equipment can transfer that oxygen with a specified efficiency (kg O_2/kWh), then the energy required is determined by this. This sets the upper and lower limits on the bioreactor volume, viz. (Grady et al., 2010):

(a) The maximum bioreactor volume that can be supported by the energy input is associated with the requirement for the oxygen transfer device not only used to transfer oxygen but also to keep solids in suspension. However, since volume is typically minimized in an MBR this is generally not a consideration.

(b) The minimum bioreactor volume is determined by the constraint on the oxygen transfer device with respect to the maximum volumetric energy input. When volumetric energy input is excessive, the shear rate becomes high enough to produce deflocculation which can then promote membrane fouling. Also, as a rule of thumb, it is generally accepted that any oxygen transfer device is physically limited regarding the maximum achievable volumetric oxygen transfer rate in $mgO_2/(L\,h)$. For air-based systems (as in an MBR) the practical limit on a sustained basis is generally about 100 $mgO_2/(L\,h)$, peaking as needed to 150 $mgO_2/(L\,h)$. This constraint determines the actual size of many MBRs, as membranes permit higher MLSS concentrations to be maintained than in CAS systems. When this is applied to the design example, the oxygen demand is 295 kg/h (Eq. (3.17)). This implies a minimum aerobic tank HRT of 2.9 h versus the calculated value (HRT_{aer}) of 5.1 h, which can therefore be considered sufficient to overcome the physical constraint of the oxygen transfer devices.

Another consideration for any ASP (including MBRs) is the trade-off between the size of the bioreactor and the solid–liquid separation specification – or the membrane area, in the case of an MBR. This relates to the flux that can be sustained at the solids loading rate to which the flux pertains. This in turn is dictated by the effectiveness of the crossflow or air scour in sweeping the solids away from the surface through the application of shear. If the applied shear is insufficient to remove the colloidal and particulate material from the membrane:solution interface this material can deposit and accumulate on the membrane surface and produce fouling and clogging. The solids loading rate is simply the flux multiplied by the solids concentration (the MLSS). Hence reducing the MLSS would be expected to have a significant impact on fouling and clogging. In practice the impact is minor, but conservative sizing of the biotank is nonetheless beneficial in providing buffering of hydraulic surges.

Unlike CASPs, the membrane separation process generates a significant MLSS concentration difference between the membrane and respective anoxic and aerobic biological tanks/zones, as determined by the recirculation ratios r_{int} and r_{mr}, respectively. In the design example, a typical maximum MLSS concentration in the membrane tank X_m of 10 kg/m^3 is combined with an assumed membrane recirculation ratio r_{mr} of 4, resulting in an MLSS concentration in the aerobic zone (X_{aer} – Eq. (3.19)) of 8 g/L. A recirculation rate r_{int} of 1.61 (to obtain an effluent NO_e concentration of 12 mg/L – Eq. (3.11)) leads to an MLSS concentration in the anoxic zone X_{anox} of 4930 mg/L (Eq. (3.18)). The MLSS gradient has to be incorporated in the design procedure to allow correct tank sizing and estimation of required aeration capacity, since the MLSS concentration impacts on the α-factor (Eq. (3.28)).

3.3.3. Aeration System Design

The design of the aeration system is one of the most important differences between an iMBR and a CASP, since iMBRs typically have both membrane and biology aeration. It is thus necessary to determine how much oxygen arising from membrane aeration contributes to the total oxygen required for carbonaceous degradation and nutrient removal (M_O – Eq. (3.17)). Dissolved oxygen (DO) concentrations in the membrane tanks can be high due to the prevailing vigorous membrane aeration, and therefore some of it will be available for biological degradation. However, since it is not possible to incorporate this in a simple stoichiometric design, for the purposes of the design calculation it is assumed that all DO from membrane aeration contributes to the total O_2 requirement for biological degradation. This is likely to be an overestimation, and a more accurate representation requires the use of the IWA ASM models under dynamic conditions (Section 3.4).

Table 3.30 illustrates the aeration system design and operating parameters, ultimately allowing cost evaluation (Section 3.5). The coarse bubble membrane aeration is evidently much less efficient than the biology aeration for providing dissolved oxygen, as indicated by the much higher air flow rate ($Q_{A,m}$ – Eqs (3.30) and (3.31)) required to transfer oxygen into the biomass. This can be attributed to several factors in Eq. (3.31): (a) the higher MLSS concentration and its impact on the α-factor (Eq. (3.28)), (b) a lower aerator depth y (determined by the membrane module length and placement) and (c) a lower oxygen transfer efficiency $SOTE_{coarse}$, all compared to the values for fine bubble aeration.

3.4. IMBR DESIGN METHODOLOGY: DYNAMIC MODEL

3.4.1. Structure and Simulation of an MBR Dynamic Model

Steady-state models (Section 3.3) are represented by algebraic equations and do not include time dependency. Dynamic models, on the other hand, typically consist of a set of ordinary differential equations (ODEs) that contain time as an independent variable. ODEs are obtained by setting up mass balances over the bioreactor for the different model components (soluble and particulate COD, NH_4, NO_3, PO_4, autotrophic and heterotrophic biomass, DO, etc.). Mass balances have the following general format:

$$\frac{dM_x}{dt} = \text{transport} + \text{conversion},$$

where M_x indicates the mass of component x in the system. The left-hand side is the rate of change of mass of this component. This can be brought about by either physical transport or conversion. The former expresses the hydraulics and mixing behaviour of the system. Often, for simplicity, fixed volume tanks (i.e. flow rate in = flow rate out) and a tanks-in-series approach are used to describe mixing if plug flow behaviour prevails (Levenspiel, 1999). The last term expresses either removal or production of a component in the system, through stoichiometry and process kinetics, taking a negative sign when the component is consumed and a positive sign when it is produced. These terms are actually similar to the rate equations for steady-state models.

TABLE 3.30 Aeration System Operating Parameters and Design Calculations

Parameter	Unit	Biology	Membrane	Eq.
Diffuser type		Fine bubble	Coarse bubble	
Standard oxygen transfer efficiency, $SOTE_x$	%/m	0.05	0.02	
Air density, ρ_A	kg/m^3	1.2	1.2	
Correction factor exponent, ω_x	–	0.084	0.084	
Mass % oxygen in air, $O_{A,m}$	%	23.2	23.2	
Mass transfer correction, solids, α	–	0.51	0.43	(3.28)
Mass transfer correction, salinity, β	–	0.95	0.95	
Mass transfer correction, temperature, Φ	–	1	1	(3.29)
Aerator depth, y_x	m	5	2.3	
Air flow rate, membrane tank, $Q_{A,m}$	Nm^3/h	–	22893	(3.30)
O_2 transferred by membrane aeration, M_m	kg/d	–	2886	(3.31)
O_2 to be provided by biology aeration, M_b	kg/d	4193	–	(3.17)
Air flow rate, biotank, $Q_{A,b}$	Nm^3/h	–	5174	(3.32)

Application of mass balances to conventional activated sludge (CAS) systems has led to the well-established family of Activated Sludge Models (ASM1, ASM2, ASM2d, ASM3). These models were built to address different scenarios and differ in the number of components accounted for and the number of processes incorporated (Henze, Grady, Gujer, Marais, & Matsuo, 2000). ASM models are widely accepted in literature and have been frequently applied to CAS for system analysis and optimization, and have more recently been employed for modelling MBR systems (Verrecht, Maere, Benedetti, Nopens, & Judd, 2010; Verrecht et al., in press; Fenu et al., in press). There is still debate in the literature as to whether the ASM models are directly applicable to MBRs, given some specific differences of the MBR compared to CAS, and specifically whether default parameter values can be reused or should be adapted. Consensus appears to have been reached on some topics, e.g. MLSS-dependency of the aeration model (Cornel, Wagner, & Krause, 2003). However, the limited amount of published information on application of ASMs to full-scale MBRs makes further agreement challenging, especially since biological and separation performance is often so high that predicting effluent quality is not a key issue. However, the more recent interest in MBR energy demand has invigorated dynamic modelling of MBRs, particularly with reference to OPEX.

Since dynamic models comprise a set of non-linear ODEs their solution demands the use of numerical solvers to compute the model outputs. Various commercial software tools exist that contain numerical solvers, and they also are offered with a suite of implemented ASM models and models dedicated to MBR systems. Examples of specific commercial wastewater treatment simulators are (in alphabetical order, websites accessed June 2010):

- AQUASIM (www.aquasim.eawag.ch)
- BioWin (www.envirosim.com)
- GPS-X (www.hydromantis.com)
- SIMBA (www.ifak-system.com)
- STOAT (www.wrcplc.co.uk/software)
- WEST (www.mostforwater.com).

3.4.2. Model Calibration and Validation

Literature models contain a large number of model parameters with default values ascribed either through experiments or through expert knowledge. However, these values may differ slightly for different cases, and thus each model developed for a specific system under study requires calibrating to allow it to be used. Model calibration can be defined as finding a unique set of model parameters that provide a good description of the system behaviour, and can be achieved by confronting model predictions with actual measurements performed on the system. A two-step procedure is generally adopted for wastewater treatment models (Fig. 3.5).

(a) A *steady-state calibration* is performed, entailing feeding of the dynamic model with a constant influent quality specification typically obtained from the average of the sewage load and composition of measurements performed at the plant. An influent characterisation (Roeleveld, & van Loosdrecht, 2002) is then performed that translates measured variables into model state variables. Subsequently, the model prediction is confronted with average values of important system variables (e.g. effluent COD, TKN, NH_4, NO_3, PO_4; DO-levels, TSS). If the model prediction is unacceptable, calibration is performed by slightly modifying parameter values. Since many processes are interrelated, the following calibration sequence is typically adopted: solids balance, aeration model, nitrification, denitrification and P-removal.

(b) A *dynamic calibration* is performed, whereby the model is fed with dynamic influent data over a prolonged time frame (typically one week)

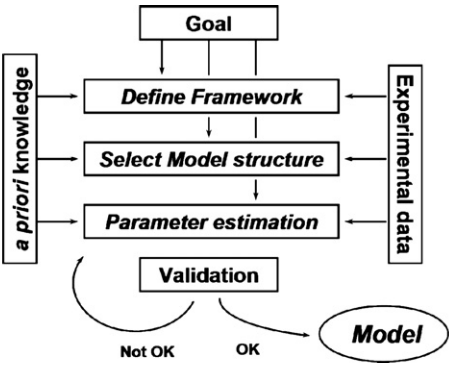

FIG. 3.5 Steps involved in modelling a system.

which can either be obtained from on-line sensors or reconstructed by using correlations between COD and nutrient load versus influent flow rate. The model output will then also be dynamic and should be compared to dynamic measurements of system states (either in the bioreactor or effluent) typically collected using on-line sensors. The latter have become quite common at WwTPs since they are used as input for control loops in the system (e.g. DO control based on DO or NH_4 levels). The dynamic calibration further fine-tunes the model parameters to provide a good match between measured data and model predictions.

Once calibrated the model needs to be validated, typically using an independently collected data set. If the calibrated model is able to describe this data set without further adjustment in the parameter values, the model can be considered validated and used to provide answers to the modelling goals. However, it is not always valid to extrapolate the model to situations substantially different to those used during calibration/validation.

3.4.3. Case Study: Dynamic MBR Model for Energy Optimisation

Dynamic modelling has been applied to a small-scale MBR for decentralized reuse to optimize the energy consumption without compromising effluent quality (Verrecht et al., 2010), in which the following modelling steps were employed:

- Modelling goal: Energy optimization of the plant whilst maintaining effluent quality.
- *A priori* knowledge: The plant was configured such that nitrification, denitrification and COD and bio-P removal occurred. Several literature models exist describing this type of system using different levels of complexity.
- Experimental data: Frequently logged influent and effluent data were available as well as DO sensor data (used for aeration control). Additional

measurements were performed for characterization of: (a) system hydraulics (through a tracer test) and (b) influent dynamics and dynamic system behaviour (through a comprehensive monitoring campaign).

- Framework definition: Definition of the dynamic system behaviour is required to provide effluent quality data and ensure it is safeguarded regardless of changing feed conditions. Moreover, energy consumption is also sensitive to system dynamics. Therefore, a dynamic model structure (i.e. a set of ODEs) is required.
- Model structure: The tracer test confirmed that the anoxic and aerobic tank of the MBR could each be modelled as a completely mixed tank. ASM2d was chosen as the biokinetic model since bio-P removal occurred in the system despite the absence of a dedicated anaerobic zone. Given the importance of the impact of aeration on energy consumption, a dedicated aeration model was used to account for coarse and fine bubble aeration and the detrimental effect of elevated TSS concentrations on oxygen transfer efficiency. Empirical correlations for the energy consumption of the unit processes – aeration, pumping, mixing – were derived from measurements on the plant.
- Model calibration: A steady-state calibration was performed on average data for influent, effluent and sludge collected over a period of approximately twice the SRT. It appeared that for steady-state simulations the growth of phosphate-accumulating organisms, and consequently sludge production, could not be simulated correctly without using unrealistic parameter values. Analysis showed the anaerobic conditions needed for bio-P removal did not occur when using a steady-state influent, indicating the need for a dynamic model to account for influent variations. The dynamic calibration was carried out using NH_4, NO_3, PO_4 and TSS data collected over a 24-h measurement campaign. The results showed that default ASM2d parameter values were sufficient to describe the biological processes accurately. The aeration model was calibrated using on-line DO sensor data and air flow measurements. Changing only the diffuser fouling factor of the aeration model was sufficient to match closely the measured DO profile, and could be justified since an inspection of the membrane diffusers showed visible fouling.
- Scenario analysis: A scenario analysis was conducted with the calibrated dynamic model to determine the optimum operating conditions by varying the following operational degrees of freedom: SRT, internal recirculation flow, DO setpoint and membrane aeration demand. A total of 486 different scenarios were simulated and ranked in terms of energy consumption while still compliant with effluent quality standards. The operational parameters that eventually were chosen for implementation are shown in Table 3.31. The new operational conditions led to an on-site reduction in energy consumption of 23%. Decreasing the membrane aeration flow and SRT had the most profound effect on total operational energy consumption, but a trade-off in achievable NH_4 removal and decreasing SRT became

TABLE 3.31 Operational Parameter Values from Scenario Analysis: Original and Optimized System (Verrecht et al., 2010)

Parameters	Unit	Original	New
Operational parameters			
Membrane aeration	Nm^3/h	84	42
SRT	d	47	35
DO setpoint	mg/L	2	1.25
Recirculation ratio	–	2.27	4.25
Energy consumption			
Measurement	kWh/m^3	4.03	3.11
Reduction	%		23%
Model prediction	kWh/m^3	4.25	2.99
Deviation from real value	%	5.1	3.9

apparent. Increasing the recirculation flow led to improved TN removal and a deterioration in TP removal, as also predicted by the model. Data collected after implementation over a period corresponding to approximately twice the SRT also indicated that membrane permeability was maintained at the original levels without changing the cleaning regime. This modelling approach thus allowed the operating envelope to be identified for meeting criteria based on energy demand and specific water quality determinants.

- Model validation: As no separate data set was available prior to the scenario analysis, the validation was performed by comparing the system measurements after having implemented the best scenario. Both effluent quality and energy consumption were found to adequately match the data.

3.5. COST BENEFIT ANALYSIS

A cost benefit analysis is applied to determine beneficial changes in plant design and operation. Two parameters can be used to quantify impacts on cost and effluent quality through dynamic modelling:

(a) the net present value (NPV, Eq. (3.44)), calculated for a plant lifetime of 30 years and taking into account all capital and operational expenditures (CAPEX and OPEX) over the plant life; and
(b) the effluent quality index (EQI, Eq. (3.45)), which quantifies the pollution load to a receiving water body expressed in kg pollution units per d (Copp,

2002). The pollution units PU_x are the product of the respective effluent concentrations at time t and a weighting factor β_x, where $\beta_{TSS} = 2$, $\beta_{COD} = 1$, $\beta_{BOD} = 2$, $\beta_{TKN} = 20$ and $\beta_{NO} = 20$ (Vanrolleghem, Jeppsson, Carstensen, Carlsson, & Olsson, 1996).

The methodology for the cost benefit analysis (Fig. 3.6, Table 3.32) can be applied to the generic nitrifying/denitrifying plant (Fig. 3.3) in conjunction with the biokinetic Activated Sludge Model No. 1 (ASM1), using the default ASM1 values (Henze et al., 2000) for the biokinetic parameters. The aeration model used incorporates the influence of the MLSS concentration on the actual oxygen transfer rate (AOTR, Eq. (3.33)) through the α-factor (Eq. (3.28)). It also accounts for variations in oxygen transfer from using coarse and fine bubble diffusers for membrane and biological aeration, respectively. AOTR is derived through calculating the actual oxygen transfer efficiency (AOTE, Eq. (3.34)), which is dependent on the diffuser oxygen transfer efficiency $SOTE_x$

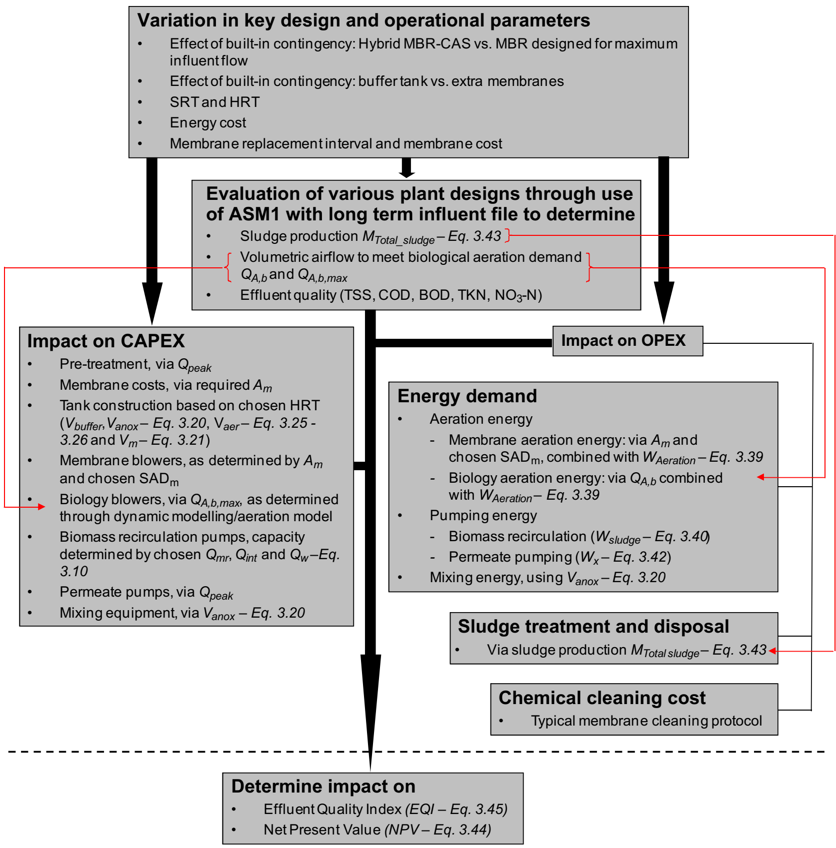

FIG. 3.6 Schematic overview methodology for cost benefit analysis.

TABLE 3.32 Equations for Determination of CAPEX, OPEX, NPV and EQI

Actual oxygen transfer rate, AOTR, kg/d	$Q_A \rho_A O_{A,m} \mathrm{AOTE}$	(3.33)
Actual oxygen transfer efficiency, AOTE, %	$\mathrm{SOTE}_x y_x \frac{(\beta C^*_{T,av} - C)}{C^*_{20}} \Phi \alpha F_x$	(3.34)
Average dissolved oxygen saturation concentration for clean water at temp. T and tank depth h_x, $C^*_{T,av}$, gO_2/m^3	$0.5 C^*_T (P_d/P_{A,1} + O_{out}/O_{A,v})$	(3.35)
Dissolved oxygen saturation concentration for clean water at temp. T and 1 atm, C^*_T, gO_2/m^3	$14.65 - 0.41T + 7.99 \cdot 10^{-3} T^2 - 7.78 \cdot 10^{-5} T^3$	(3.36)
Pressure at bottom of aeration tank, P_d, Pa	$P_{A,1} + \rho_{sludge} g h_x$	(3.37)
O_2 in air leaving the surface of the aeration tank O_{out}, %	$\frac{O_{A,v}(1-\mathrm{AOTE})}{1 - O_{A,v}\mathrm{AOTE}}$	(3.38)
Blower power requirement, $W_{Aeration}$, kW	$\frac{P_{A,1} T_{K,1} \lambda}{2.73 \times 10^5 \xi (\lambda - 1)} \left[\left(\frac{P_{A,2}}{P_{A,1}} \right)^{1-\frac{1}{\lambda}} - 1 \right] Q_{A,x}$	(3.39)
Derived sludge pumping power requirement, W'_{sludge}, kWh/d	$\frac{E_q}{t_e - t_0} \cdot \int_{t_0}^{t_e} [Q_{int}(t) + Q_{mr}(t) + Q_w(t)] dt$	(3.40)
Power required for sludge pumping, W_{sludge}, kW	$\rho_{sludge} g Q_{sludge} \Delta h$	(3.41)
Power requirement for permeate pumping and backwashing, W_x, kWh/d	$\frac{1}{t_e - t_0} \int_{t_0}^{t_e} \frac{\Delta p_x Q_x}{\xi_p} dt$	(3.42)
Total sludge production, $M_{total\ sludge}$, kg/d	$\frac{(M(\mathrm{TSS}_{system})_{t_e} - M(\mathrm{TSS}_{system})_{t_0} + \int_{t_0}^{t_e} \mathrm{TSS}_w(t) Q_w(t) dt)}{t_e - t_0}$	(3.43)
Net present value, NPV, €	$\sum_{t=0}^{29} \frac{(CAPEX)_t + (OPEX)_t}{(1+i)^t}$	(3.44)
Effluent quality index, EQI, kg PU/d	$\frac{1}{t_e - t_0} \int_{t_0}^{t_e} [\mathrm{PU_{TSS}}(t) + \mathrm{PU_{COD}}(t) + \mathrm{PU_{BOD}}(t) + \mathrm{PU_{TKN}}(t) + \mathrm{PU_{NO}}(t)] Q(t) dt$	(3.45)

and fouling factor F_x (Krampe and Krauth, 2003 after Tchobanoglous et al., 2003; Germain et al., 2007; Henze, van Loosdrecht, Ekama, & Brdjanovic, 2008; Verrecht et al., 2008, 2010a; Stenstrom, & Rosso, 2008; Maere et al., 2009).

The various plant designs can be subjected to a long-term dynamic influent file including all phenomena that are typically observed in a year of full-scale WwTP influent data (Gernaey, Rosen, & Jeppson, 2006). The average influent flow (Q) is 20,851 m^3/d, while the maximum instantaneous flow is 59,580 m^3/d. This modelling approach allows determination of the normal and maximum oxygen demand, along with the corresponding air flows $Q_{A,b}$ and $Q_{A,b,max}$ (Eqs (3.33) and (3.34)), the sludge production M_{Total_sludge} (Eq. (3.43)) and the effluent quality over the evaluation period ($t_e - t_0$). These parameters can be used to calculate EQI, and as input for dedicated cost models for calculation of CAPEX and OPEX, and so NPV, using otherwise similar basic assumptions as those used for the steady-state calculation in Section 3.3.

3.5.1. Capital Costs

CAPEX evaluation demands pricing information, obtained from suppliers and/or based on costs provided by end-users for similar items of equipment at full-scale MBR plants (Table 3.33). The maximum plant capacity (Q_{peak}) relates to $J_{net,max}$, assumed to be 40% higher than J_{net} (Section 3.3).

Biological tank volumes (Eqs (3.20), (3.21), (3.25) and (3.26)) are determined by the required minimum HRT of 8 h at average influent flow conditions, or a minimum HRT at maximum flow conditions of 4 h, whichever is the larger, which correlates values reported for full-scale MBR plants (Chapter 5). V_m (Eq. (3.21)) is incorporated in V_{aer} (Eqs (3.25) and (3.26)). Each 10,000 m^2 of membrane area is assumed to demand one membrane tank, and it is preferred to have four or more membrane tanks to allow sufficient flexibility in operation and cleaning (Sections 5.3.1.2–5.3.1.6). A typical tank is rectangular, has a total height of 5.7 m and is filled to a 5 m depth.

A 6-mm coarse screening step followed by 0.75 mm fine screening is appropriate pre-treatment for an HF plant (Section 3.2.2.2), with the screens sized to treat Q_{peak} with 50% redundancy to allow one of the three screens to be shut down for cleaning or maintenance. An appropriately conservative value for SAD_m is 0.3 $Nm^3/(m^2\ h)$ (Section 3.2.2.1), and each membrane tank demands one blower with one more on standby. The biology blowers are sized based on $Q_{A,b,max}$ (Eqs (3.33) and (3.34)) to maintain a DO of 2 mg/L over the simulation period, assuming 50% redundancy. Pumps for biomass recirculation (as determined by Q_{mr}, Q_{int} and Q_w – Eq. (3.10)) and permeate (via Q_{peak}) are sized based on those typical of a large-scale plant, with one standby, as with the mixing equipment for the anoxic zones where one agitator is installed per 450 m^3 of V_{anox} (Eq. (3.20)) and one standby. Costs of land, civil engineering, other electrical equipment and construction are excluded, since they are all highly location-specific.

3.5.2. Operating Costs

OPEX relates mainly to energy demand for aeration, pumping and mixing, costs for sludge handling and disposal and chemical cleaning.

TABLE 3.33 Parameter Values and Key Assumptions for Determination of Capital and Operational Costs

Parameter	Unit	Value	Parameter	Unit	Value
J_{net}	LMH	20	*Assumptions for CAPEX calculation*		
$J_{net,max}$	LMH	28			
f_{anox}	–	0.33	Membrane cost	€/m^2	50
θ_x	d	25	Civil works		
r_{mr}	–	4	Structural concrete	€/m^2	400
r_{int}	–	3	Foundations	€/m^2	171
Membrane packing density: area per membrane tank, φ_{tank}	m^2/m^3	45	*Assumptions for OPEX calculation*		
SAD$_m$	Nm^3/(m^2 h)	0.3			
Blower inlet pressure, $P_{A,1}$	Pa	101,300	Mixing power demand	kW/10^3 m^3 V_{anox}	8
Blower outlet pressure, $P_{A,2}$	Pa	160,300	Energy cost	€/kWh	0.0918
Blower inlet temperature, $T_{K,1}$	K	293	Sludge treatment cost	€/ton of DS	150
Blower efficiency, ξ	–	0.6	Citric acid 50%	€/ton	760
Specific heat capacity of air, λ	J/(kg.K)	1.4	NaOCl 14%	€/m^3	254
Aerator depth, y	m	5	*Assumptions for NPV calculation*		
Total headloss in pipework, Δh	m	3			
Sludge pump efficiency, $\xi_{p,sludge}$	–	0.50	Membrane life	Year	10
Permeate pump efficiency, ξ_p	–	0.75	Inflation	%	3
Transmembrane pressure, ΔP_m	Pa	35,000	Discount rate i	%	6

3.5.2.1. Energy Demand

Aeration energy: By incorporating the dedicated aeration model (Eqs (3.28), (3.33)–(3.38)) in the plant model, the air flow rate required to meet biological aeration demand $Q_{A,b}$ (Eqs (3.33) and (3.34)) can be determined over the evaluation period, while membrane aeration is determined by A_m and SAD_m. The theoretical blower power consumption $W_{Aeration}$ is given by Eq. (3.39), and based on the parameters displayed in Table 3.31 a value of 0.025 kWh/Nm^3 of air is determined for the aeration energy demand. This corresponds well with literature values (Verrecht et al., 2008) and data from blower manufacturers. Total aeration energy is obtained by summing the membrane aeration and biology aeration energy demand values.

Pumping energy: Power required for sludge pumping, W'_{sludge}, can be derived (Eq. (3.40)), evaluated for the different sludge pumping requirements Q_{int}, Q_{mr} and Q_w (Eq. (3.10)). Using the parameters in Table 3.31, a sludge pumping energy demand E_q (Eq. (3.40)) of 0.016 kWh/m^3 of sludge pumped arises, assuming a linear dependency of W_{sludge} on Q_{sludge} (Eq. (3.41)). Additional power is also required for permeate pumping (W_{perm} – Eq. (3.42)) and backwashing (W_{bw} – Eq. (3.42)).

Mixing energy: A typical constant mixing power requirement of 8 W per m^3 of V_{anox} (Eq. (3.20)) is appropriate (Tchobanoglous et al., 2003), with no supplementary mechanical mixing required for the aerobic, membrane and buffer tanks.

3.5.2.2. Other Operational Costs

Sludge production ($M_{total\ sludge}$ – Eq. (3.43)) can be calculated from the change in total biomass in the tanks and the amount of biomass wasted over the evaluation period. Since reported costs for sludge handling and disposal vary widely and are location-specific, a broad range of values arise. A typical membrane cleaning protocol and frequency, required for chemical demand determination, comprises a weekly clean in place (CIP) with 500 ppm NaOCl and 2000 ppm citric acid supplemented with biannual cleaning out of place (COP) using 1000 ppm NaOCl and 2000 ppm citric acid.

3.5.3. Cost Sensitivity Analysis

3.5.3.1. Effect of Built-in Contingency: Hybrid Plant Versus Plant Designed for Maximum Flow

Table 3.34 shows a breakdown of costs for two scenarios:

(a) MBR component of a 'hybrid' plant (i.e. an MBR parallel to a conventional activated sludge plant): the MBR is designed to treat a constant daily flow,

TABLE 3.34 CAPEX, OPEX and Resulting NPV for an MBR Treating Steady-State Influent, As Part of a Hybrid Plant and an MBR, Designed for Maximum Flow without Buffer Tanks (Adapted from Verrecht et al., in press).

	Unit	MBR Part of Hybrid Plant	Plant Designed for Maximum Flow
Average plant influent flow	m^3/d	20,851	20,851
Maximum flow to the MBR	m^3/d	20,851	59,580
Total tank volume	m^3	6949	9975
Average plant utilization	%	100%	34%
Effluent Quality Index	kg PU/d	8364	5720
$COD_{average}$	mg/L	32.6	29.7
$NH_4\text{-}N_{average}$	mg/L	0.18	0.36
$NO_3\text{-}N_{average}$	mg/L	17.7	11.9
Total CAPEX	**€**	**4,070,432**	**7,135,044**
Screens	%	10.1	9.2
Membranes	%	53.4	62.4
Tank construction	%	26.6	20.2
Biology blowers	%	1.6	0.91
Membrane blowers	%	1.7	1.8
Permeate pumps	%	2.0	2.6
Mixing equipment	%	2.1	1.6
Recirculation pumps	%	2	2
Total OPEX	**€/year**	**649,266**	**882,364**
Energy	%	81.2	84
Sludge treatment and disposal	%	16.4	12.4
Chemicals	%	2.4	3.6
Net present value	**€**	**19,189,118**	**29,563,025**

while excess flow is treated by a conventional ASP that is not taken into account in the analysis such that reported NPV and EQI values represent only the MBR part of the hybrid plant); and

(b) a plant designed for maximum flow conditions ($Q_{peak} = 3Q$), whereby the average energy demand for the 'maximum flow' plant is about 40% higher (1.08 versus 0.77 kWh/m^3) due mainly to the under-utilization of the available membrane capacity, and the resulting excess aeration.

The contingency provided for changes in feedwater flow and composition impacts significantly on net present value (NPV). The analysis shows that any deviation from the ideal 'hybrid' plant, where the MBR treats a constant influent stream, leads to plant under-utilization and a resulting cost penalty manifested as an increase of up to 54% in NPV for a plant designed for three times the mean flow. The effluent quality (EQI) is lower for the 'hybrid' plant, which can mainly be attributed to higher average NO_3-N values arising from the smaller biological tank. The influence of changes in other operational and design parameters, over ranges typically encountered in full-scale plant operation, on the NPV and EQI evaluated for a default plant design (Table 3.35) is shown in Table 3.36.

3.5.3.2. Buffer Tank Versus Extra Membranes

Adding buffering capacity for flow equalization permits a smaller plant with a reduced membrane surface requirement and so greater mean plant utilization. A reasonable practical size limit to the buffer tank is 2 days (corresponding to 80% of the design flow). No influent is allowed to bypass the plant under storm conditions, so the combination of the buffer tank and plant capacity must cope with the maximum instantaneous flow of 59,580 m^3/d. Addition of a maximum size buffer tank yields a 21.8% decrease in NPV from €29.5 million down to €23.1 million, produced from a 9.2% decreased CAPEX, 20% decreased OPEX, and 22% increased average plant utilization (from 34 to 52%). Effluent quality, as indicated by the EQI, is largely unaffected and deteriorates by no more than 4% over the buffer tank size range considered (Table 3.36).

The cost of land required for the buffer tank is excluded from the analysis. However, for an additional required land projected value of less than €6.4 million it would always be beneficial to build a buffer tank. Assuming a total plant footprint equalling 2.5 times the combined footprint required for the biotanks and buffer tanks (Brepols, 2010; Brepols, Schäfer, & Engelhardt, 2010), a plant with the largest buffer tank requires 9693 m^2 extra land compared to one without. Land costs would have to increase to €660 per m^2 before addition of a buffer tank becomes economically unviable, demonstrating that whilst MBRs are regarded as being spatially frugal, there are palpable benefits to investing in flow equalization notwithstanding the increased footprint. Assuming a CASP to incur 2.7 times the footprint of an MBR, the combination of an MBR with the maximum sized buffer tank would actually

TABLE 3.35 Selected Base Design Parameter and Cost Values

Parameter	Units	Value
Design capacity	m^3/d	31,000
Maximum plant capacity*	m^3/d	43,400
Total tank volume	m^3	7233
Membrane area	m^2	64,583
SRT	d	24.3
Buffer tank size	m^3	13,071
Maximum flow out of buffer tank*	m^3/d	12,400
Max HRT in buffer tank	d	1.05
Effluent quality index	kg PU/d	5940
NH_4-N	mg/L	0.14
NO_3-N	mg/L	15.9
COD	mg/L	33.6
Net present value	Euro	€ 24,488,963

**As determined by design requirement that maximum sustainable flux = 140% of design flux.*

be ~10% larger than a CAS treating the same flow, though the latter would then demand tertiary treatment to achieve the same EQI.

3.5.3.3. *Influence of SRT and HRT*

A shorter design SRT decreases CAPEX due to decreased installed aerobic tank blower capacity at the lower MLSS concentrations and the resulting decreased aeration demand. However, the cost for the process blowers is less than 2% of total CAPEX, so the potential influence is negligible. The bulk of the reduction in NPV arises from OPEX, and specifically the decreased aeration demand at lower MLSS concentrations (due to its influence on the α-factor, Eq. (3.27)) at shorter SRTs. This cost benefit outweighs the cost penalty of increased sludge production according to this analysis, corroborating recent trends of working at lower MLSS concentrations (Trussell, Merlo, Hermanowicz, & Jenkins, 2006). However, this correlation is very sensitive to sludge treatment and disposal costs; costs of sludge tankering to a centralized sludge processing facility can be very significant at some sites, outweighing the energy cost reduction arising from lower SRT operation. Also, effluent quality requirements place a lower

TABLE 3.36 Sensitivity of NPV and EQI to Changes in Base Assumptions Defined in Table 3.35

	Net Present Value		EQI	
	M€	% Change	kg PU/d	% Change
Design parameters				
Buffer tank				
0 d HRT (No buffer tank)	29.6	21	5720	−3.7
2 d HRT (Maximum considered)	23.6	−5.6	5920	−0.34
Solids retention time (SRT)				
7 d	24.1	−1.3	6575	+11
56.1 d	26.3	+7.6	5562	−6.4
Hydraulic residence time (HRT)				
6 h	24.4	−0.4	6174	+3.7
15 h	24.9	+1.6	5243	−12
Costs				
Energy costs				
Annual increase of 4%	26.0	+6.2	5940	0
Annual increase of 7%	32.5	+33	5940	0
Sludge treatment costs (excluding hauling)				
43 €/ton of DS	23.0	−6.1	5940	0
259 €/ton of DS	26.0	+6.2	5940	0
Membrane costs				
20 €/m^2 membrane surface	20.8	−15	5940	0
100 €/m^2 membrane surface	35.0	+43	5940	0
Membrane costs − halving every 10 years	21.9	−11	5940	0
Membrane life − 5 years	31.0	+26	5940	0

limit on the SRT operating range, as EQI deteriorates when SRT decreases (Table 3.36), and operation at lower MLSS concentrations may lead to higher permeability decline rates (Trussell et al., 2006; Ouyang, & Li, 2009). Compared to the influence of SRT, the effect of HRT on NPV is minimal (if land costs are excluded), but a higher average HRT leads to improved effluent quality.

3.5.3.4. Energy Cost

Assuming an annual energy price rise of 4%, in line with the historical average between 1969 and 2008 (EIA, 2009), an increase in NPV of 6% arises compared to the base case where energy costs follow inflation. A 'worst case scenario' of 7% annual increase, corresponding to a doubling of energy prices roughly every 10 years, increases the NPV by 33%.

3.5.3.5. Membrane Replacement and Cost

As shown in Table 3.34, membrane costs make up the bulk (53−64%) of total CAPEX, while the other process equipment combined contributes about 20% to the total CAPEX. A 'worst case' membrane lifetime of 5 years (i.e. 6 membrane installations in the projected plant lifetime of 30 years) results in an NPV 26% higher than that of the base scenario with a membrane replacement every 10 years. A halving of the membranes cost every 10 years reduces the NPV by 10%; conversely, an increase of the membrane cost to €100 per m^2 of membrane area increases the NPV by 97% for a 5-year membrane lifetime and by 69% when membranes are replaced every 10 years. Since membrane replacement is such a decisive factor towards NPV, it is unsurprising that considerable attention has been paid to optimization of membrane lifetime by operating under a sustainable regime and developing adequate cleaning strategies. Recent developments suggest that an assumed membrane life of 10 years is not unreasonable, with membrane life of up to 13 years demonstrated at some plants (Porlock, United Kingdom, Section 5.2.1.1) and several other plants which have operated for more than 10 years without a membrane refit (Grand Targhee Resort, WY, USA; Thetis Lake Trailer Park, Canada).

3.6. OPERATION AND MAINTENANCE

3.6.1. Introduction

As already observed, the immersed MBR technology was introduced in the early 1990s, some 20 years after the commercialization of the first sidestream technology. A review of publications based on MBRs for wastewater treatment appearing in the scientific literature during the two decades following 1990 reveals a pronounced focus on fouling and foulant characterization (Section 1.5), and specifically extracellular polymeric substances (EPS). Other key research topics, such as nutrient removal and micropollutant fate, are most obviously driven by legislation.

Membrane fouling is generally perceived by the academic community as being the primary challenge to effective MBR operation (Sections

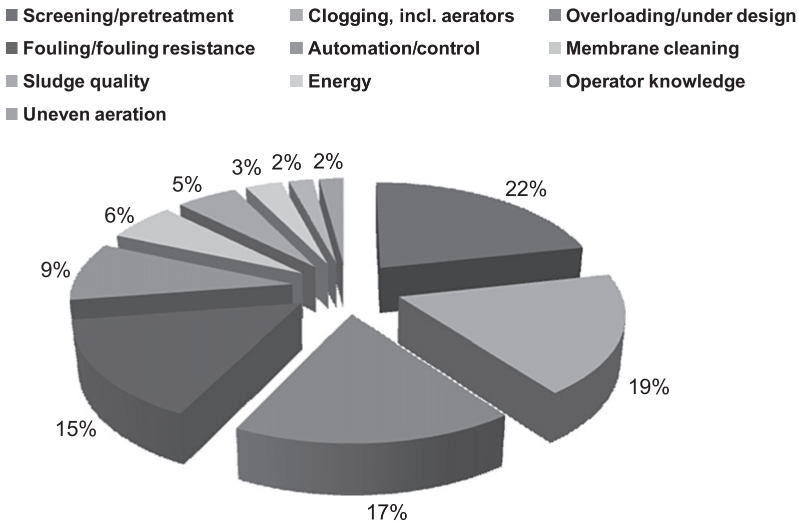

FIG. 3.7 Results of survey of MBR practitioners. *(Santos, Ma, & Judd, in press)*

2.3.6–2.3.9), and the nature of the work conducted is based on the premise that fouling refers to the reduction of permeability through deposition of EPS constituents at the membrane surface. However, a survey of 48 MBR practitioners, each independently challenged with the open question '*What is the main technical problem that stops MBRs working as they should?*' identified a number of topics which do not necessarily relate to fouling of the membrane surface (Fig. 3.7). Aside from economic, managerial and logistical issues, which are possibly generic, key technical limitations identified by this survey relate to the appropriate operation and maintenance of specific plant components or operations, plant design or process control. Key aspects specified include membrane aeration, over-optimistic design fluxes with insufficient contingency and, most significantly, screening – assumed to determine the degree of membrane clogging.

3.6.2. Clogging

Clogging is the agglomeration of solids within or at the entrance to the membrane channels. Whilst this is to be clearly distinguished from membrane surface fouling regarding both its mechanism and amelioration, the impact of both fouling and clogging is identical in that both are manifested as a decrease in the membrane permeability (Fig. 3.8). However, whereas fouling can generally be substantially removed through the application of an in situ chemical clean, i.e. cleaning in place (CIP), this course of action is not necessarily effective against clogging since in this case the materials are

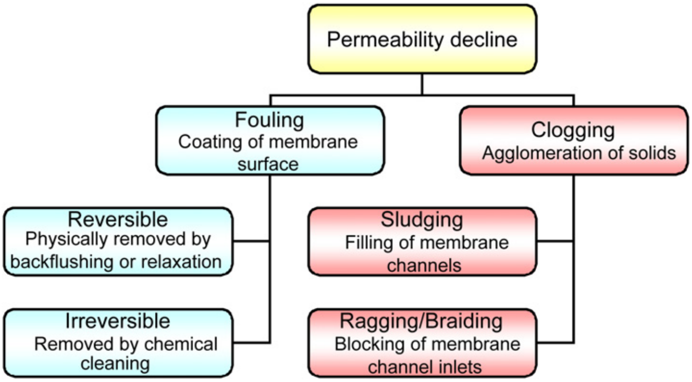

FIG. 3.8 Permeability decline phenomena.

physically lodged between the membrane surfaces rather than coated onto them. Severe clogging is generally only countered by removal of the membrane from the tank and cleaning the membrane modules individually with a low pressure hose. Such a level of manual intervention risks compromising the integrity of the fibres (Section 3.6.6).

As well as being indistinguishable in terms of impact, fouling and clogging can be related – as in the case of localized dewatering. When fouling takes place in a region of a membrane other unfouled regions become hydraulically overloaded. This can then cause rapid draining of the sludge in that region. If the forces causing solids deposition and sludge dewatering are greater than those associated with the scouring air then clogging (or 'sludging') takes place in that region.

Whilst clogging is inferred by an ineffective chemical clean, for FS modules clogging may also be deduced from a visual inspection of the top of the stack if the sludge level is lowered to expose the permeate outlet tubes and the top of the membrane panels. Clogging can sometimes be visible to the naked eye as a dark brown colour completely filling the 6–9 mm membrane channel. For transparent permeate outlet tubes, clogging may be inferred from a distinct dark brown discoloration which arises from extensive biofilm growth in the stagnant permeate water when no permeate flow takes place. Such an effect may also arise either from absolute fouling or from panels for which the membrane integrity has been compromised (Section 3.6.6), but such effects are much less commonly encountered than channel clogging.

Anecdotal evidence from operation of full-scale wastewater treatment works indicates clogging to be a more significant problem than fouling. In a survey of 15 European MBR plants published in 2008, eight had experienced problems of clogging (Itokawa, Thiemig, & Pinnekamp, 2008). Clogging within membrane channels has been noted in both of the main membrane configurations employed for iMBRs, hollow fibre (Fig. 3.9a) and flat sheet

FIG. 3.9 Clogging (sludging) of MBR membrane channels in (a) a hollow fibre and (b) a flat sheet module.

(Fig. 3.9b). In the case of municipal wastewater treatment the problem of clogging of membrane channels by gross particles in the MBR is exacerbated by their apparent tendency to agglomerate into long 'rags' or 'braids' up to 1 m in length (Fig. 3.10a) which may collect at the channel entrances (Fig. 3.10b). The rags appear to be made up primarily of cellulosic fibres, from bathroom tissue, and hairs. Such extensive agglomeration is referred to as 'reconstitution of rags', 'ragging' or 'braiding', and the occlusion of the channel entrances sometimes referred to as 'matting'. Rags may also agglomerate at the membrane aerator, which is extremely deleterious to the process since clogging rapidly ensues without scouring air to displace the solids from the membrane interstices.

There is currently no accepted non-intrusive method of assessing clogging propensity, other than (a) filtration of the mixed liquor through a coarse screen (3−6 mm) and (b) visual observation of aeration patterns in the tank. Whilst clogging impedes the passage of air bubbles passing through the membrane channels, there is a synergistic relationship in that reduced aeration encourages clogging. This underlines the importance of aerator design and installation, and specifically rigorous levelling of the diffusers to prevent poor air distribution between tanks or stacks. Also, small variations in water levels between reactors can significantly disturb air distribution, which is exacerbated in small plants equipped with less automation to remediate such imbalances.

Amelioration of clogging is primarily through the rigorous screening of the feedwater and, for the HF configuration especially, limiting the solids concentration in the membrane tank. Sludging/localized dewatering tends to arise only within some regions of specific units in some of the trains: it rarely

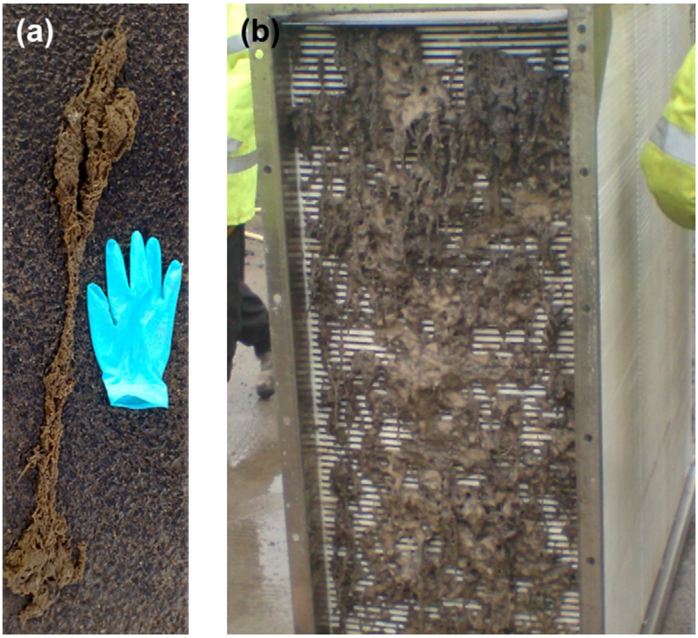

FIG. 3.10 (a) An ~80 cm 'braid' and (b) 'braiding'/'matting'/'ragging' of membrane channel entrances in a flat sheet module.

arises throughout the plant and can normally be attributed to local regions of high fluxes (>~40 LMH), inadequate air scouring and, in the case of hollow fibres where direct air souring of the membrane is more limited, high MLSS concentrations. A common scenario is that encountered during storm flows when the recycle ratios subsequently decrease to below 2 and the membranes become hydraulically overloaded. In the case of the FS membranes where permeation is driven by the hydrostatic head, an unregulated change in differential pressure across the membrane can cause high fluxes, which may then lead to clogging. On the other hand FS membranes are less prone to clogging from high MLSS: FS-based membrane thickeners operate at around 10 LMH and solids concentrations up to 4–5% without clogging problems, provided aeration is maintained and the flux is regulated by controlling the pressure differential.

Notwithstanding the paucity of information in this area, there are a number of aspects of clogging are self-evident:

1. The solids agglomeration rate in the channels relates to the rate at which water is drained from the sludge. This in turn is dependent on both the flux and the residence time of the sludge in the membrane channels, since the extent of dewatering increases at longer residence times.

2. The residence time in the membrane channel itself is directly related to membrane aeration, with respect to both the distribution of the air bubbles throughout the channels and the overall aeration rate.
3. Agglomeration must also depend both on the concentration and the characteristics of the particles, since particles which, for whatever reason, more readily adhere to the membrane and/or each other can be expected to agglomerate faster. These may be presumed to be partly related to feedwater physicochemical parameters, since these are known to impact on sludge quality (Section 2.3.6.1), and the physical nature of the inert solids specifically.

In fact, the same parameters which determine the extent of membrane fouling also similarly influence membrane channel clogging, and the manifestation of the two phenomena (reduced permeate flow) is also the same. It can only be speculated as to whether precisely the same chemical foulants which have been associated with fouling, such as colloidal polysaccharides or proteinaceous materials, are also responsible for particle agglomeration and/or irreversible deposition within the membrane channels. However, monitoring of the physical sludge characteristics can provide an indication as to whether incipient clogging is likely. The time to filter standard method 2710H (APHA, 1999), modified slightly with a smaller pore size filter paper, provides data on changes in sludge filterability and thus some indication of biomass health. Also, most obviously for immersed systems: (a) the solids concentration must be kept as low as possible, generally no more than 25% more than that of the biotank and a maximum of ~15 g/L for most HF systems; and (b) the membrane aeration system must be functioning correctly to ensure an even distribution of air over the membrane surface, possibly with increased aeration during storm flows. For pumped sidestream systems the high shear imparted by the crossflow permits rather higher MLSS concentrations and thus greater contingency, albeit at the expense of energy demand. Air-lift sidestream systems, on the other hand, are susceptible to clogging and matting, though the latter is apparently readily removed by periodic draining of the membrane tubes.

3.6.3. Pre-treatment

3.6.3.1. Screening

Screening is generally recognized as being crucial in suppressing clogging of both the membrane modules and the aerators. Whilst the standard rating of a screen at the inlet of a classical sewage treatment works is 6 mm, for an MBR the rating ranges from 3 mm for an FS membrane down to 1 mm or less for the HF configuration (Sections 3.2.2.2 and 2.3.9.1). The quantities of the screening generated in an MBR process are therefore considerably greater than that produced by conventional sewage treatment, and the management of this waste stream has to be taken into account.

A comprehensive testing programme of screens challenged with raw municipal wastewater has been conducted at the Chester Le Street WwTW in the United Kingdom (Thompson, & Marlow, 2003), and more limited smaller scale tests have been conducted in Germany (Frechen et al., 2007). The performance of different types of screens varies significantly for a given aperture shape, and available screen types for fine screening of sewage upstream of the MBR comprise bar, wedgewire and perforated plate. For municipal wastewaters in particular, screens with circular aperture (i.e. perforated plate) are preferred for rigorous removal of fibrous material which might otherwise pass through the slit-like apertures provided by bar or wedgewire screens. However, 1 mm bar or wedgewire screens provide a similar performance to 3 mm perforated plate; such screens may be considered more suited to small sites or industrial sites.

The other critical component of the screen system is the mode of cleaning. Most fine screens have automated cleaning, although coarse bar screens can be periodically manually raked clean. Brush cleaning of the screens produces a lower volume waste but the action of the brushes can force some fine material through the apertures. Washing of the screens can also introduce the risk of screenings being washed downstream. Backwashing offers the lowest risk of entrainment of fibrous matter, with band screens being the least exposed in this regard. Plants should be designed to prevent any bypassing of the fine screens; they must be operated and maintained correctly – preferably through a maintenance contract with the supplier. An advantage of a hybrid system is that excessive flows that would otherwise bypass the screens can be directed to the conventional activated sludge stream. For a classical MBR, on the other hand, duty and standby screens are essential to ensure that fine screening is never compromised.

However, some plants appear to be susceptible to ragging despite rigorous inlet screening and, in such cases, RAS screening would appear to be the only reasonable solution. Given that ragging has thus far attracted no research and quantitative information from full-scale plant is also extremely limited it is unclear as to whether such remedial action is effective. The company Eflo have fitted simple manually raked 10 mm bar screens to the RAS lines of two MBR plants in the UAE, having encountered problems with ragging (or 'Hair and Fibre Reinforced Biomass', HFRB) at smaller plants in the region over a prior 18-month period. There are now an increasing number of plants operating with RAS screening, specifically Swanage (Section 5.3.1.2), Heenvliet (Section 5.3.1.1) and Ulu Pandan (Section 5.3.1.7).

3.6.3.2. Other Pre-Treatment

Aside from screening, two other important pre-treatment steps are flow balancing and FOG traps (for fats, oils and grease). Flow balancing is highly desirable to limit the shock loads imparted on the MBR system, since rapid changes in both the hydraulic load and the F/M ratio can promote a number of deleterious effects, principally membrane fouling, foaming (Sections 2.3.6.3 and 3.6.5) and nitrification inhibition. FOG can similarly promote both foaming and fouling, with fouling arising from both direct adsorption of FOG onto the membrane and from EPS generated by the filamentous micro-organisms.

3.6.4. Fouling and Cleaning

As already discussed, a large number of things can lead to a diminution in flux or permeability. Generally it is change in the feedwater flow and/or quality, or that of the sludge directly, which causes changes in permeability. This includes temperature, hydrophobicity (possibly from FOG in the feed) and shock loads of salinity or toxic chemicals which may promote EPS generation. Other factors impacting on permeability through EPS concentration include a high F/M ratio and low DO concentration. The extent of potentially onerous colloidal fouling can be assessed through a comparison of permeate and supernatant COD, from standard centrifugation, which gives a measure of the levels of fine flocculant materials and colloidal particles in the sludge which are retained by the membranes, leading to membrane pore plugging. Such fouling may be ameliorated by chemical cleaning, but in such instances it is better to identify the root cause.

In general, surface membrane fouling is a greater operational impediment in industrial effluent treatment than in municipal and, also in most cases, is ameliorated primarily by chemical cleaning. The plethora of research into membrane foulant (and specifically extracellular polymeric substances or EPS) characterization in municipal wastewater treatment has arguably done little to inform actual operation and maintenance of full-scale plant. Generally for these applications the use of a combination of cleans based on hypochlorite, sometimes adjusted to an alkaline pH, and citric acid, or occasionally oxalic acid and in either case often supplemented with mineral acid, is ubiquitous (Section 3.2.2.3) and has been so almost since the installation of the first iMBR in 1990. Any departure from this practice can generally be attributed to clogging, when greater intervention is required, constraints on waste discharge, when hydrogen peroxide may be used instead of sodium hypochlorite, or changes in wastewater quality.

For industrial applications the range of candidate cleaning chemicals is more extensive, and may include detergent and chelating or anti-scaling chemicals at a pre-defined temperature and duration of application. Thus, whilst the use of hypochlorite is almost ubiquitous in MBR membrane cleaning, it is not necessarily the most effective reagent for some industrial applications where more foulant appraisal may be required, particularly for more challenging effluents and/or unusual membranes. For example, internal fouling of a ceramic membrane by iron from a landfill leachate source has been encountered, requiring an oxalic acid cycle to be included in the CIP procedure. Ceramic membranes generally appear to benefit more from mineral acid cleaning than hypochlorite.

The protocol of a clean in place can, in the case of HF membranes, involve repeated short backflush intervals (or pulsing) and hence resemble a chemically enhanced backflush (CEB, Fig. 2.11). The sequence of cleaning agents is usually (alkaline) hypochlorite followed by organic acid, and is particularly prevalent for municipal wastewater treatment. This arises because it is generally considered that finishing with an alkaline cleaner can promote precipitation of metal hydroxides and carbonate salts, and as such the acid clean should always follow an alkaline clean for waters containing significant concentrations of scaling compounds. However, reversing the sequence has been shown to be effective at some sites or for some membrane products. Maintenance cleaning is applied regularly — often twice weekly — for HF systems, and it is nearly always more effective to employ both reagents consecutively on every clean, rather than alternating between cleans. For FS systems, with no maintenance cleaning, recovery cleaning is usually applied based on a set threshold pressure, but also time limited if extended operation without reaching the threshold pressure is encountered. The cleaning frequency is then generally between quarterly and annually.

Other components of the system may also require cleaning; aerator flushing with sludge is normally conducted according to the manufacturer's recommendations, the standard frequency being daily and for each start-up of the blower for flushable centipedal or ring aerators typically used for FS systems. This is essential to remove any sludge which might otherwise collect inside the aerators and dry out in the air flow to form a tenacious solid residue. For HF systems using cyclic aeration, aerator flushing is not considered necessary.

Another important issue is the management of the chemical waste stream generated from chemical cleaning, and recovery cleaning in particular. For maintenance cleaning, provided the total load of sodium hypochlorite exerted is not too large relative to the bioreactor, it can be flushed into the mixed liquor (through displacement with permeate) and consumed by it without sacrificing significant biomass activity. This can generate EPS as a consequence of stresses imposed on the biomass, but this is generally not significant. In the case of recovery cleaning, where membranes are soaked in tanks filled with more concentrated cleaning reagent, the quantities of reagent involved are much larger. In such cases the chlorine residual can be quenched by dosing it with some of the sludge which has been displaced by the cleaning reagent. The spent waste reagent must then be disposed of appropriately, normally to the head of works. If quenching with sludge is not appropriate then chemically dechlorination with alkaline bisulphite solution may be necessary before returning the spent solution to the head of works.

Notwithstanding the general guidelines provided above and in Section 3.6.3, the control system should provide sufficient flexibility to allow different cleaning reagent concentrations and cleaning sequences to be applied. It can also be advantageous to study the impact of the head of sludge or water in the tank, since this imposes a back pressure which can influence the cleaning efficacy. Given that fouling is ubiquitous, that its precise nature cannot be predicted, and that chemical cleaning represents the only means at the operator's disposal for recovering permeability, it is important that the design of the CIP control system (Section 3.6.9) does not prevent exploration of different cleaning protocol options.

3.6.5. Foaming

Foaming in domestic wastewater treatment plants is attributable most often to imbalances in the *F*/*M* ratio causing bacteriological changes (Section 2.3.6.3). Other causes, such as shock loads of synthetic surfactants, are easily distinguished from *F*/*M*-based foaming and much less frequent in most plants. The most common type of foam, caused by filamentous micro-organisms, can be onerous to conventional sewage treatment because of its impeding of the conventional clarification process – a phenomenon referred to as bulking. However, excessive foaming is onerous to any biotreatment process. For MBRs, the negative impact is exacerbated by the association of foaming with an increased sludge fouling propensity due to the hydrophobic nature of the micro-organisms concerned and the elevated levels of EPS associated with them.

A high *F*/*M* ratio, such as that found in the start-up of an activated sludge process without any seed sludge from an existing process (or too little seeding), yields copious amounts of white foam which rapidly builds up to completely cover the aeration tank. This foam is inherently fragile, being dislodged by gusts of wind. This white foam tends to diminish as the biomass concentration increases and disappears as the target MLSS concentration is approached (>5–7 g/L) and the *F*/*M* ratio is in balance. This white foam can incrementally accumulate if the *F*/*M* ratio is rapidly increased, for example through wasting excessive amounts of sludge and consequently decreasing the MLSS concentration.

Conversely, a stable light brown foam can be promoted at low *F*/*M* ratios when the biomass population is extremely high. Such foaming is common, particularly in the membrane tanks and during periods of low loading such as overnight and at weekends. A thin scum-like layer on the bioreactors, and the anoxic zone in particular where microbubbles of nitrogen gas are released, is not necessarily problematic provided it does not accumulate. Indeed, it is actually advantageous in the membrane tanks to have a thin foam layer since this provides a useful indication of regions of reduced aeration, which is indicative of clogging (Section 3.6.2). However, such foaming becomes extremely onerous if it builds up to the point where it overflows from the tanks, forming a viscous sludge as the bubbles collapse. On returning to a balanced *F*/*M* ratio the accumulation will normally cease, but the foam disperses only very slowly over a period of days or even weeks. Over this period the foam condenses as air bubbles collapse and the layer turns a darker brown due to oxidation. In an unmodified system, the dispersal of the foam is only through the mechanical action of the aerated sludge on which it resides, which eventually re-suspends the solid particles.

Since foaming is a common problem in classical biotreatment processes foaming management strategies are reasonably well advanced. Strategies for tackling foaming include the following:

a. reduction of the MLSS concentration,
b. increasing the RAS rate,
c. nutrient dosing,
d. reseeding with imported sludge,
e. chlorine dosing,
f. alum dosing (for foaming relating to *Microthrix parvicella* in particular),
g. minimizing the membrane aeration,
h. antifoam dosing (using non-silicone-based reagents such as Nalco MPE 50 or CIBA Burst, though these are only marginally effective against stable brown foam) and
i. surface spraying.

Of these the first six actually address the possible cause of the problem, if it is microbiological in origin, rather than the symptom. On the other hand, the impact (if any) of surface spraying and antifoam dosing is usually immediate. Alum dosing works by blocking the fat-splitting enzyme in *Microthrix parvicella*, and is generally dosed at a concentration of 60 g Al^{3+} per day of sludge age per kg MLSS. Whilst it has been successfully demonstrated for some CASPs, it appears to be species specific and is ineffective against some filamentous micro-organism species (such as 021N and 1863), and also of limited efficacy at SRTs $<$ 8 d and low DO concentrations. Chlorine dosing, either through surface spraying or dosing directly into the RAS stream at concentrations of 100−150 mg/L, appears to be more effective for dissipation of Nocardia-based foam.

As with most other wastewater treatment processes, it is highly desirable for there to be the most appropriate design with sufficient contingency built into the plant, rather than be faced with reactive measures which prove to be ineffective. The most desirable design facets for reducing or mitigating the risk of foaming are the following:

(a) overflow from the membrane tank to the biotank,
(b) equalization of loads through buffering,
(c) installation of surface sprayers (although careful consideration should be given to odour and aerosol generation),
(d) good mixing characteristics, with no accumulation of anaerobic sludge or formation of septic regions,
(e) sufficient freeboard and/or surface wasting − with appropriate management of the wasted foam, which is much less readily dewatered than waste activated sludge (WAS),

(f) the option for supplementary dosing with nutrient, antifoam reagents, alum or chlorine,
(g) avoiding overly extended SRTs,
(h) careful control of nitrification/denitrification,
(i) DO monitoring, and selection of conservative low set points and
(j) adequate FOG removal, if appropriate.

Ultimately, foaming – along with many other facets of biological wastewater treatment – is a consequence of microbiology. Foaming arises primarily because the death of cells leads to the release of their DNA, which is proteinaceous and surface active and so causes foaming. Thus, anything causing extensive inactivation of the micro-organisms can produce foaming, including low temperatures, insufficient organic loads and excessive chemical cleaning. Given that at least some of these are occasionally unavoidable, it is advisable to have some contingency measures in place should foaming arise, if no other mitigation of risk is possible.

3.6.6. Membrane Integrity

The most commonly employed, and by far the most convenient, method for assessing membrane integrity is through on-line turbidity measurement. However, the efficacy of this measurement is significantly impaired both by shedding of colloidal particles from biofilms formed in the permeate stream and, more commonly, by air bubbles, both of which give false positives. Alternative methods include the following:

- pressure decay testing (for hollow fibres) using pressurized air in the permeate channels,
- suspended solids measurement,
- particle count,
- bacteriological count,
- SDI (Sild density index) test, if RO is installed downstream and
- visual inspection.

None of the above is entirely satisfactory, since these methods are predominantly off-line and particle counting is subject to the same false positives as turbidity monitoring. Pressure decay testing (PDT) is the standard accepted off-line method for integrity monitoring, and is certified by most environmental/water regulatory bodies. The rate at which the air pressure on the permeate side decays follows a logarithmic relationship with the rejection of micro-organisms (or a log:normal relationship with the LRV, the log rejection value). Compromised fibres may also be detected through the appearance of bubbles during a PDT. Severely compromised panels or modules can be identified simply through inspection of the individual permeate lines, provided these are transparent or translucent, or of the permeate via a sight glass.

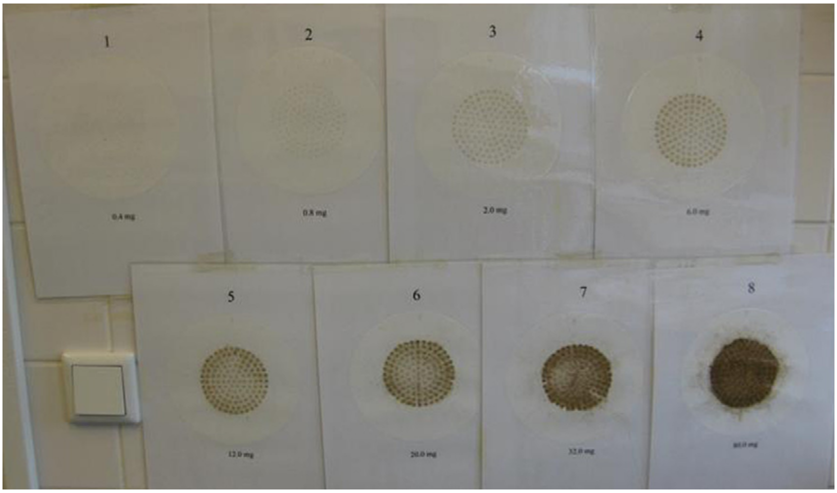

FIG. 3.11 Reference filter papers for membrane integrity assessment *(from the Heenvliet site).*

An alternative method for semi-quantification of the suspended solids in the permeate is through the creation of a set of filter papers used for the filtration of increasing quantities of sludge solids (Fig. 3.11), which can each be laminated and kept for reference. These provide an indication of the total quantity of sludge solids in the sample. Permeate collected from the sampling points can then be filtered by the same method, either by gravity or vacuum filtration, to create the sample filter paper. The filter is dried and the hue created by the sludge compared with that of the standards to allow an estimation of the sludge solids concentration. This method thus permits an assessment of the concentration of both fine and gross particles and is less vagarious than the on-line turbidity reading. It is also quite flexible, in that the amount of treated water filtered can be changed. Clearly, though, it relies on some recognizable link between the estimated product water sludge solids concentration and the LRV.

3.6.7. Biological Operation and System Shutdown

MBRs have traditionally operated at relatively long SRTs (>15 d) to take advantage of the improved biological efficacy, manifested primarily as higher nitrification rates and reduced sludge yields compared with conventional processes. However, this then leads to an increased risk of formation of some filamentous micro-organisms with the associated onerous impacts (Section 2.2.5), and a reduced oxygen transfer efficiency associated with the high MLSS concentration (Section 2.2.5). It is the latter which has caused increasing concern within the industry, given the commensurately increased energy penalty incurred.

Over the past 5–7 years there has been a noticeable trend towards lowering sludge age, particularly for large plants where the total energy saving may be significant and the increased sludge yield is made less onerous by the availability of on-site sludge management. However, a reduction in the SRT can increase the risk of nitrification failure as well as of fouling, although clogging is slightly ameliorated at the reduced solids concentrations in the membrane tank. The lower MLSS also reduces the buffering capacity of the reactor to shock loads. The latter can be significant if the plant receives trade discharges. There is therefore a balance of risks to be considered at the plant design stage, which then determines the operational envelope.

Seasonal shutdown of an MBR is easier to manage than that of a classical system because no washout of the biomass can take place. The MLSS level can be run down to similar levels to that of a classical ASP (3–4 g/L) to maintain the *F*/*M* ratio and then fed with a carbon source and/or nutrients to maintain it. Whilst this may lead to the release of EPS, this is of little consequence if there is little or no flow through the system. Re-seeding of an MBR, if necessary, demands that the seed sludge is screened to remove gross solids and rags in particular, if it is taken from a conventional municipal ASP reactor. MBRs incur a lower start-up time than conventional ASPs since they can be operated with no sludge wastage and no washout at any required HRT: the uncoupling of HRT and SRT is one of the key advantages of the MBR process. Whilst the microbial population may not reach steady state for a number of weeks, an MBR will generally produce effluent quality within specification a matter of days after restarting.

3.6.8. Industrial Versus Municipal MBRs

The concept of an industrial MBR does not differ from that of a municipal in that it must be designed to meet both the biotreatment and hydraulic objectives given the quality of the feedwater and, subsequently, the biomass and sludge generated. However, there are some key differences in the nature of the water to be treated which in turn determine differences in the design and thus the operating envelope (Table 3.37).

Industrial effluent flows tend to be smaller than municipal ones, making for smaller plants. The variation in flow is also less for industrial plants, with flows tending to decrease at weekends but otherwise remaining stable. For municipal plants, on the other hand, storm flows can exert huge changes in the hydraulic load. Municipal plants are thus sized for wet weather conditions, such that for much of the time there is a surfeit of membrane area. An advantage offered by the sidestream systems, both pumped and air-lift, is the ease with which the membranes can be taken off-line when not required, conserving both energy and membrane life. For large municipal iMBR plants, separate stormwater management is highly desirable to prevent storm flows to the MBR.

TABLE 3.37 Municipal versus Industrial Wastewater Treatment by MBRs

Facet/parameter	General Municipal Trend	General Industrial Trend
Feedwater		
Flow	Relatively high with significant diurnal variation, increasing during storms	Relatively low but stable, decreasing at weekends. Possible seasonal variation
Organic load	Relatively low with little variation	Relatively high with large fluctuations
Toxic shock	Occasional, particularly salinity shocks at coastal sites	Relatively frequent
Temperature	Seasonal variation	Relatively high with little variation
MBR		
Membrane and process configuration	FS/HF iMBRs; air-lift MT sMBR	All types; pumped sMBRs more common
Flux	Higher (generally >18 LMH)	Lower (generally <22 LMH)*
Fine screening	Critical for removal of fibrous material	Removal of gross solids

**for iMBRs; for sMBRs fluxes can exceed 150 LMH*

The generally higher organic loading at higher temperatures for industrial effluents can demand greater aeration intensity, possibly through jet aerators or even a pure oxygen supply, and thus greater aeration energy. This is especially so for pumped sMBR systems which tend to operate at higher MLSS levels and therefore inherently lower oxygen transfer efficiencies. Organic loading is also more variable for industrial effluents, demanding more prolonged buffering, and operating temperatures in excess of 50 °C demand that the bioreactor is acclimatized to thermophilic conditions.

Notwithstanding the higher operating temperatures, which lower the viscosity, industrial effluents tend to produce sludge of higher fouling propensity and thus operate at lower fluxes (for iMBRs) and permeabilities. Permeabilities are particularly low for pumped sMBR systems where fluxes are generally up to an order of magnitude (80−150 LMH) higher than those of immersed systems, whilst TMPs may be up to 50 times higher at 2.5−4 bar. Cleaning protocols may also differ more widely for industrial effluents, reflecting the more varied nature of fouling (Section 3.6.4). Scaling by hardness and other multivalent metal ions is exacerbated if substantial amounts of CO_2 are stripped from the bioreactor by the action of the aerators, causing an elevation in pH. The tenacity of scalants and other foulants can ultimately impact on the choice of membrane, since some materials (ceramic and PTFE, for example) are more highly resistant to aggressive cleaning chemicals, though ceramic membranes appear to be more susceptible to scaling by multivalent metal ions.

Maintaining biotreatment is usually a more significant issue for industrial effluents. The wide variation in effluent composition across the entire industrial sector means that dosing with nutrients, assimilable organic carbon or other supplements may be necessary to sustain the microbiology in some cases. Limited biodegradability of the organic carbon is a familiar problem in biotreatment of landfill leachate and pharmaceutical synthesis wastewaters, sometimes demanding multi-stage biotreatment. By the same token, the quality of the seed sludge has to be high, with screening to ensure that extraneous gross particles are removed.

3.6.9. Process Control

Process control is fundamental to all wastewater treatment processes, and an excellent précis of the subject as pertaining to biological processes as a whole has recently been provided by Olsson (2008). The design approach for MBR control systems is guided by the application/duty, and in particular the scale, and the membrane configuration. Large-scale commercial systems require stability of operation, optimum use of energy and chemicals, and robust control when challenged with changes in hydraulic and/or organic loadings. Pilot plants, on the other hand, require flexibility of control strategies and built-in dependable comprehensive data logging to allow different modes of operation to be studied and the performance monitored. Furthermore, while the selection of sensors and final control elements (i.e. pumps and valves, Lewis, 2007) is similar for the two system types, commercial systems are larger with a commensurately greater choice of control equipment. In contrast, for pilot systems there is a more limited pool of control components with respect to both manufacturers and design architecture.

Due to demand for longevity of membrane life linked with ever stricter contractual agreements, membrane suppliers routinely stipulate limiting values of key process operational parameters as part of their warranties. These primarily comprise permeability (i.e. TMP and flux) and cleaning chemicals, but often dictate the entire process control scheme revolving around the filtration sub-system via proprietary algorithms (Ishida et al., 1993; Liu et al., 2006; Bartels & Papouktchiev, 2008). This is especially so in larger systems, where warranty contracts are stringent and cost of membrane replacement can form a significant proportion of the operating cost (Section 3.5.3.5). Therefore in large MBRs, control system design needs to be focused on selection of dependable, high-end primary sensing elements and final control elements to try to minimize process failure risk (Table 3.38). Control design for pilot systems is challenging in that all possible variations in operation that may need

TABLE 3.38 MBR Failure Modes

ID	Stimulus	Primary Impact	Secondary Impact	Ameliorative Measure
A	OC shock	*F/M* increase	Increased foaming	Anti-foaming strategies Buffer tank, reactor oversizing
B	OC shock	Unprocessed sewage increase	Increased fouling	Chemical clean
C	Saline shock	Biomass stress/ inactivation, EPS generation	Increased fouling and possible foaming	Sewerage remediation See A–B above
D	Toxic shock	Biomass stress/ inactivation, EPS generation	Increased fouling and possible foaming	See A–B above
E	Feedwater dilution (e.g. storm flows)	Biomass starvation	Reduced water quality	OC/nutrient fortification
F	Sludge discharge failure	SRT and MLSS increase	Oxygenation efficiency decrease, clogging propensity increase	Sludge discharge repair*
G	Aeration blower failure	Solids accumulation in membrane channels	Sludging	Blower monitoring/ maintenance
H	Aeration port clogging	Localized solids accumulation in membrane channels	Localized sludging	Aerator flushing/ cleaning*
I	Screen failure/ overtopping	Increased levels of gross solids	Braiding/Matting	Screening efficiency monitoring RAS screening*
J	RAS pump failure	Solids concentration profile across tank	Localized sludging Denitrification failure	Pump monitoring*
K	Inappropriate tank dimensions	Solids concentration profile across tank	Localized sludging	RAS rate increase*
L	Backflush pump failure	Foulant accumulation	Increased fouling	Pump monitoring
M	Relaxation failure	Foulant and solids accumulation	Increased fouling and clogging	Motorized valve monitoring*
N	Membrane integrity failure	Impaired water quality	–	Membrane integrity monitoring and repair

(*Continued*)

TABLE 3.38 MBR Failure Modes—cont'd

ID	Stimulus	Primary Impact	Secondary Impact	Ameliorative Measure
O	Inappropriate chemical cleaning protocol	Reduced/zero permeability recovery after CIP	Increased fouling	Increase reagent pumping time Conduct optimization test
P	Overdosed hypochlorite during CIP	Biomass inactivation Foulant generation	Reduced water quality	Check sludge DO and permeate ammonia levels, then microbiology. Re-seed if necessary
Q	Solids agglomeration	Braiding		See I
R	System shutdown	Biomass inactivation	Clogging and fouling propensity increase	Maintain aeration, nutrient/OC dose. Re-seed with screened sludge if required

**External chemical clean of membranes normally required*

to be explored for system optimization or operational sustainability must be anticipated. This is usually accomplished programmatically by creating extensive lists of user selectable configuration options offering a multitude of process steps for inline and recovery cleaning as well as complete freedom to select each individual valve state, pump flow/speed setpoint and time duration for the process step concerned.

Currently, polymeric membranes for MBR applications have permeabilities and recommended operating fluxes that translate into operating pressures generally ranging from ~70 to 140 mbar for immersed systems following a chemical clean, and up to 600 mbar for sidestream systems. Some membrane modules are not designed to tolerate back pressure. Therefore measurement of pressure should have a precision of 7 mbar or better. The transition from individual process cycles (i.e. from filtration to backflush and back again) has to be accomplished with careful sequencing of actuation of valves and pumps so as to avoid pressure surges exceeding 70 mbar, which would then exacerbate fouling. In this regard, it is favourable to have progressive flow evolution at the beginning and at the end of the filtration step. Ultimately, sustained operation relies on arresting any permeability decline, manifested as an increase in the TMP at constant flux. In systems of 0.4 MLD or more nominal flow capacity, a limiting rate of change of 50–100 mbar/s between the physical cleaning period and the filtration cycle is desirable. Excessive sudden hydraulic loads during this part of the sequence, leading to correspondingly large rates of change in the TMP, can encourage irreversible fouling. For pilot plants the pressure transient can be higher − 100−200 mbar/s.

For large systems with membranes operating under vacuum in filtration mode, initial priming of the filtrate suction pump and pipe headers connecting the membranes is accomplished at constant vacuum pump pressure, usually with dedicated air priming pumps. When the filtrate lines are primed and air extracted, the priming pumps are isolated and the filtrate pumps engaged. Since the filtrate is removed by vacuum, gas tends to precipitate from solution and coalesce in the pump housing. For most pump designs this causes a rapid loss of liquid flow. The underlying reason for this is that if the pump speed increases unconstrained then sudden air displacement takes place associated with rapid liquid flow. This in turn generates a significant pressure surge on the filtrate side of the membrane that can compromise membrane integrity or else accelerate the membrane fouling rate. To properly detect such events, the control system needs to scan pressure at least once every 100 ms, demanding fast-acting pressure transducers. Closed loop control for the filtrate flow or pressure should sustain the filtrate pump speed and prevent any sudden change in pump speed; slow stepwise acceleration of the pumping can be introduced in attempt to correct the flow. Graduation of the rate of change at the final control elements responsible for production or regulation of the filtrate flow is an essential part of control system narratives.

For larger systems, membrane modules are arranged as several parallel trains, each with its own membrane compartment. Often a number of these trains are fed with single set of variable frequency drive (VFD) feed pumps (Fig. 3.12). In such cases the flow for each train is controlled individually by a dedicated flow regulating valve with its own controller. Although possible, feed flow provided by pumping at high pressure with severe throttling to control flow to the individual trains is to be discouraged. Such an arrangement is more difficult to regulate and can lead to valve clogging, and the shear imparted can promote floc breakage and so membrane fouling from the

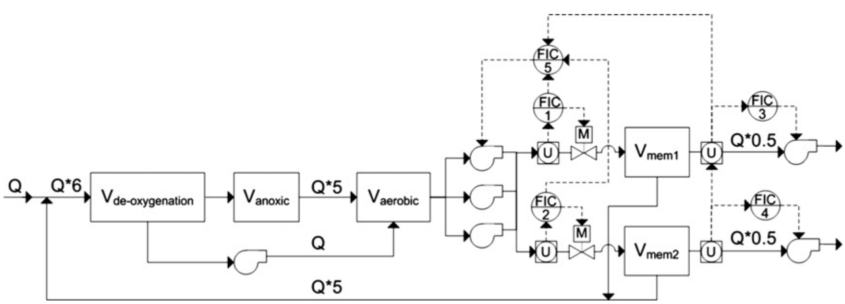

FIG. 3.12 An example of a piping and instrumentation diagram (P&ID) for a single train, with two membrane compartments fed by three pumps and the return from the membrane tank directed to the head of the train.

released EPS. This can be avoided by controlling flow distribution across the membrane trains with only light throttling (i.e. predominantly open valves) with the feed pumps set at a speed sufficient to achieve the necessary combined feed/crossflow for all trains. Subsequent redistribution of flow among trains is achieved by moderate and slow feed valve actuation. Since reaching balanced flow conditions takes longer than achieving the required combined feed flow, it is advisable to keep the feed flow unchanged during physical cleaning rather than attempting to save energy by reducing the feed flow pumping rate. This does not apply to chemically enhanced backflushes (i.e. maintenance cleaning), since in this instance it is necessary to preserve a high chemical concentration near the membrane surface which would be disturbed by the crossflow.

Offline chemical cleaning efficiency is crucial for preserving membrane permeability within the prescribed limits of the manufacturer. Maintaining constant temperature, liquid stream velocity and chemical concentration for the duration of the cleaning process is achieved by precise closed-loop control, with a response time – the time taken to reach 63% of the setpoint value – no faster than one-third of the cleaning loop HRT. The time limit is arbitrary, but it is the approximate time taken for any reaction, from the point of introducing the cleaning reagent, to be manifested. This may include demand on the chemical reagent imposed by the sudden dislodging of significant layers of fouling material from the membrane. Failure to impose this delay before responding can cause overdosing of chemicals or reaching terminal temperatures above manufacturers' recommended limits.

Optimization of operation on MBRs is based on ability to measure process parameters related to dynamics of substrate utilization, microbial growth and decay rate, membrane fouling and energy use. As a first step, the recorded temperature-compensated permeability is compared to the value predicted from maximum and minimum permeability limits following and prior to a chemical clean (assuming a linear decrease in permeability between cleans and based on a desired cleaning frequency – possibly once or twice a week for a typical maintenance clean). For a faster than predicted decrease in permeability, sustained over a period of several hours, adjustment can be made of the appropriate O&M parameters, such as flux, filtration time, membrane aeration rate intermittency or the protocol for the following maintenance clean. Conversely, if the fouling rate is lower than predicted then the flux can be increased and/or the aeration rate decreased.

Additionally, implementation of inline measurement of MLSS and oxygen uptake rate enables monitoring of the specific substrate utilization rate. If, due to diurnal variation in flow, a shift in organic loading takes place, a reduction in the flux may be possible as a means of mitigating the increased fouling tendency caused by the release of EPS. Furthermore, deviation in specific substrate utilization rate may be indicative of either a toxic load or nutrient deficiency. Finally, in the case of a rapid loss of permeability, to attempt to prevent costly and time-consuming off-line recovery cleaning, the MBR control system should actuate pre-emptive inline chemical cleaning at a higher frequency to recover permeability. This can be combined with reduction in flux or recovery, whichever parameter is considered variable (Bartels & Papuktchiev, 2008). It is common to pre-negotiate direct implementation of ready-to-use algorithms provided by membrane suppliers on royalty-free bases on procuring their MBR products.

3.6.10. Overview

It must be conceded that, notwithstanding the advantages offered by the technology, an MBR plant is significantly more complicated, both in design and operation, than a conventional activated sludge plant and therefore more exposed to a greater risk of process failure (Table 3.38). As such there is a greater onus on the operators to engage in the maintenance of all the constituent parts, including both the membranes and peripheral equipment such as the screens, blowers, motorized valve drives and process instrumentation.

It is also the case that more conservative operation − less variable and lower applied fluxes and loads coupled with higher and/or uniformly applied membrane aeration rates − alleviates the primary causes of process performance deterioration, these being fouling and clogging, and that rigorous screening retards the latter. Anecdotal evidence from full-scale plants suggests that it is those plants operating under such conditions which are subject to significantly less unscheduled manual intervention. Given the energy penalty incurred by higher membrane aeration rates, it would seem that the industry may in the future have to contend with higher capital costs associated with both larger membrane area and flow equalization to allow operation at lower fluxes and more regulated loads.

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Chapter 4

Commercial Technologies

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*PURON® is a registered trademark for the Koch submerged membrane module in Germany and other countries.

4.1. INTRODUCTION

Developing and commercially available membrane bioreactor (MBR) membrane module products can be classified according to membrane configuration (Section 4.3). Products based on flat sheet (FS), hollow fibre (HF) and multitube (MT) configurations are described in Sections 4.2, 4.3 and 4.4, respectively. Many such products exist and many more are being developed; a comprehensive description of all technologies available globally is not possible given that new products are almost continually being brought to market. However, most of those visible at the time of writing, and for which the basic technical details of the membrane element (Appendix C) were available, are included. In addition, established technologies based on specific membrane modules are also described at the end of each section.

4.2. IMMERSED FLAT SHEET (FS) PRODUCTS

4.2.1. Kubota

The Kubota membrane module was developed in the late 1980s in response to a Japanese Government initiative to encourage a new generation of a compact wastewater treatment process producing high-quality treated water. The first pilot plant demonstration of Kubota membranes was conducted in 1990, prior to the first commercial installation soon after. As of August 2009, there were more than 3300 Kubota MBR plants worldwide with a total installed (average) capacity in excess of 900 megalitres a day (MLD).

The original FS microfiltration (MF) membrane, type *510*, which is still widely used, comprises a 0.5 m × 1 m flat panel, 6 mm thick, providing an effective membrane area of 0.8 m^2. The membrane itself is a hydrophilicized, chlorinated polyethylene (PE) membrane, supported by a very robust non-woven substrate (Fig. 4.1), which is welded on each side to plate with a spacer material between the membrane and plate. The plate contains a number of narrow channels for even collection of the permeate across the surface. The average membrane pore size is 0.2 μm (normally rated 0.4 μm) but, due to the formation of the dynamic layer on the membrane surface, the effective pore size in operation is considerably lower than this and can be in the ultrafiltration (UF) range.

Type *510* membrane panels (Fig. 4.2a) are securely fitted into a membrane case (Fig. 4.2a) to form a module (Fig. 4.3), providing equal spacing of the membranes across the module. The spacing, combined with a 3-mm-rated bi-directional screen, has been found to be important in suppressing clogging between the panels. Flow from outside to inside the panel is either by suction or, more usually, under gravity – routinely between 0.05 and 0.13 bar hydrostatic head. Permeate is extracted from a single point at the top of each

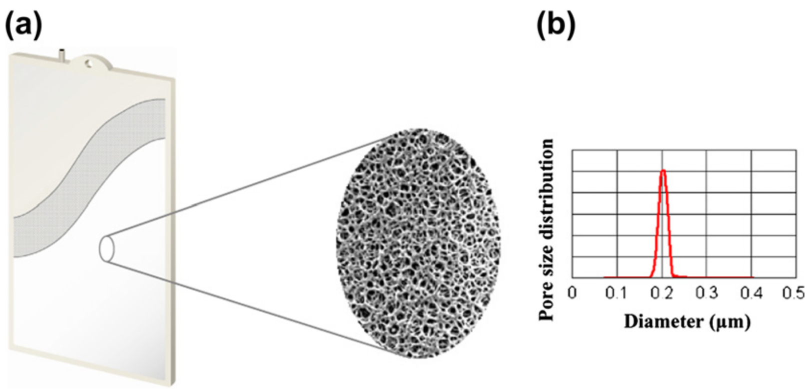

FIG. 4.1 Kubota membrane: (a) substrate and membrane surface and (b) pore size distribution.

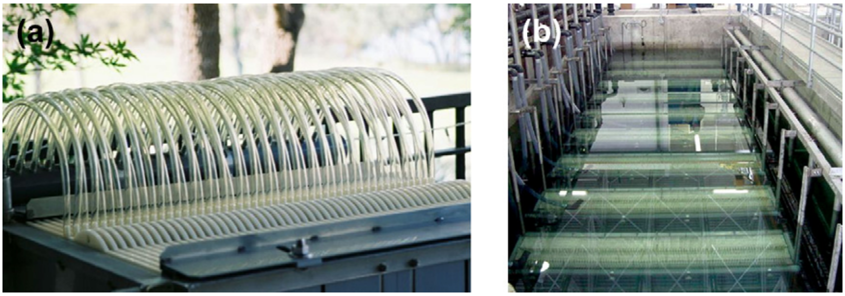

FIG. 4.2 Kubota *510* membrane panel in the membrane case (a) and (b) in situ *(photos courtesy of Ovivo).*

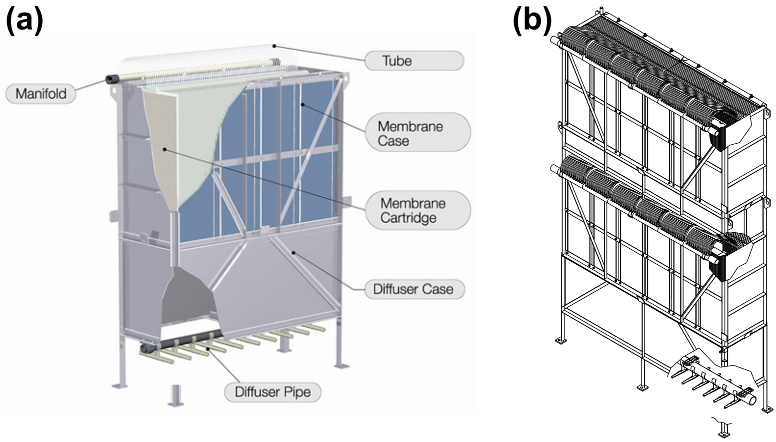

FIG. 4.3 Kubota: (a) single-deck (*ES* Model) and (b) double-deck (*EK* Model) modules, based on the *510* membrane panel.

membrane panel via a transparent tube (Fig 4.2a). Aeration via coarse bubble aerators is applied at the base of the tank so as to provide some oxygenation of the biomass in addition to aerating the membrane module. The original loop-type aeration pipe has been largely superseded by a patented sludge flushable aerator. This 'centapedal' aerator comprises a central pipe with smaller open-ended lateral branch pipes at regular intervals (Fig. 4.3). Cleaning of the aerator is achieved by briefly opening an external valve connected via a manifold to the ends of the central pipe(s). This allows vigorous backflow of sludge and air into the tank. This backflow clears sludge from within the aerator and helps to prevent clogging of the aeration system. To prevent air bubbles from escaping without passing through the membrane case, a diffuser case is fitted, in effect providing a skirt at the base of the module.

Membrane modules are presented in single- or double-deck configurations (*ES* and *EK* series, respectively; Table 4.1). The single-deck *ES* series

TABLE 4.1 Specifications for Kubota Panels and Modules

Membrane or Module Proprietary Name, Model	*510, ES*	*510, EK*	*515, RM*	*515, RW*
Panel dimensions, length × width × thickness, mm	1020 × 490 × 6	1020 × 490 × 6	1560 × 575 × 6	1560 × 575 × 6
Panel effective membrane area, m^2	0.8	0.8	1.45	1.45
Module dimensions, length × width × height, mm	1140–2920 × 600–620 × 2030	2200–2920 × 600–620 × 3500	2250–2930 × 575 × 2490	2250–2930 × 575 × 4290
Number of panels per module	75–200	300–400	150–200	300–400
Total membrane area per module, m^2	60–160	240–320	217–290	435–580
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	0.75	0.53	0.42	0.29
Launched	1991	2003	2010	2009

(Fig. 4.3a) has been employed in package plants, decentralized systems and small industrial plants. The introduction of the double-deck *EK* design in 2002 (Fig. 4.3b) provided improved efficiency for medium-sized plants. Capital costs per unit membrane area have decreased since the doubling of the specific membrane surface area to plant footprint, and the number of diffuser cases required halved with a single membrane case used for mounting two banks of panels. In addition, the operational costs were reduced due to a 30% decrease in specific aeration demand over that incurred by the single-deck ES series. By August 2009, there were over 200 double-deck membrane plants in operation, treating more than 600 MLD.

In 2009, Kubota developed a larger panel, type *515* (Fig. 4.4a), and the new *RW* series for medium and large municipal plants. This panel provides a total area of 1.45 m^2 and has two nozzles to improve the hydrodynamics of permeate extraction. The *RW400* membrane case incorporates 400 panels and 580 m^2 surface filtration area. The increased panel length and higher design flux reduce both the footprint and the membrane aeration demand; the recommended guide value of SAD_m for the double-deck module is 0.29 $Nm^3/(m^2\,h)$ for the double-deck module (Fig. 4.4c). At a total height of 4.3 m, the module is around 2.1 times the height of the original *ES* module. The operating conditions, the materials and the concept design of the original and new modules are otherwise the same. The first MBR plant based on the double-deck *RW* module began operation in May 2009. Kubota launched the *RM*

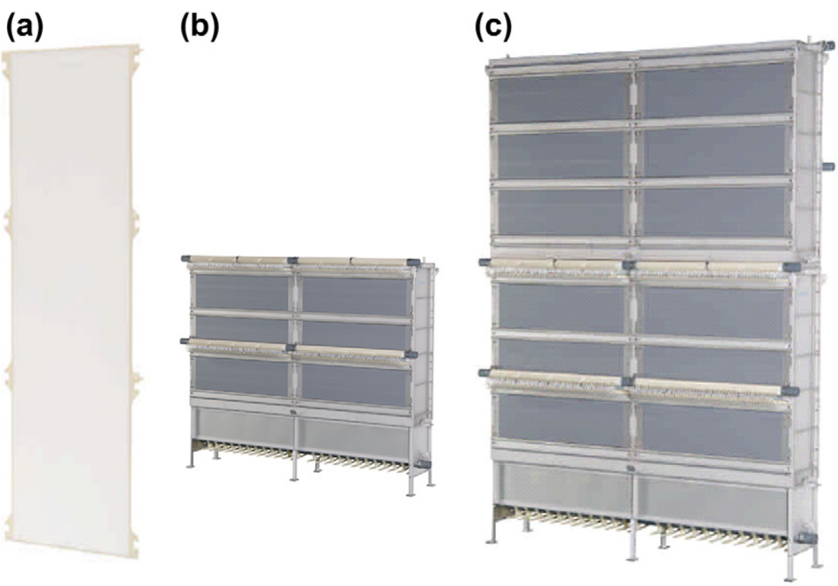

FIG. 4.4 (a) Kubota *515* panel, (b) *RM* single-deck module and (c) *RW* double-deck module.

series (Fig. 4.4b), single-deck modules accommodating type *515* membrane cartridges, in early 2010, following successful sales and operation of the *RW* series.

4.2.2. A3

The German company A3 Water Solutions GmbH began life in year 2000. A3 stands for *Abfall–Abwasser–Anlagentechnik* ('Waste sewage system technology'), and the company specializes in membrane module production, small wastewater treatment systems and treatment of fermentation residues. Their *MaxFlow* range of products comprises the *M70* and *U70* membrane modules, the former being based on a 0.14-μm PVDF membrane and the latter on a 150-kDa (~0.07 μm) PES membrane, which are designed for MBR duties. Both membranes are formed into panels 1040 mm long, 700 mm wide and 6 mm thick, providing an area of 1.36 m^2. A module (Fig. 4.5a) comprises 51 panels, which therefore provides a membrane area of almost 70 m^2, and operates at a recommended SAD_m of 0.69 $Nm^3/(m^2\,h)$ for a single deck and 0.35 for a double deck (Fig. 4.5b). The maximum recommended transmembrane pressure (TMP) is 0.25 bar and the panels are backwashable at a maximum pressure of 0.05 bar, made possible by using the filtrate pump in reverse operation. There are currently more than 40 installations based on these products, including containerized systems in army field camps and ship-board wastewater treatment plant as well as many other municipal and industrial applications. The product has undergone extensive

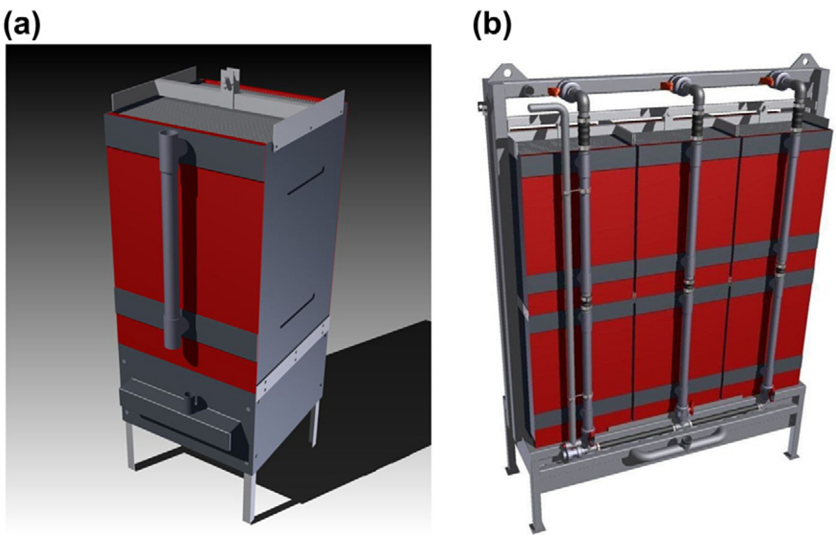

FIG. 4.5 The A3 module: (a) single and (b) double deck.

testing at Anjou Recherche in Paris (Grélot, Weinrich, Tazi-Pain, Lesjean, & Trouvé, 2007).

4.2.3. Alfa Laval

Alfa Laval is a global provider, founded in 1883, of specialized products and engineering solutions based on its key technologies of heat transfer, separation and fluid handling. The company has approximately 12,000 employees in more than 50 countries and operates in around 100 countries. The membrane activity was originally established in 1965 as DDS Filtration – one of the first European membrane companies. DDS was part of the Dow company from the mid-1980s until 1997 and was eventually acquired by Alfa Laval in 2002.

The company's *Hollow Sheet* MBR product is based on a 0.2 μm pore PVDF membrane and was introduced in 2006. The membrane sheet (Fig. 4.6a) is roughly square in aspect, almost 1.2 m wide, and has a membrane area of 1.81 m^2. Modules of 85, 170 and 255 sheets are possible providing areas of 154, 308 and 462 m^2, respectively, for the single, double (Fig. 4.6b) and triple decks, with no separation between each deck required as permeate is discharged through the sides of the element rather than through an extraction tube. The separation between the sheets themselves is 7 mm. The sheets are non-rigid, with the perforated spacer comprising a series of open channels to aid permeate transport within the sheets towards each side to the

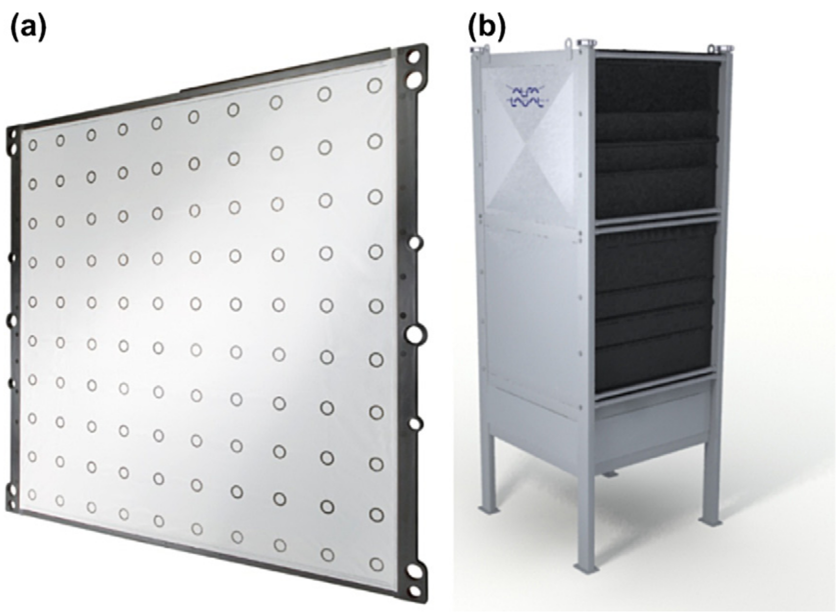

FIG. 4.6 The Alfa Laval: (a) modular sheet and (b) double deck module.

built-in permeate collector frame. The modules demand a minimum aeration rate of 0.48, 0.3 and 0.24 $Nm^3/(m^2\ h)$, respectively, for the single, double and triple decks. They have a typical operating TMP range of 0.01−0.04 bar at average flux 10−30 LMH, up to a TMP of 0.08 bar at peak flux. The sheets are flushable by circulating water or CIP solution from one side of the module to the other via the spacer. This is done by slightly overpressurizing up to 0.01 bar.

The modules are employed by a number of contractors and system builders incorporating MBR technology in plants in France, Italy, the Netherlands and the USA among others. The first full-scale MBR plant based on the technology, at 0.48 MLD design flow capacity, was commissioned at a modified potato starch production facility in Denmark in March 2007. Currently more than 20 full-scale and pilot plants are in operation or under commissioning.

4.2.4. Brightwater

Brightwater has been established since 1990. The company designs, supplies and commissions plants for the treatment of sewage, industrial wastes, water and sludges. Brightwater is part of the FLI Environmental group, who acquired the company from Bord na Móna Environmental in November 2007, Bord na Móna having originally acquired Brightwater in July 2000. The Brightwater *MEMBRIGHT*® system (Fig. 4.7) is an FS immersed MBR (iMBR) with 150 kDa (~0.07 μm) polyethylsulphone (PES) membranes mounted on a rigid polypropylene (PP) support. The module is almost square in aspect, 1120 mm in length, 1215 mm in width for a 50-panel unit (715 mm for 25-panel unit) and

FIG. 4.7 The Brightwater *MEMBRIGHT*® module.

1450 mm high. The respective membrane areas provided by the two sizes are 46 and 92 m^2, with each square panel 950 mm in length, providing an area of 1.84 m^2. The membrane spacing is ~9 mm, the panel support spacing being 10 mm, and the panels are clamped in place within a stainless steel frame to form the module. The module is fitted with an integral aerator which ensures even distribution of air across the module.

There were five installations based on the technology as of the end of 2009, mainly in Southern Ireland. The system has also been engineered and marketed for single household applications using modules containing six double-sided plates, each 400 mm wide by 600 mm deep, and several such units have been sold in the USA for this purpose.

4.2.5. Colloide

Colloide Engineering Systems (CES) provide various treatment technologies for water and wastewater in both the industrial and municipal sectors, concentrating mainly on small-to-medium-scale plants. The company first introduced the MBR technology in 2001. The CES *SubSnake* system is unusual in that the membrane modules are bespoke and fabricated from a continuous 0.04 μm PES membrane sheet which is cut to size and then glued at the edges to form an FS module. The membrane is then wrapped, snake-like, around a purpose-built steel or plastic frame comprising a number of rigid vertical poles at each end to make a multiple FS module with a membrane sheet separation of 10 mm. A single tube is inserted into the permeate channel of each FS sheet for permeate extraction under suction into a common manifold (Fig. 4.8a). The maximum depth of the module formed from this serpentine arrangement is dictated by the width of the sheeting, and the total membrane area of the module by its overall length. Thus far, the largest modules offered by the company are 10 m^2 (Fig. 4.8b), provided by 10 sheets of 1 m depth and 0.5 m width.

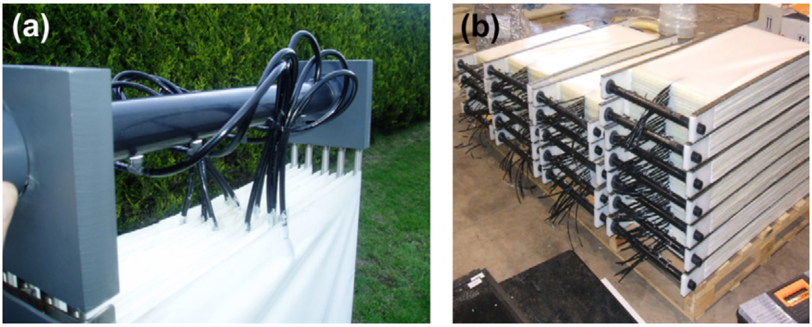

FIG. 4.8 The Colloide *SubSnake*: (a) panels, showing the permeate outlet tubes and manifold and (b) the 10 m^2 modules.

4.2.6. Ecologix: *Ecoplate*™ and *EcoSepro*™

Ecologix Technologies Asia is a Taiwanese manufacturer of both FS and HF membrane modules. Its membranes are provided by Sepromembrane Inc. in California, and the company manufactures the membrane module under licence from Ecologix Technologies in the USA. The product range includes membrane fine bubble diffusers and a rotary drum screen − the key process components of an iMBR. The company supplies two FS products: the *EcoPlate*™ and the *EcoSepro*™. Both products can be fitted with either PES or PVDF membranes of pore sizes of 0.08, 0.1 and 0.4 μm. The recommended aeration rates for the modules are 0.72−1.0 $Nm^3/(m^2\,h)$ for the *EcoPlate*™ and 0.5−0.8 $Nm^3/(m^2\,h)$ for the *EcoSepro*™.

The *EcoPlate*™ is based on a rigid 6-mm-thick ABS plate with the membrane supported by a PET non-woven substrate. The membrane is glued and ultrasonically welded to the backing plate, which is fitted with a single nozzle for permeate extraction. The panel dimensions are 1000 × 490 mm and provide a membrane area of 0.8 m^2 (Fig. 4.9a); the panels are fitted in a cassette to give a spacing of 6 mm. Modules (Fig. 4.9b) can be both single- and double-deck in design, with up to 400 panels for the 3.4 m tall double deck, and have a recommended maximum TMP of 0.3 bar.

FIG. 4.9 The Ecologix *Ecoplate*™: (a) panel and (b) module.

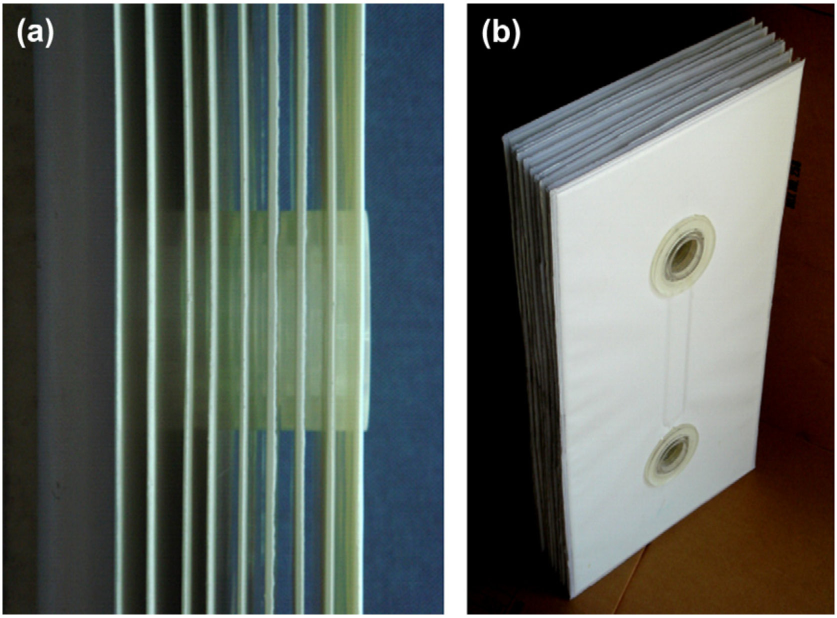

FIG. 4.10 The Ecologix *EcoSepro*™: (a) permeate collector and (b) panels.

The *EcoSepro*™ comprises a self-supporting membrane sheet. A dual 50 mm diameter permeate collection manifold runs through the centre of each panel, and the panels are fitted with plastic gaskets to allow a resin seal to be made (Fig. 4.10a). The panels themselves are non-rigid and only 3 mm thick, permitting high packing densities, and are 1000 × 320 mm, providing a surface area of 0.6 m^2. As with the *EcoPlate*™ system, the panels are separated by 6 mm in the module (of up to 720 panels, Fig. 4.10b), and the module operates at a maximum TMP of 0.3 bar. The decreased panel thickness decreases the module footprint (m^2 membrane area per m^2 base cross-section) by a factor of two to around 340 m^2/m^2 for the double-deck module.

4.2.7. Huber

Huber SE is a German company of more than 175 years standing with over 900 employees worldwide and a turnover of over $150m. The company provides water and wastewater treatment technologies, and is primarily known within the water sector for its screening and sludge treatment equipment.

The Huber Vacuum Rotation Membrane (*VRM*®) product is differentiated from almost all the other MBR systems by having a moving membrane module, which rotates at a frequency of 1–2 rpm. The small shear created by this, combined with the intensive scouring of the membrane surface by air from the central coarse bubble aeration, apparently obviates chemical cleaning. The

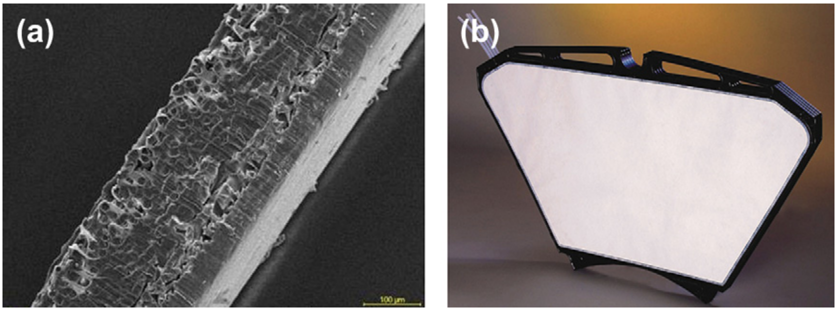

FIG. 4.11 The Huber *VRM*® membrane: (a) material and (b) element.

rotary action means that all areas of the membrane receive air scour, such that the solids do not collect in any region of the module.

The membrane material is based on 0.038 μm pore size PES material of around 300 μm in thickness (Fig. 4.11a). The membrane elements themselves comprise a four-plate segment of a hexagon or octagon (Fig. 4.11b), thereby making up one-sixth or one-eighth of a complete plate. The individual elements are thus relatively small (0.75−1.5 m^2 for one plate; four plates create one module of 3−6 m^2) and, since each is fitted with a permeate extraction tube, the permeate flow path is relatively short. The plates themselves are 6 mm thick and separated by 6 mm. The membrane modules are assembled into the *VRM*® units of 2 m (*VRM*® 20) or 3 m (*VRM*® 30) diameter. Each module is positioned and fixed in a drum and then connected to the collection tubing. The mixed liquor is taken from the aeration tank and is circulated around the unit and between the plates where permeate water is withdrawn. The water flows via the tubing into the collecting pipes (Fig. 4.12a). The pipes are joined to the central collection manifold, and the product water is discharged from this central pipe. The whole module is aerated by two coarse bubble aerators which are placed in the middle of the membrane unit (Fig. 4.12b) at a recommended membrane scouring rate of 0.15−0.25 $Nm^3/(m^2\,h)$.

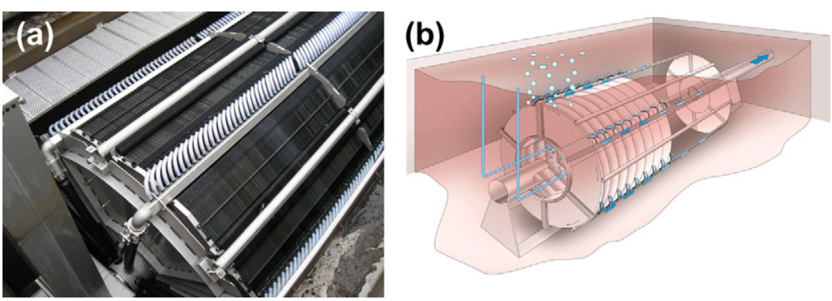

FIG. 4.12 The Huber *VRM*® module: (a) top view, showing permeate collection tubes and (b) schematic.

There were around 30 VRM plants installed by the end of 2009 with a mean capacity of 0.8 MLD, the largest plants having a capacity of up to 23,000 p.e. The company also has two conventional rectangular membrane panel products (the *ClearBox*® and the *BioMem* system) based on the same PES membrane material. The panels for these have dimensions of 800 × 400 × 4 mm with a total membrane area of 3.5 m^2 (*ClearBox*) and 25 m^2 (*BioMem*).

4.2.8. Jiangsu Lantian

Jiangsu Lantian Peier Membrane Co., Ltd (or 'Peier Membrane Industry') was established in Gaocheng Town, Yixing City in China in 2007. The company is a small-to-medium enterprise of less than 100 staff but nonetheless claims to have China's largest closed production facility for clean membrane fabrication with associated testing and laboratory facilities. The company provides 518 mm width rectangular panels of three different lengths, along with a fourth much smaller bench-scale panel, fitted with PVDF membranes of pore size in the range of 0.1−0.3 μm mounted on an ABS backing plate. The panels of 1190, 1780 and 2000 mm length provide a membrane area of 1, 1.5 and 1.75 m^2, respectively, with corresponding recommended SAD_m values of 0.72, 0.48 and 0.41 $Nm^3/(m^2\,h)$; the company thus provides the largest commercially available flat panel. The normal module size is 100 panels (thus between 100 and 175 m^2). A number of reference sites in China are identified by the company for this product.

4.2.9. LG Electronics

The LG *Green Membrane* was originally produced by KOReD, a Korean company founded in 1999 and originally focusing its activities on developing membrane technologies for water purification before more recently shifting its business activities towards water/wastewater engineering. The flat sheet *Neofil*® MBR membrane panel was launched in 2003. The company formed a joint venture with the Korean electronics giant LG Electronics in 2009, and LG are in the process of developing the MBR and reuse products further. The 1 m^2 *Green Membrane* panel is 1200 mm long by 490 mm wide and is fitted with a 0.01−0.2 μm PES membrane and a single permeate extraction nozzle. The panels are 4 mm thick − one of the slimmest of all commercially available rigid FS modules − and separated in the module by a channel width of 7 mm (Fig. 4.13). The module SAD_m value is 0.6−0.9 $Nm^3/(m^2\,h)$.

4.2.10. MICRODYN-NADIR

MICRODYN-NADIR GmbH is a German company specializing in membrane filtration. The company was formed from the merger between Microdyn

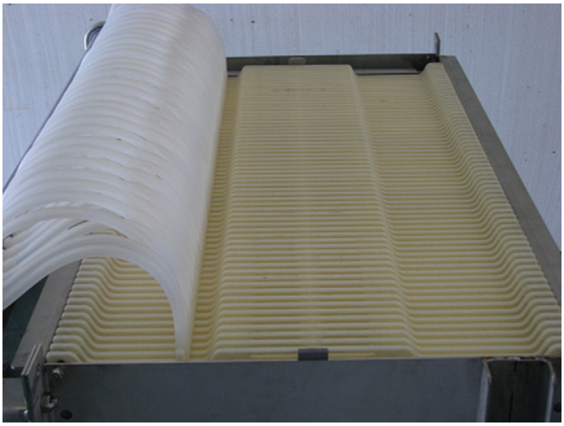

FIG. 4.13 LG Electronics *Green Membrane* module.

Modulbau and Nadir Filtration in 2003. The MICRODYN-NADIR *BIO-CEL*® MBR membrane technology was launched in late 2005.

The *BIO-CEL*® is a composite laminated material with the 0.04 μm PES membrane being permanently affixed to a macroporous fibrous separator (Fig. 4.14a), which has a very open structure (Fig. 4.14b). As such it combines the high-permeability properties of the conventional rigid flat sheet panels with the flexibility and strength associated with the braided hollow fibre products, as well as permitting high packing densities (222 m^2 membrane area per m^3 internal module volume) as a result of the reduced panel thickness of 2 mm. The membranes are backflushable up to pressures of 0.15 bar and can operate in filtration mode at pressures up to 0.4 bar. The membrane sheets for the standard *BC100* module are 1 m^2 in area, and are housed in four 'cassettes' of 25 sheets (Fig. 4.15a) each, which are spring mounted to the sides of the frame (Fig. 4.15b). The permeate is extracted via a 50 mm tube running through the centre of each sheet (Fig. 4.15c). Modules are available in sizes of 50, 100 and 400 m^2, the latter (the *BC400*) having dimensions of 1440 × 1152 × 2722 mm

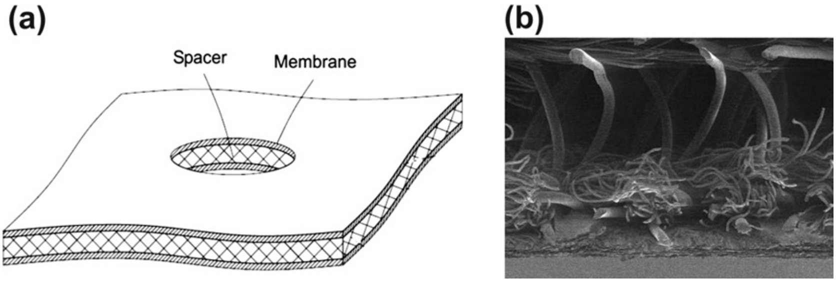

FIG. 4.14 The MICRODYN-NADIR *BIO-CEL*® laminated membrane: (a) schematic and (b) micrograph of cross-section.

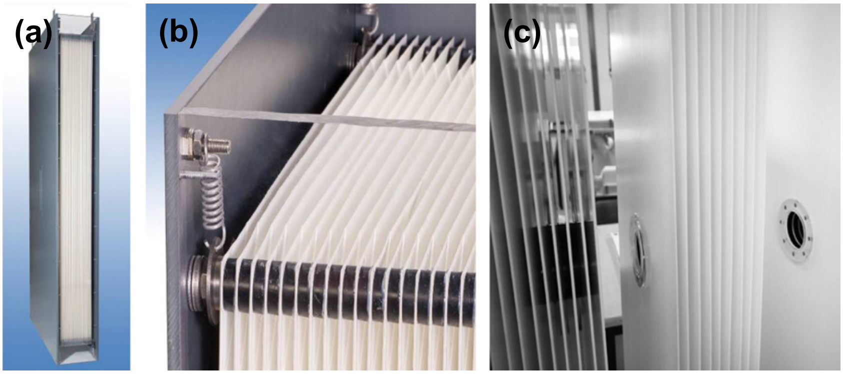

FIG. 4.15 The MICRODYN-NADIR *BIO-CEL*®: (a) cassette; (b) spring mounting; and (c) central ports for permeate manifold.

high. At 241 m^2 per m^2 footprint, this module provides the highest membrane area per unit area footprint of all the FS products with one of the lowest recommended SAD_m values: 0.21–0.4 $Nm^3/(m^2\,h)$ for the *BC400* compared with 0.3–0.6 for the *BC100*.

The product has been trialled at the University of Darmstadt as well as at other sites, and a 20-MLD MBR plant at Ji'an in China is to employ the technology.

4.2.11. Shanghai MegaVision

Shanghai MegaVision Membrane Engineering & Technology Co., Ltd is a small membrane manufacturing company which commercialized its FS membrane product in 2006. The company provides a 1-m^2 FS membrane product based on both PVDF and PES and with pore sizes of 0.1 and 0.3 μm. The panels, which are based on a PVC frame, are 930 mm high × 610 mm wide × 16 mm thick (including the panel separation) with a single permeate extraction port. The single-deck modules (Fig. 4.16a) are available as 100 and 150 panel units, and the recommended aeration value is 0.75 $Nm^3/(m^2\,h)$.

4.2.12. Shanghai Sinap

Shanghai Sinap was co-founded by the Shanghai Institute of Applied Physics and Shanghai Filter Co., Ltd. The company provides 0.1 μm pore size PVDF membrane panels of four different sizes, the two largest being 0.8 and 1.5 m^2 in membrane area. The dimensions of these two panels are 1000 mm × 480 mm × 7 mm and 1800 mm × 510 mm × 10 mm, spaced by 7 mm in the stainless steel frame module (Fig. 4.16b) which holds 150 panels. The technology, which has a minimum recommended SAD_m of 0.72 $Nm^3/(m^2\,h)$, has

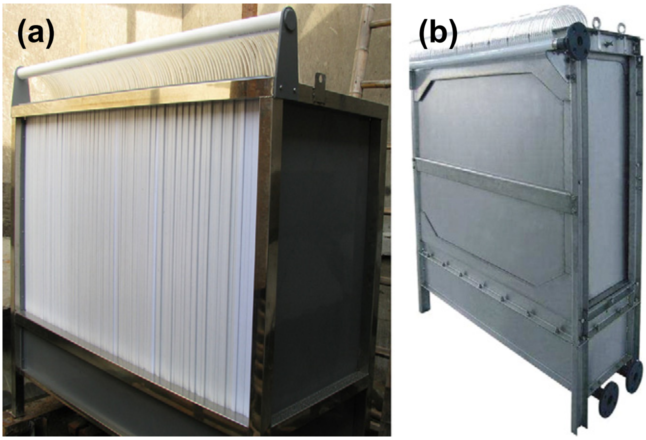

FIG. 4.16 The MBR modules of: (a) Shanghai MegaVision and (b) Shanghai Sinap.

apparently been applied to oil-bearing, laundry and pharmaceutical wastewaters, as well as landfill leachate and domestic sewage, with one single application employing 13,500 m^2 of membrane.

4.2.13. Toray

Toray is an established Japanese membrane manufacturer of some 30 years standing, specializing principally in reverse osmosis (RO) membranes for pure water applications. The company launched its FS MBR membrane product in 2004, now registered as *MEMBRAY*® in most regions of the world. The membrane material used is 0.08 μm PVDF, with a standard deviation of 0.03 μm. It is reinforced with a polyethylene terephthalate (PET) non-woven fibre and mounted on an ABS support, into which a number of 1−2 mm permeate channels are cut. Permeate is extracted via a single outlet tube. The panel (*TSP-50150*, Fig. 4.17a) has dimensions of 515 mm by 1608 mm, providing a membrane area of 1.4 m^2, and is 7.5 mm thick, with a panel separation of 6 mm. A smaller panel of 515 mm × 1059 mm also exists, its use being apparently limited to applications where height is constrained (such as on board ships).

Panels are assembled in a stainless steel frame to form modules ranging from 45 m^2 total membrane area (50 panels, *TMR090−050S* module, Fig. 4.17b) to 140 m^2 (100 panels, *TMR140−100S* module). The modules can then either be doubled in width (*TMR140−200W* module) or stacked (*TMR140−200D* module) to form larger modules. The *TMR140−100S* module has dimensions of 1620 mm long, 810 mm wide and 2100 mm high. A design flux of 33 LMH is assumed (though the quoted range is between 8.3 and 62.6 LMH for peak operation), along with a maximum TMP of 0.2 bar. The

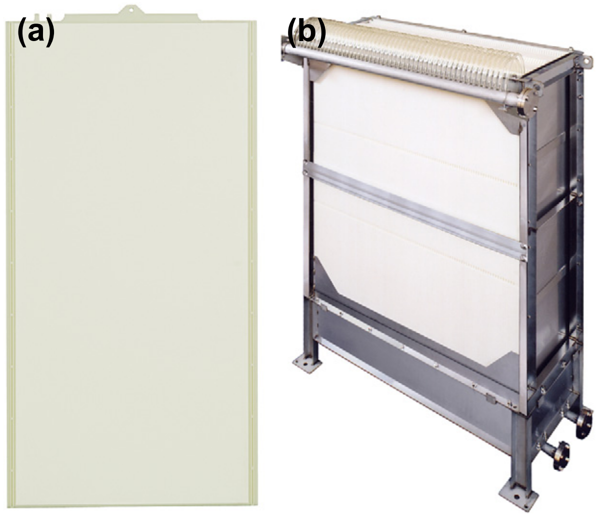

FIG. 4.17 The Toray: (a) element/panel and (b) module.

recommended aeration rate is 0.56 $Nm^3/(m^2\,h)$ for the single-deck unit, and pre-screening to 3 mm is stipulated. As of the middle of 2009 there were over 100 operating plants worldwide, with a mean size of just over 1 MLD, the largest undergoing commissioning in 2010 being the Yas Island plant in the UAE (Section 5.2.8.3).

4.2.14. Vina

The Suzhou Vina Filter Company provide both FS and HF membrane products for MBR duty, as well as membrane products for potable water and gas filtration. The *VINAP-150* FS product (Fig. 4.18) is based on a 0.05 μm pore

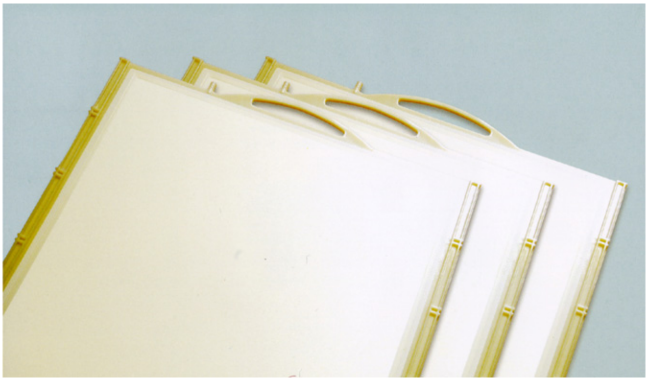

FIG. 4.18 The Vina *VINAP-150* FS panel.

size PVDF membrane with panels available in three different sizes, the two largest providing 1.0 and 1.5 m^2 of membrane area. The dimensions of these panels are 490 × 1000 mm, and 1780 × 510 mm, respectively, both 6 mm thick and spaced by 7 mm in the module. The minimum recommended SAD_m is 0.48 $Nm^3/(m^2\ h)$.

4.2.15. Weise

Weise Water Systems GmbH was founded in 2001. The company provides a series of submerged UF FS module products in their MicroClear® range, and had over 900 installations based on the technology at the end of 2008, compared with approximately 160 at the end of 2005.

The Weise system is one of the few which does not employ a single, top-mounted permeate collection nozzle. Instead, there is a manifold ('filtrate collector') fitted to the side of the membrane cassette (Fig. 4.19a). This collects permeate from the 24 (for the *MC03* model) or 21 (for the *MCXL*) membrane plates, which are based on 0.05 μm PES membranes. The plates are 2 mm thick and separated by 5.5 mm, such that the packing density provided by the system is amongst the highest of all the FS MBR products. The 492 × 165 mm *MC03* plate provides a membrane area of 0.146 m^2, or 3.5 m^2 from a 24-plate cassette. The company introduced the new *MCXL* filter in 2008, based on wider and slightly thicker plates (490 × 375 × 3.5 mm, 0.333 m^2 membrane area), and for which the 21-plate cassette offers a membrane area of 7 m^2. The largest module provided by the company, the *MA04-150* (Fig. 4.20), is based on a stack of 75 of these cassettes (5 × 3 × 3 deep), and has overall dimensions of 1950 × 1250 × 2100 mm high and demands an SAD_m of 0.76 $Nm^3/(m^2\ h)$ – the lowest for the

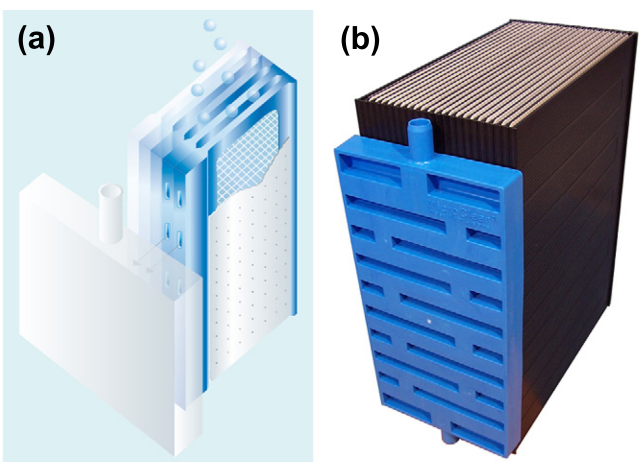

FIG. 4.19 The Weise *Microclear®* module permeate manifold: (a) schematic and (b) photograph.

FIG. 4.20 The Weise *Microclear*® *MA04-150* module.

product range. This 525 m^2 module has a footprint of 2.43 m^2. The membrane area per m^2 footprint offered by this module is thus 216 – one of the highest for the FS modules.

4.2.16. Other Flat Sheet MBR Membrane Products

A number of the FS products listed above have only recently been introduced to the market, and more are currently under development. Whilst it is not possible to list these comprehensively, a few are worthy of some attention.

The inge *FiSh* ('Fibre Sheet') PES membrane panel appears to have been successfully demonstrated through pilot plant trials conducted at Anjou Recherche (Grélot et al., 2009). The panel is a self-supported 3-mm-thick sheet of 0.2 μm PES membranes (Fig. 4.21). The membranes are backflushable and the modules potentially simple to fabricate. Another thin flat sheet membrane panel is being developed by VITO in collaboration with Agfa, the photographic film company. For this technology two membrane layers are coated to the faces of a spacer-fabric, which forms an 'integrated permeate channel' (*IPC*) membrane envelope. The spacer-fabric which is used as a membrane support, comprises a monofilament PET layer sandwiched between two 0.3-mm woven layers, onto which the membrane layers are coated (Fig. 4.22). A 0.3-μm-rated membrane sheet formed in this way has been shown to be readily backwashable at TMPs as high as 1 bar. This facilitates the stability of MBR operation (stable permeability) and allows higher flux levels (Doyen,

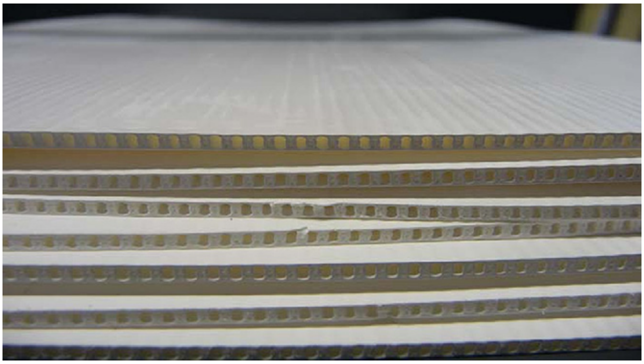

FIG. 4.21 The inge *FiSh* membrane.

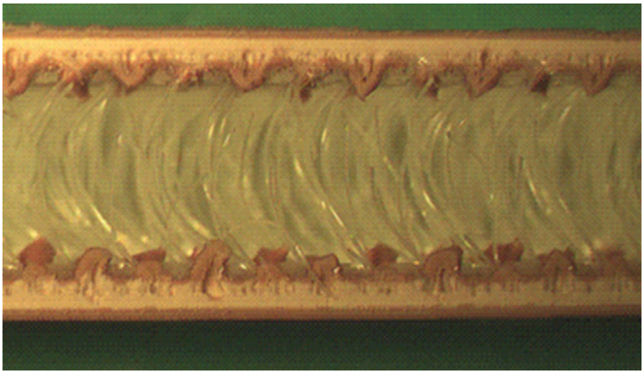

FIG. 4.22 The VITO-Agfa *IPC* membrane envelope.

Mues, Molenberghs, & Cobben, 2010). Both of these developments would be expected to lead to further backflushable FS MBR technologies.

A further more unusual development is that reported by the Institute of Water Quality and Waste Management at the University of Hanover, in collaboration with ITN Nanovation GmbH. These workers have tested a ceramic (α-alumina) FS panel, 0.08−0.3 μm in pore size, with internal permeate channels of 3 mm and a panel thickness of 6.5 mm. The module has been tested against municipal wastewater and paper and brewery industrial effluents, operating at permeabilities between 30 and 300 LMH/bar, with the lowest permeabilities recorded for the municipal wastewater.

Finally, the Singaporean company Hyflux have secured the contract to build a 68-MLD MBR water reclamation plant at Jurong in Singapore, and have developed their own PVDF-based FS MBR presumably with a view to employing it at this site.

4.2.17. Other Flat Sheet MBR Technologies

Whilst the membrane product is an important component of an MBR, there are other aspects of the process that contribute to the technology. A number of established companies offer package plant technologies based on available flat sheet membrane products. These include Hitachi and Busse.

Hitachi Plant Technologies Ltd have developed a technology based on 0.1 μm PVDF FS membrane panels having dimensions 0.5 m wide by 1.0/1.5 m tall. Treatment plants having capacities between 30 and 750 m^3/day have been installed, with the smaller plants (between 30 m^3/day and 100 m^3/day) being containerized, and the company has the capability to provide larger MBR plants. The modules can be stacked two (Fig. 4.23) or three deep. Hitachi have had industrial effluent installations in Japan dating back to 1997 and more recent sites in the UAE. The latter have all been based on their three standard package plant sizes of 0.25, 0.5 and 0.75 MLD, the smallest being 13.6 m long by 4.13 m high and the largest 17.8 m long by 5.44 m high. The quoted energy demand is 0.72–0.81 kWh/m^3 across the three plant sizes. Hitachi have supplied many MBR compact units in Middle East and Asia regions, as well as targeting other markets such as Australia, Africa, Europe and the Americas region.

Busse IS GmbH provide package plant MBRs down to single household size, the smallest (the *MF-HKA4*, Fig. 4.24) being for 4 p.e. and 0.6 m^3/day flow. Due to its modular design the system can be expanded to up to 500 p.e.

FIG. 4.23 The Hitachi membrane unit *(all rights reserved, Copyright © 2010 Hitachi Plant Technologies, Ltd).*

FIG. 4.24 The Busse package plant.

The patented system is based on Kubota membrane panels, and was originally introduced in 1997. As such, Busse is the most established provider of domestic-scale wastewater treatment technology, and had already installed around 270 such plants by the end of 2005 (Lesjean & Huisjes, 2008) and more than 500 by the end of 2009 – in Germany and in 12 other countries throughout the world. The two-tank system comprises an aerated pre-cleaning stage, with a coarse screen fitted to the outlet, followed by the bioreactor fitted with the submerged membrane. As with most FS MBRs the filtration is driven by the hydrostatic head. The small size coupled with the requirement for highly conservative operating conditions means that the specific energy demand is 2.5–4 kWh/m^3, though this only amounts to a maximum power of 200 W at the domestic wastewater flows concerned. The system is produced, distributed and serviced by Busse Innovative Systeme GmbH in Germany and by co-operating partners in 12 other countries.

Busse generally have service contracts for their package plants, involving on-site servicing at fixed intervals of 6–12 months. The membrane units are removed and disassembled, and the panels replaced with reconditioned ones. The used panels are then carefully thoroughly cleaned off-site for reuse. By maintaining a pool of reconditioned membranes the company is always able to provide panels when required, and is able to limit the costs incurred and waste generated. The relatively benign operating conditions of the plant mean that the membrane life can exceed 10 years.

There are additionally a number of other MBR products for which comprehensive product information could not be obtained by the time this book went to press. These include the products of Pure Envitech ENVIS from Korea, Kang Na Hsiung (KNH) from China, and Martin Systems AG from Germany. Pure Envitech supply the *ENVIS* product, which is based on a 0.4-μm pore size 8-mm-thick membrane panel of 0.98 m^2 area, and has two permeate extraction points at the top of the panels. Martin Systems AG provide the *siClaro*® range of products, which comprises both a rotating membrane (the *DM*) and a conventional rectangular panel (the *FM*), with up to triple-deck units provided for the latter. It is unclear as to whether the KNH product is an MBR or a fixed film reactor, the fixed film apparently being based on a non-woven felt.

There continue to be further MBR membrane product developments and joint ventures around the world, most notably in the Far East, and some are indicated in Appendix C. In April 2010, for example, the Japanese chemical and pharmaceutical company Teijin Ltd formed a joint venture with Membrane-Tec Co., Ltd to co-develop wastewater treatment systems for rural communities in China. The agreement was signed with Yixing City Water Works & Construction Investment Co., Ltd of the Jiangsu Province of China. Teijin have a proprietary multi-stage biological wastewater treatment technology which is to be combined with membrane filtration provided by Membrane-Tec's PTFE MF FS product.

4.3. IMMERSED HF PRODUCTS

4.3.1. GE–Zenon

GE has over 300,000 employees (323,000 in 2008), a turnover of $182b, assets of almost $800b and a market value of $90b, according to Forbes. As such it ranks as one of the largest companies in the world, and the 12th largest in 2009 based on revenue. It was one of the 12 original companies on the Dow Jones industrial average in 1896, and continues to be active in electrical power generation technology, lighting and home electrical. The conglomerate has extended its activities, either through growth or acquisitions, to such areas as jet engine technology, computer technology, finance, entertainment and health technologies. GE Energy's renewable energy business has expanded greatly, to keep up with growing US and global demand for clean energy. Under its 'Ecomagination' programme, GE has invested more than $850m in renewable energy technology since entering the renewable energy industry in 2002. As part of the same initiative the company is committed to reducing its own water by 20% by 2012, and exporting water-saving and recycling technology to those emerging economies affected by shortages. The group plans to employ water recycling technologies at more than 1000 plants around the world – mainly in the USA, Europe and Asia.

In June 2006 GE acquired Zenon Environmental Inc., a company that pioneered the development and commercialization of membrane technologies for water and wastewater treatment, forming the Water and Process Technologies group (GE W&PT). Zenon's earliest product, the tubular membrane

Permaflow (introduced in 1983), was successfully applied in small industrial MBR applications from the mid-1980s onwards, including in the development of the *ZENOGEM* MBR process in partnership with General Motors. In 1989, it developed a hollow fibre concept product called *Moustic*, which, although never commercialized, paved the way for the development of Zenon's most successful technology, the *ZeeWeed®* immersed hollow fibre membrane, which was introduced in 1993.

ZeeWeed® is now a core membrane technology in the portfolio of products and solutions offered by GE W&PT. The Zenon acquisition has made GE W&PT one of the world's largest manufacturers of UF HF membranes, capable of producing enough *ZeeWeed®* membranes on an annual basis to treat over 4.75 billion litres (1.25 billion gallons) of water per day. A leading global supplier of water and wastewater treatment solutions, GE is the largest MBR process technology company in the world, with an operating installation base of over 650 plants in 47 countries representing a combined average daily wastewater treatment capacity of 3000 MLD – about double that of its nearest competitor – and 6800 MLD-years of experience.

The *ZeeWeed®* membrane used for MBR applications consists of a woven reinforcing braid on which a UF PVDF membrane is cast. The inner braid provides the tensile strength required to operate over the long term in high solids applications, eliminating the risk of fibre breakage. Thousands of vertical fibres form modules fitted with polymeric top and bottom collection headers into which the fibre open ends are potted with polyurethane. The double header construction maximizes module performance by avoiding excessive pressure drop as the water travels along the length of the fibre lumen. Membrane modules are in turn assembled into cassettes, designed to minimize system connections for ease of installation and maintenance, which are connected via permeate and aeration headers to form trains (Fig. 4.25). Permeate is extracted via the permeate header with a permeate pump or gravity, depending on the particular site hydraulics. During normal operation, the membrane filtration system is operated with a repeated filtration cycle, which consists of a production period (permeation) followed by a short relaxation or backpulse period, where backpulsing comprises short bursts of permeate flowing in the reverse direction for more effective physical cleaning of the membrane.

The first generation of commercially available *ZeeWeed®* membrane products was the *ZeeWeed® 145*, developed in 1993 (Fig. 4.26) and so-named because it offered 145 ft^2 (13.5 m^2) of membrane surface area. The next membrane developed was the *ZeeWeed® 150* (150 ft^2 or 13.9 m^2), which was released in 1995 and was the first self-supporting immersed vertical hollow fibre membrane design based on combining modules in a system to increase plant capacity. There followed the *ZeeWeed® 500A* (500 ft^2 or 46.5 m^2) in 1997 and the *ZeeWeed® 500C* (220 ft^2 or 20.4 m^2 initially and later 250 ft^2 or 23.4 m^2) in 2001, both of which optimized fibre spacing and were housed in standardized cassette frames. The latest version of the *ZeeWeed® 500* product

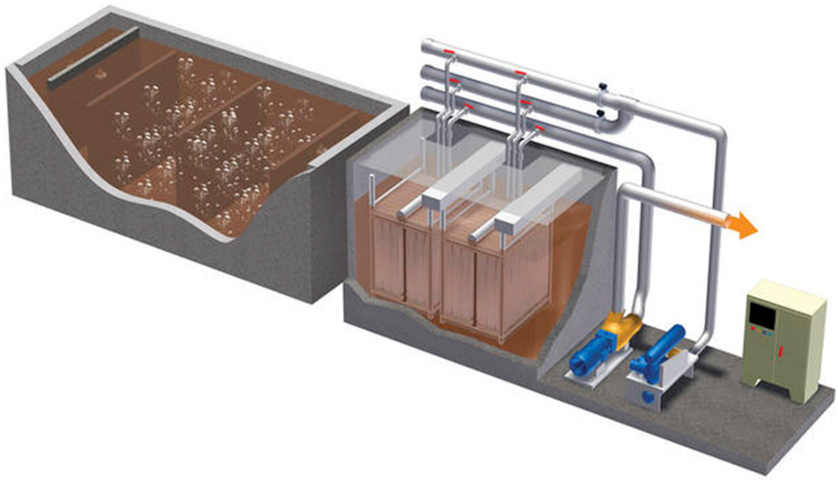

FIG. 4.25 *ZeeWeed®* membrane train.

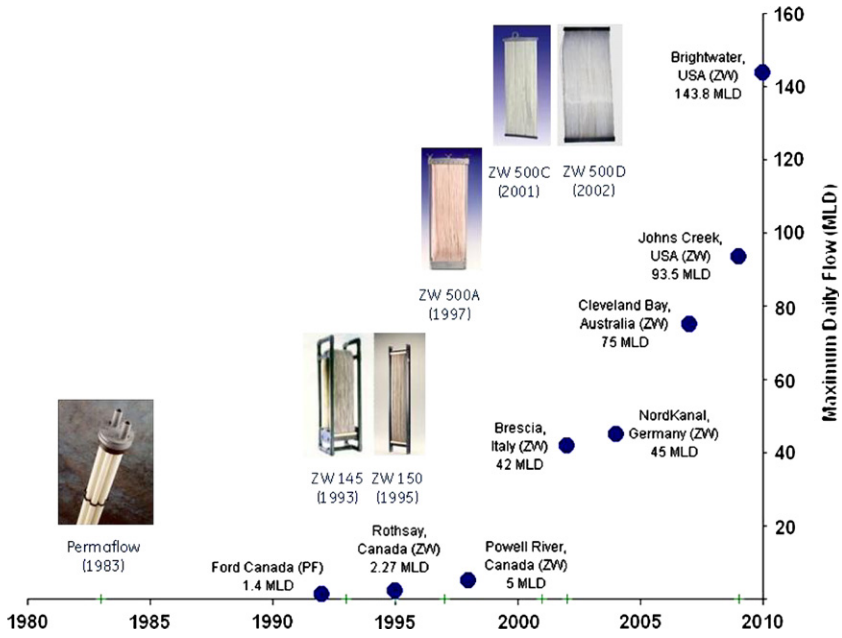

FIG. 4.26 The *ZeeWeed®* product timeline with major project milestones.

for MBR applications is the *ZeeWeed® 500D* family, which includes four available module surface areas and three cassette frame sizes (Fig. 4.27 and Table 4.2). The *ZeeWeed® 500D-S* model is identical to the *500D* but is shorter to match the height of the *500A* and *500C* generation of products. The

FIG. 4.27 The *ZeeWeed® 500D* MBR product family.

ZeeWeed® 500D-S module in combination with the 16 module cassette frame can be used to retrofit or replace these earlier generations of products while providing additional benefits associated with the *500D* product, including reduced aeration rates. Integral to the membrane cassette frame are coarse bubble aerators that deliver the air used to scour the membrane surface. The *ZeeWeed® 500D* modules are arranged in two parallel rows within the cassette with a central permeate header. The *500D* product design maximizes filtration capacity, reduces plant footprint and minimizes aeration energy.

For the *ZeeWeed® 500D*, membrane aeration employs low-pressure air from the aeration header to the specially designed coarse bubble aeration system at the base of each cassette, scouring the membrane surface and transporting mixed liquor out of the fibre bundle. Membrane aeration also provides a portion of the biological process oxygen requirements, with the remainder provided by more efficient fine bubble diffused aerators in the biological reactor. GE's patented method for controlling membrane aeration minimizes aeration energy demand by aerating for 10 s in a 40-s period: '10/30 aeration', first introduced in 2005 (Fig. 4.28). During peak flow events aeration strategy changes to 10/10 sequential aeration, where aeration is for 10 s in a 20-s period. This is achieved by manipulating air distribution valves and controlling either the number of blowers in service or the blower speed. The use of variable aeration cycling according to the imposed flux has been shown to reduce significantly the energy consumption without impairing process hydraulic performance.

TABLE 4.2 *ZeeWeed®500D* **MBR Series Module Specifications**

Parameter	*ZeeWeed® 500D*		*ZeeWeed® 500D-S*	
Year of introduction	2002	2010	2009	2010
Membrane area per module, m^2	31.6	34.4	25.2	27.9
Nominal pore size, μm	0.04	0.04	0.04	0.04
Membrane material	PVDF	PVDF	PVDF	PVDF
Reinforced fibre	Y	Y	Y	Y
Flow direction	Outside–in	Outside–in	Outside–in	Outside–in
Number of cassette sizes	2	2	1	1
Maximum number of modules per cassette	16 or 48	16 or 48	16	16
Packing density (m^2/m^3)	228	253	221	242

The maximum capacity of a *ZeeWeed®* MBR plant has increased dramatically since the release of the first *ZeeWeed®* modules and today the *ZeeWeed® 500* membrane is employed in many of the world's largest MBR plants. These include the largest operating MBRs (in terms of maximum daily

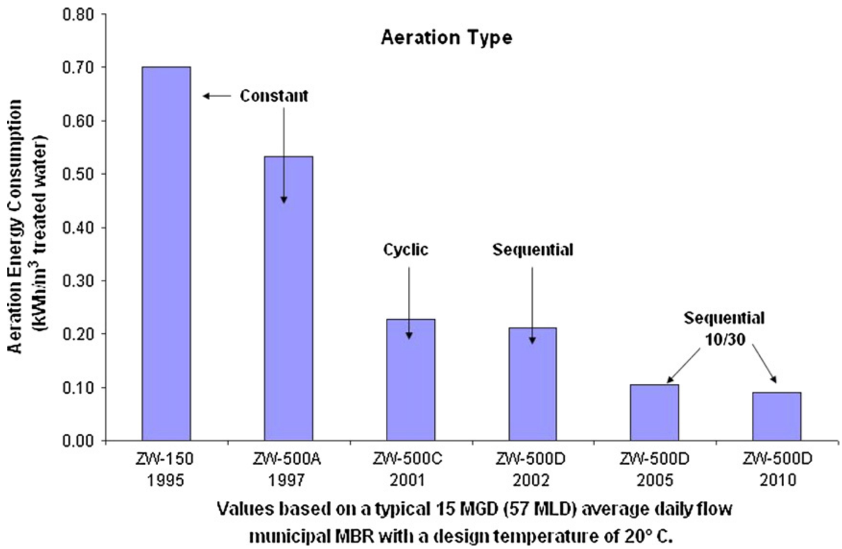

FIG. 4.28 *ZeeWeed®* MBR membrane aeration energy by-product.

FIG. 4.29 *ZeeWeed*® MBR *Z-Mod*™ packaged plant.

flow, or MDF) such as Johns Creek Environmental Campus, Georgia, USA (41 MLD average daily flow, or ADF, 94 MLD MDF); Cleveland Bay, Australia (29 MLD ADF, 75 MLD MDF); and Loudoun County, Virginia, USA (38 MLD ADF, 71 MLD MDF), as well as the largest MBR projects currently under design or construction such as Brightwater, Washington, USA (117 MLD ADF, 144 MLD MDF); Jumeirah Golf Estates, Dubai, UAE (113 MLD ADF, 135 MLD MDF); and City of North Las Vegas, Nevada, USA (95 MLD ADF, 132 MLD MDF). In addition to custom-designed MBR installations such as these, GE offer a range of packaged MBR systems with treatment capacities in the range of 0.015−15 MLD (Fig. 4.29). GE offer the *Z-Mod*™ line of pre-designed, engineered and skid-mounted MBR equipment, which contribute to lower overall project costs and shorter delivery timelines. These package plants are designed around the *ZeeWeed*® *500* membrane products. To date, over 100 package system MBR facilities have been installed worldwide for a diverse range of commercial, municipal and industrial customers.

4.3.2. Asahi Kasei

Asahi Kasei Chemicals Corporation is part of the Asahi Kasei group in Japan. The company has been manufacturing HF UF/MF membranes and modules for various industrial applications since the 1970s. The *Microza*® PVDF HF MF membrane for water treatment applications was introduced in 1998, and is in widespread use for pure water treatment as a pressurized module in their industrial and municipal water treatment system. Key features of the 1.3 mm diameter *Microza*® membrane fibre are the narrow pore size distribution (rated 0.1 μm determined by rejection of uniform latex), and the high tolerance to oxidative chemicals (Fig. 4.30). Tests conducted against both 0.5 wt% sodium hypochlorite and 4 wt% sodium hydroxide have shown no significant decrease in tensile strength over a 60-day period. This has been attributed to the high crystallinity of the *Microza*® PVDF material.

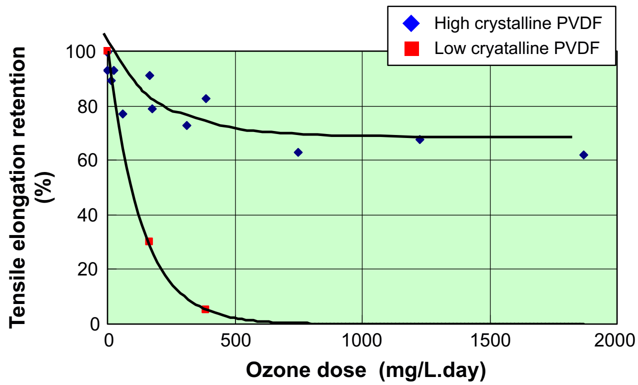

FIG. 4.30 Tensile strength against ozone loading, *Microza*® fibre.

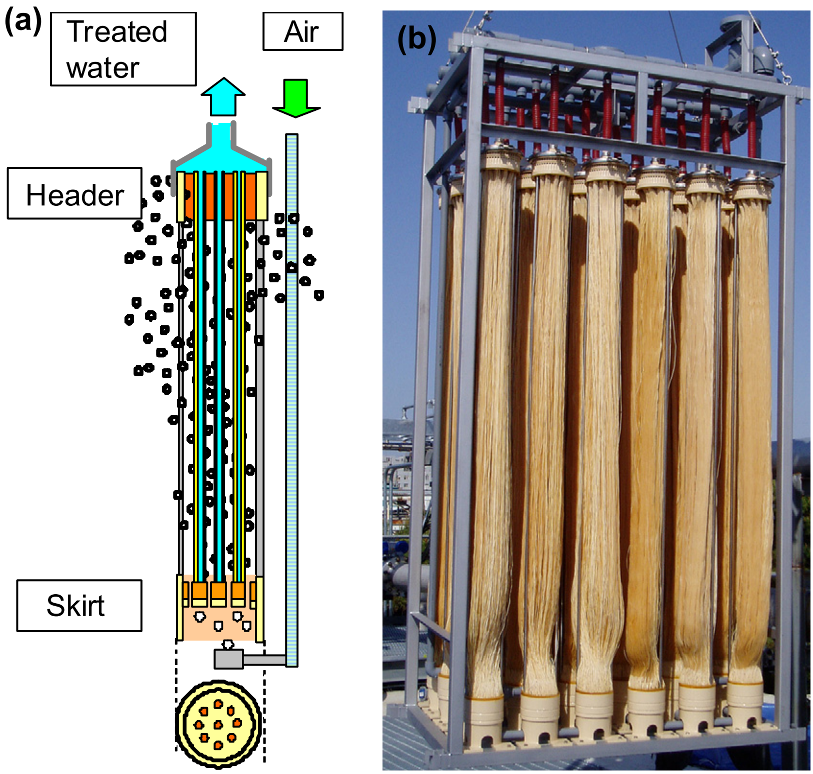

FIG. 4.31 Asahi Kasei: (a) schematic of module and (b) 24-element rack.

The MBR membrane module (*MUNC-620A*) has been under development since 1999 and was formally launched in 2004. The rack comprises a number of 167 mm diameter × 2163 mm high vertical modules mounted onto a steel frame (Fig. 4.31a), each module offering a membrane area of 25 m^2. The fibres are potted at both ends and gathered into bundles at the top of the module to encourage the escape of suspended matter. Air is provided by a separate aerator and is directed into the module via a series of ~10 mm holes in the module base around which fibre bundles are potted. Air bubbles are thus introduced at the centre of each fibre bundle thereby ensuring good air—fibre contact at the base (Fig. 4.31a). The standard 24-module rack, such as that shown in Fig. 4.32b, provides 600 m^2 membrane area in a 1400 × 920 × 2900 mm high frame. Whilst the packing density within each element is over 530 m^2 membrane area per m^3, for the overall rack it is around 161 m^2/m^3 indicating the extent to which the fibre bundle expands in the air stream.

Original pilot trials on the Asahi Kasei MBR technology were conducted on the module in collaboration with the Japan Sewage Works Agency at Mooka, based on a flow of 0.036 MLD. In these trials a net flux of 30 LMH was sustained yielding an SAD_p of 13.3 m^3 air/m^3 permeate, employing the recommended SAD_m of 0.2—0.28 $Nm^3/(m^2\,h)$. The first installation, for food effluent treatment, was subsequently commissioned in 2005. By the middle of 2009 there were eight municipal plants based on the technology providing a total installed capacity of 180 MLD, along with industrial plants with a total installed capacity of 91 MLD. One hundred megalitres a day of the municipal wastewater treatment capacity is provided by the Wen Ye He plant in Beijing (Section 5.3.2.1). This plant, currently the world's second largest in terms of peak flow capacity, was installed in 2007 as part of the preparation for the 2008 Athletic Boat Games.

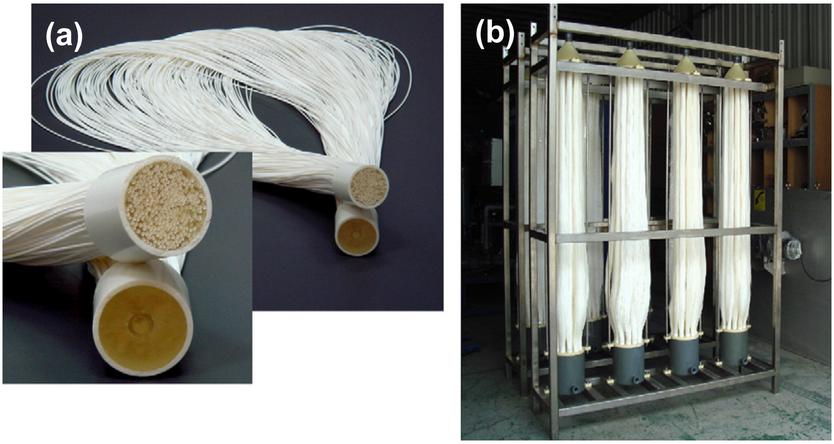

FIG. 4.32 *Ecofil*™ and *Ecoflon*™ products: (a) fibre bundle and (b) modules.

4.3.3. Beijing Origin Water

Beijing Origin Water Technology Company (Origin Water) conducts business in environment protection and resource recycling, with its core business relating to the application of membrane filtration to municipal and industrial sewage treatment and recycling. The company was established in 2001 and, following a major expansion in 2003, achieved an annual income in 2005 of more than 120m Yuan ($15m) which had increased sixfold by 2008. Origin Water is a well-known and established brand in China.

The company provides a PVDF HF MBR technology with three different pore sizes (0.3, 0.1 and 0.02 μm) and two different filament diameters (2.0 and 2.4 mm, the *BSY* and the *RF*, respectively). The modules for all membrane types are rectangular, 2000 mm high × 1250 mm wide × 30 mm thick, with a membrane area of 26.5 m^2 for the *BSY* and 27.5 m^2 for the larger diameter *RF*. The stack on which the module is based contains 60 modules, providing 1602 m^2 for the *BSY* and 1650 m^2 for the *RF*, with dimensions of 3076 mm high × 3334 long × 1760 mm wide in each case. The company developed the technology for the Wen Yu River plant in Beijing, based on the Asahi Kasei membrane product (Section 5.3.3.1), and the Shen Ding River plant in Hubei (Section 5.3.3.1), which at 110 MLD capacity and employing 108 modules was the world's largest MBR plant as of April 2010 and is based on the *BSY* membrane.

4.3.4. Ecologix: *EcoFil* and *EcoFlon*

In addition to the two flat sheet products (Section 4.2.6), Ecologix also produces HF modules in two different membrane materials: PVDF and PTFE. Both products feature 1.3 mm diameter filaments in 160 mm diameter modules supplied in lengths of 1000, 1500 and 2000 mm, providing a membrane area of 14.4, 21.6 and 28.8 m^2, respectively, with the active fibre length being 500 mm less than the total element length. The fibres are gathered into cylindrical bundles of around 40 mm diameter (Fig. 4.32a) at the base and the top of the module, which then forms a rack of eight modules (Fig. 4.32b).

4.3.5. ENE

'Energy and Environment' (ENE) are a Korean company whose MBR product, the *SuperMAK*, is based on a 0.4-μm pore size 2-mm diameter PVDF fibre. Lengths of the fibre appear to be wrapped around the base of the PVC frame of the module (Fig. 4.33), which is free standing with no other supporting framework. The module dimensions are 400 × 160 × 720 mm high, with the single permeate header at the top of the module, providing an area of 10 m^2, with 14 such modules inserted in a skid of 1400 × 1000 × 800 mm high. Stacking of this module appears to be possible, since a 1600-mm high unit of

FIG. 4.33 The ENE *SuperMAK* module.

the same footprint is also available based on 28 modules. The recommended SAD_m value is 0.6 $Nm^3/(m^2\,h)$.

4.3.6. Hangzhou H-Filtration

As with a number of MBR membrane suppliers in China, The Hangzhou H-Filtration Membrane Technology & Engineering Co., Ltd is in effect a university spin-out company, in this case from Zhejiang University. The company's activities have been focused on the development and production of HF membrane separation technology and engineering for more than a decade, with their product range encompassing water filtration and gas separation applications in the pharmaceuticals, chemical, food and beverage and automotive industries. A number of small industrial effluent plants, ranging in capacity from 0.01 to 0.2 MLD, have been installed based on the technology.

Their MBR *MR* product is a PP HF membrane module (Fig. 4.34a) with filaments of 0.45 mm diameter − one of the lowest of all commercially available MBR membranes − with a 70−80 μm wall thickness. It appears that the modules can be arranged either horizontally or vertically in the stack. The polymer is assumed to be dry-spun (Section 2.1.2), and hence with slit-like pores. The module, which has dimensions of 810 mm wide × 525 mm

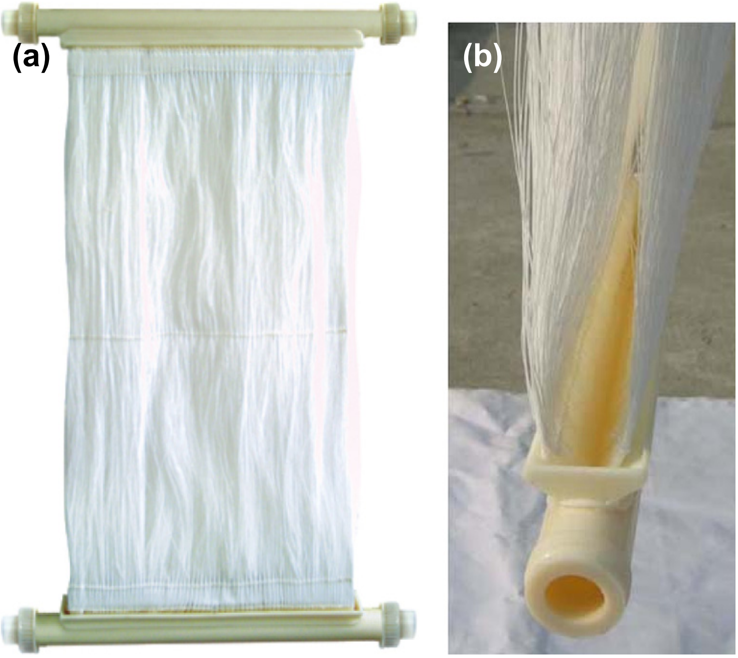

FIG. 4.34 The Hangzhou MBR: (a) module and (b) header.

high $\times$ 55 mm thick, provides a membrane area of 8 m^2 from two layers adjoined to the headers (Fig. 4.34b), and the SAD_m value is relatively low at 0.1–0.16 $Nm^3/(m^2\,h)$ – as is the recommended flux. The module can be stacked to provide a double- or triple-deck unit, the latter giving a membrane area of 24 m^2 from a unit around 2 m high, and can be operated up to 0.3 bar.

4.3.7. Koch Membrane Systems

Koch Membrane Systems is an established pure water membrane company of some 35 years standing, with over 15,000 global installations in both the municipal and industrial sectors such as food and beverage, biopharma, paint and pigments, automotive, power generation, oil and gas, pulp and paper and microelectronics. The company provides products and support for all pressure-driven membrane processes, from reverse osmosis through to microfiltration, and owns the membrane brands *ABCOR*®, *FLUID SYSTEMS*® and *ROMICON*®, as well as the MBR membrane module brand *PURON*®.

The company Puron was originally formed in late 2001 as a spin-out company from the University of Aachen, and was subsequently acquired by Koch Membrane Systems in 2004. The membrane is based on PES, and is

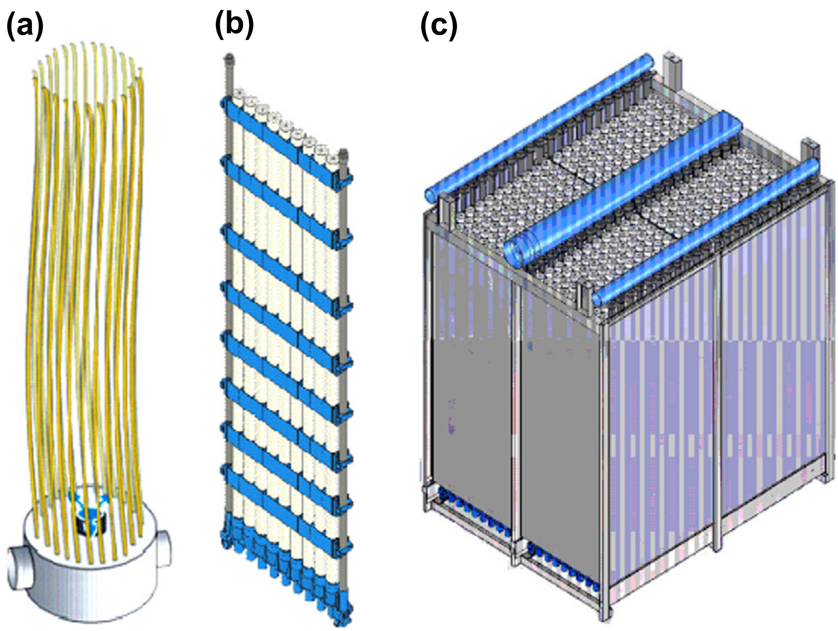

FIG. 4.35 The *PURON*® submerged membrane module: (a) fibre bundle, (b) module row and (c) module.

2.6 and 1.2 mm in external and internal diameters, respectively. An unusual feature of the *PURON*® submerged membrane module is the securing of the fibres only at the base, with the membrane filaments individually sealed at the top (Fig. 4.35a). Scouring air is injected between the filaments intermittently by means of a central air nozzle at the module base so as to limit the degree of clogging in that region. The free movement of filaments at the top of the module is designed to allow gross solids, such as hair and agglomerated cellulose fibres, to escape without causing clogging in this region. The fibres are strengthened by an inner braid, since the lateral movement of the filaments subjects them to a certain degree of mechanical stress. In the normal operational mode, aeration is applied for between 25% and 50% of the operational time at a rate of 0.133−0.3 $Nm^3/(m^2\,h)$ depending on the application and module size.

Filtrate is withdrawn from the manifold at the base of the cylindrical element which also houses the aerator (Fig. 4.35a). The individual fibre bundles are 1830−1990 mm high and provide a membrane area of between 3.5 and 3.8 m^2, the area and height both depending on the module size. These bundles are connected in rows (Fig. 4.35b), with several of these rows mounted into a stainless steel frame to form a membrane module (Fig. 4.35c). The *PURON*® submerged membrane module is available in standard sizes of 250, 500 and

FIG. 4.36 The *PURON®* *PSH 1500* submerged membrane module.

1500 m^2 membrane area, the latter (Figs 4.35c and 4.36) having dimensions of 2244 × 1755 × 2530 mm high (and hence a footprint of 10 m^2).

4.3.8. Korea Membrane Separation

KMS has been a major player in the Korean MBR market since 2000, with over 1000 installations in total (albeit most of them less than 0.5 MLD in capacity). Since 2003 their product has been based on one of the more unusual membrane materials (high-density polyethylene), made by stretching to produce elliptical pores (Fig. 2.4a) to produce an asymmetric structure with a very porous inner surface and denser outer surface. This replaced the original PP membrane. The *KSMBR* process, developed by cooperation with SsangYong Engineering and Construction (SsangYong E&C) and Korea Water Resources Corporation (K-Water), was introduced in 2005. By 2007 there were over 600 sites, mainly small domestic units and predominantly in Korea, based on the KMS and *KSMBR* technologies, and by 2009 this had risen to over 1000 sites for all KMS-based systems, including the *KSMBR* (Kowaco Ssangyong Membrane Bioreactor).

The company provides relatively small flat membrane elements ('sub-units', Fig. 4.37a) of hydrophilicized HDPE HF fibres which are 0.65 mm in

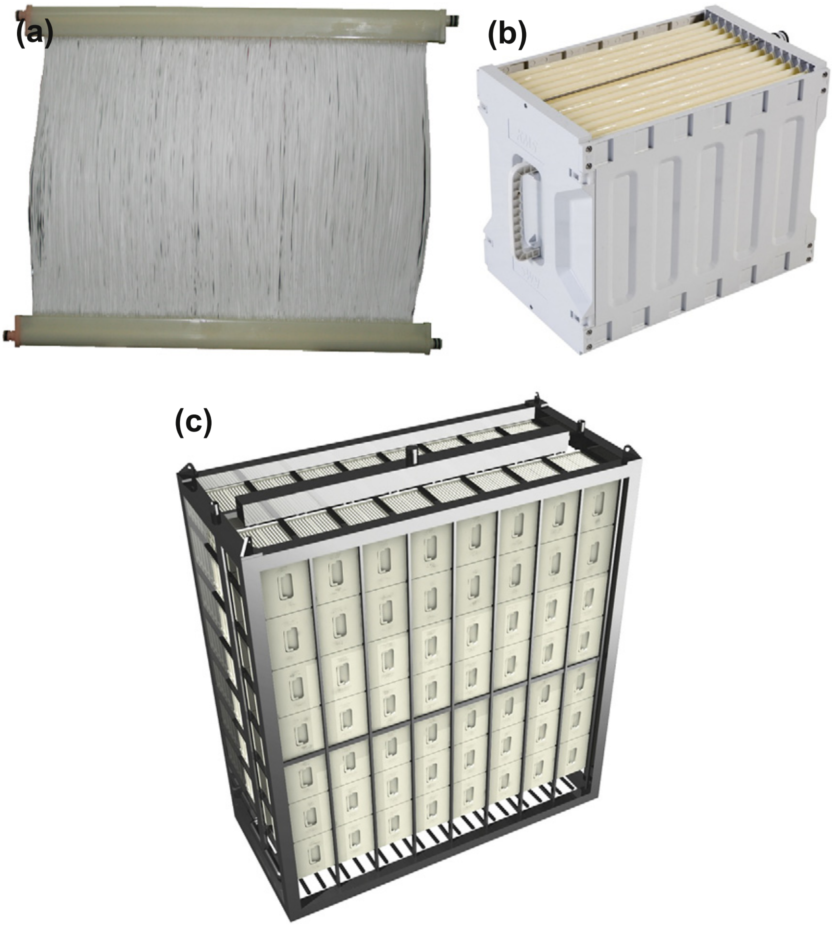

FIG. 4.37 The KMS membrane: (a) element (sub-unit), (b) module (cartridge) and (c) stacks.

outer diameter and are rated 0.4 μm. The elements sit inside modules ('unit cartridges', Fig. 4.37b) which can be stacked up to eight deep within a metal frame fitted with an integral aerator (Fig. 4.37c). The elements are 14 mm thick and provide 1.385 m^2 membrane area. Thirteen of these elements sit inside the modules, which are 536 × 320 × 396 mm high, and offer a membrane area of 18 m^2. The geometry and construction of the cartridges, i.e. the symmetrical nature, the flush fitting of the base and top and the overall weight of only 7 kg when dry, mean that they can be manually inverted to ameliorate problems of localized sludging and extend the operational period between recovery cleans.

The respective packing densities with reference to membrane area in m^2 per unit internal volume in m^3 of the element, module and stack are around 605, 265 and 159 respectively, and the largest stack offered by the company, the *6007CF*, provides a membrane area of 2016 m^2 from 112 cartridges within a frame of 2956 × 1326 × 3240 mm high. The relatively low cost of the membrane material means that the module can be economically operated at low flux, without backflushing, and low specific aeration demand. At the supplier's recommended flux of 12.4 LMH for the *6007CF* and 0.15 Nm^3/h aeration per m^2 membrane area the SAD_p value is around 12.

The *KSMBR* process is essentially the KMS MBR process modified to permit biological nutrient removal, and thereby achieves New Environmental Technology certification in Korea. The process had been installed in 139 sites by the end of 2009, predominantly within Korea. Installations include the 25 MLD Dalsung site (Section 5.3), which receives both municipal and industrial effluents and a large 73 MLD municipal plant for expansion of an existing works.

4.3.9. Litree

The Hainan Litree Purifying Technology Company Ltd was founded in 1992 as Litree Enterprises, and is an established manufacturer of HF UF membranes and modules in Asia for industrial and municipal water supply.

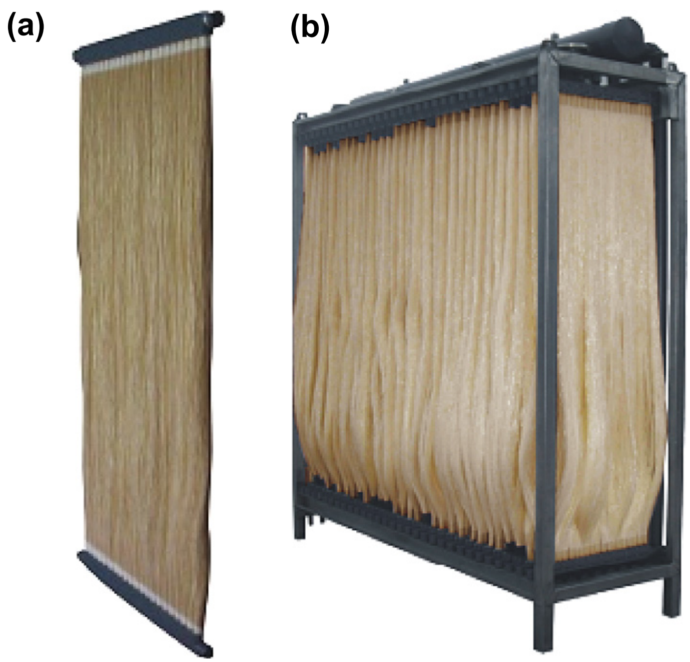

FIG. 4.38 The Litree: (a) module and (b) stack.

The company has two manufacturing facilities in the Hainan and Jiangsu provinces of China capable of supplying 3 million m^2 of membrane per year. The MBR HFLH3 membrane module is based on PVDF material with a pore size of 0.02 μm. The fibres are 1.8 mm in outside diameter (1.0 mm internal diameter), and the rectangular modules (Fig. 4.38a) are 721 × 70 mm thick (the latter assumed to include the module separation in the stack, Fig. 4.38b) with a height of 1222−2122 mm, providing an effective fibre length of 1100−2000 mm and a membrane area of 13, 18 or 25 m^2. These modules appear to comprise 20 fibre bundles around 25−30 mm in diameter. The stack has dimensions of 2138 × 855 × 2210−2710 mm height providing a membrane area of 468 or 650 m^2 from 26 of the 18 or 25 m^2 modules, respectively.

4.3.10. Memstar

Memstar Technology Ltd is a Singaporean company listed on the main board of the Singapore Exchange. The company specializes in the development, manufacture and application of PVDF HF membranes and membrane products. The company's headquarters and R&D centre are located in Singapore, with manufacturing plants based in Guangzhou City and Mianyang City in China. The company's membrane and membrane products are used mainly for water and wastewater treatment systems such as MBRs and continuous membrane filtration (CMF) technologies for various applications. The company had more than 16 reference sites as of 2009 providing a total installed capacity of 148 MLD. Memstar has also won a contract to provide the membrane product for a 100-MLD MBR plant in Guangzhou, one of China's largest MBR plants in terms of treatment capacity, which was due to be completed by the middle of 2010.

The Memstar MBR technology is based on a PVDF HF membrane of <0.1 μm pore size and 1.2 mm in outside diameter set in 16 cylindrical bundles across the width of the rectangular module (Fig 4.39a). The modules are 571 × 45 mm thick and supplied at heights of 815 and 1535 mm which provide module membrane areas in m^2 of 10 (the *SMM-1010*) or 12.5 (*SMM-1013*), and 20 (*SMM-1520*) or 25 (*SMM-1525*), respectively. The *SMM-1520* and *SMM-1525* modules are fitted with two 40 mm permeate headers, whereas the shorter *SMM-1010* and *SMM-1013* modules have a single header at the top. The coarse bubble aerator is integrated with the module, forming part of the base of the ABS frame. A 96-module stack (or skid, Fig. 4.39b) of the *SMM-1520* provides an area of 1920 m^2 within the module having dimensions of 3500 × 1430 × 2410 mm high, and the respective packing densities in m^2 membrane area per m^3 module volume within the module and skid are 159 and 507, respectively. The maximum recommended pressure for both filtration and backwash is 0.5 bar. The recommended membrane aeration rate ranges from 0.05 to 0.15 $Nm^3/(m^2\,h)$.

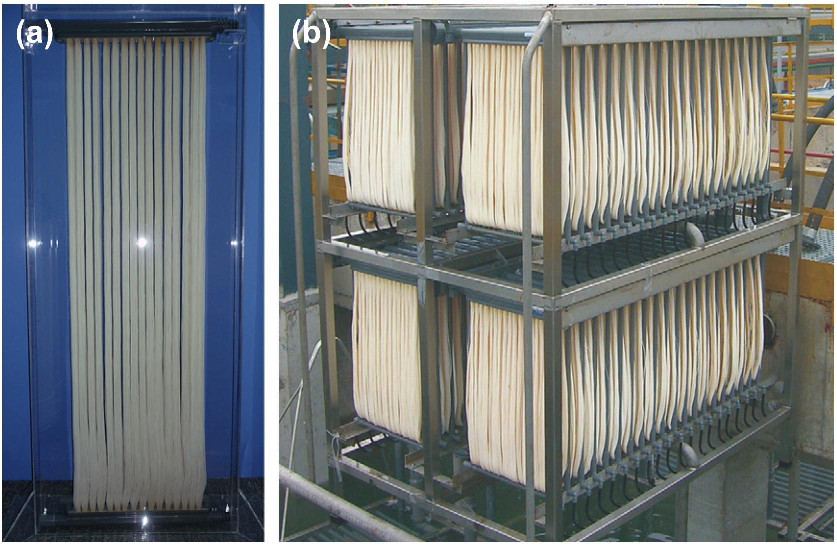

FIG. 4.39 The Memstar: (a) module and (b) stack (or skid).

4.3.11. Micronet Porous Fibers

Porous Fibers S.L. was created in 2000 following several years of research and development into the membrane filtration technology, subsequently registering the name *Micronet*®. The company provides reinforced HF membranes at two different pore sizes – 0.2 μm and 0.02 μm of external diameter 2.45 and 2.1 mm, respectively. The technology is provided as 106 mm square modules, 1935 mm in height, which have a membrane area of 6–7.5 m^2. Ninety-one of these modules form a stack of dimensions 2375 × 1020 × 2590 mm height to provide a membrane area of over 540 m^2. The recommended SAD_m value is 0.4–0.8 $Nm^3/(m^2\ h)$ depending on the application.

4.3.12. Mitsubishi Rayon

Mitsubishi Rayon Engineering Co. Ltd (MRE) represents possibly the largest MBR membrane supplier in Asia and the third largest MBR membrane supplier worldwide with respect to installed capacity, after GE Zenon and Kubota, and, as with these two companies, introduced its original MBR membrane product in the early 1990s (1992). MRE was spun out of the Mitsubishi Rayon Company Limited in 1975. The company, which has a turnover of around $635m, operates in a number of areas relating to polymeric materials, and their product range includes membrane filtration as applied to both the industrial and municipal sectors. As of the end of 2009, the company had around 3000 MBR installations worldwide, predominantly in East Asia.

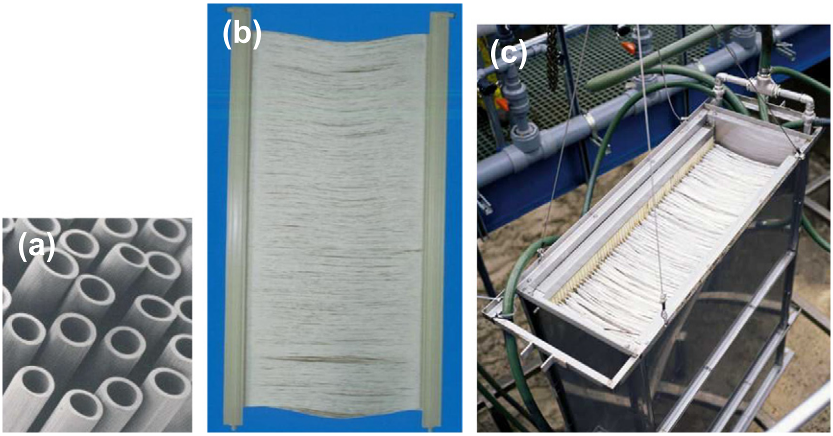

FIG. 4.40 The *STERAPORESUR*™ membrane: (a) fibres, (b) module and (c) single-deck unit.

There are two Mitsubishi Rayon Engineering modules, the original *STERAPORESUR*™ (or simply *SUR*™) introduced in 1995, and the more recent *STERAPORESADF*™, introduced a decade later, which is designed for large-scale municipal wastewaters and features in the Beijing Miyun Plant (Section 5.3.6.1). The *SUR*™ comprises PE filaments of 0.54 mm outer diameter and 60−70 μm wall thickness which are, unusually, horizontally oriented (Fig. 4.40a). For this membrane, slit-like pores of nominally 0.4 μm are generated by stretching to produce an isotropic membrane material. The fibres are potted with polyurethane resin at either end within ABS plastic permeate collection pipes to form the 3 m^2 modules (Fig. 4.40b) which have dimensions of 524 mm wide by 1035 mm long and 13 mm thick. Modules are mounted within a stainless steel frame to form units containing up to 70 modules, providing a membrane area of up to 210 m^2. These can be either single- (Fig. 4.40c) or double- deck, with permeate withdrawn from each deck. The single-deck unit is 1538 × 725 × 1442 mm high.

The *SADF*™ module is based on vertically oriented PVDF fibres. These fibres have a pore size of 0.4 μm, and are 2.8 mm in outside diameter. The modules (Fig. 4.41a) are 2000 mm high × 1250 mm × 30 mm thick, with a membrane surface area of 25 m^2. The modules (Fig. 4.41b) contain 20 elements (and hence provide 500 m^2 total membrane area) and have dimensions of 1555 × 1610 × 3124 mm height.

4.3.13. Motimo

Tianjin Motimo Membrane Technology Co. Ltd is one of the largest manufacturers of HF membranes in China, with an annual HF membranes output of 3 million m^2. It is an industrial high-tech chain enterprise offering R&D,

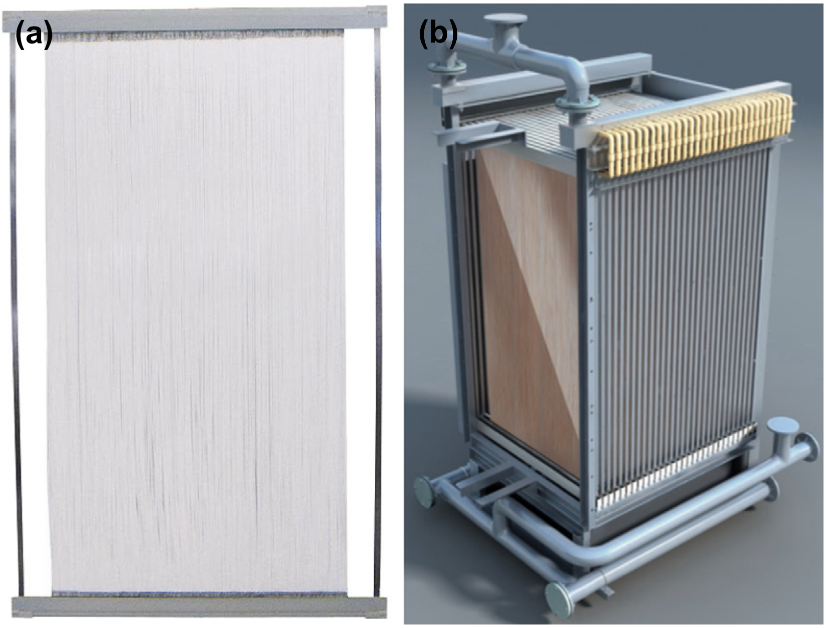

FIG. 4.41 The *STERAPORESADF*™: (a) element and (b) 30-element module.

engineering design and fabrication of membrane materials, modules and equipment based predominantly around their PVDF membrane. Membranes are offered for pumped and submerged configurations for municipal and industrial applications with, in addition to municipal plants, installations in the steel and iron, petrochemicals, textile, food and pharmaceuticals industrial sectors across East Asia, as well as some in Russia and the USA.

Motimo arose from the membrane institute of Tianjin Textile College (now part of the Tianjin Polytechnic University), which has a history of over 36 years of scientific and technology research into membrane materials. The company, originally the Motian Membrane Technology Co., subsequently formed a joint venture with the Lam Group in May 2003 to form the new company. By the end of 2009 the total installed capacity of its membranes across all water applications exceeded 2000 MLD, including the 30 MLD MBR plant for industrial effluent treatment installed in Tianjin in 2007.

The company's *Flat Plat FPII* module comprises 1.2-mm diameter PVDF HF fibres, 0.2 μm in pore size, set in a 534 × 1523 mm high module, with the membrane bundle width being 450 mm. A module, which contains two layers of membranes fed into a 32-mm diameter permeate collection tube (Fig. 4.42a), provides a membrane surface area of 20 m^2. Forty modules are fitted into a 2000 × 1400 × 1700 mm high steel frame to form an 800 m^2 stack (Fig. 4.42b). The quoted SAD_m value is 0.15 $Nm^3/(m^2\,h)$.

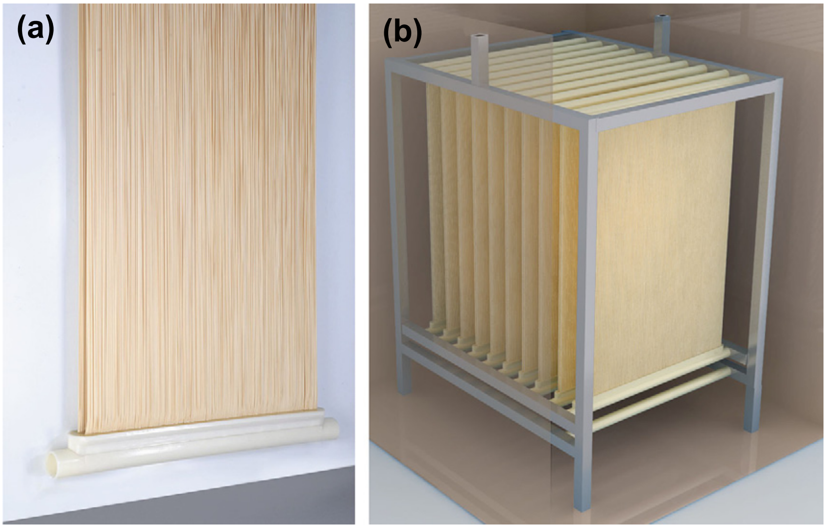

FIG. 4.42 Motimo *Flat Plat* HF: (a) module, showing permeate manifold and (b) a 10-module stack.

4.3.14. Philos

The Philos Company Ltd of Korea was established in June 2002, and provides HF membranes for filtration and gas separation. Their MBR membrane product (Fig. 4.43) comprises 0.1-μm pore size PVDF fibre of 2.35 mm outside diameter and a braided core for reinforcement. The modules comprise bundles of fibres, approximately 75 mm in diameter, which each provides a membrane area of 1.2 m^2. Three module sizes are offered, containing 42, 63 or 105 bundles with overall module lengths of 700, 910 and 1330 mm, respectively, the width (620 mm) and height (1980 mm) being the same for all three modules.

4.3.15. SENUO

The SENUO Filtration Technology Company Ltd in Tianjin produce a range of PES and PVDF HF membranes branded *SENUOFIL*. Their products have apparently been employed within a range of industrial sectors, including pharmaceuticals, food and electrophoretic painting, as well as for pre-treatment in seawater desalination and reclamation of municipal wastewater. The MBR HF membranes produced by the company are 1.3 mm in outside diameter and have a pore size of 0.1 μm. The membranes are formed into 15 m^2, 160 mm diameter bundles which are 1640 or 2150 mm long. The shorter of these forms modules (the *SN-MBR-0660* model) which have dimensions of 1180 × 560 × 2060 mm height and provides a membrane area of 120 m^2 from eight of the

FIG. 4.43 The Philos MBR module.

bundles. The recommended SAD_m value is 0.267 $Nm^3/(m^2\,h)$. One of the few published MBR pilot studies of pulp and paper treatment is based on this product (Zhang, Ma, Ye, Kong, & Li, 2009).

4.3.16. Shanghai Dehong

The Shanghai Dehong Biology Medicine Science and Technology Development Co., Ltd develops products for bio-pharmaceutical, clinical and other pure water industrial users, and has developed an MBR as part of the wider membrane activity which includes MF, UF and MD. The PVDF fibres have a pore size of 0.06–0.08 μm and the module is based on cylindrical bundles which are in the region of 32–50 mm in diameter (Fig. 4.44). Approximately 32 of these bundles sit in a metal frame to form a module 900 × 850 × 2650 mm high and provide a membrane area of 100 m^2.

4.3.17. Siemens

Siemens is a German-based conglomerate with sales exceeding $100bn and a market value of $44bn, according to Forbes. Based on the year to 2009 it was the world's 30th largest company in the Fortune 500 listing, and the fifth largest in Germany. The company employed around 420,000 people in 2009, of which around 6000 work in Siemens Water Technology. SWT operates in a number of

FIG. 4.44 The Shanghai Dehong fibre bundles.

areas of water purification, including desalination, potable water filtration and wastewater reclamation. The industrial sectors in which the group is active are extensive, including oil and gas (produced water treatment), automotive and microprocessors (twin pass RO). The company has acquired a number of established brands such as *Wallace and Tiernan*®, *Stranco*® and *Memcor*® to enhance its water activities.

Memcor® is a well-established membrane brand dating back to 1982. The membrane product originates from Australia. The company was acquired by US Filter in 1997 who were then themselves subsequently acquired by Vivendi (now Veolia) and then by Siemens in July 2004. At the time of the last acquisition, US Filter had a turnover of $1.2b and 5800 employees worldwide.

The first *Memcor*®-based MBR system based on the PVDF 0.04 μm-pore HF membranes was the *MemJet*®, introduced in 2002. In this system air bubbles entrained in mixed liquor were introduced into the module using a patented two-phase mixing system. This design was subsequently modified, and the aeration device changed to produce the *Mempulse*™ system. The module packing density was improved by changing from a cylindrical to a square geometry, 210×210 mm in cross-section and 1600 mm long

(2000 mm overall if the header and the 'skirt' at the base are included). The module, which has permeate and air headers running across the top of it appears to comprise six rows of fibre bundles with the jets of air pulsed between these rows at regular intervals. The pulsing effect is generated using the proprietary *Mempulse*™ device fitted at the base of the module. The B40N module provides an area of 38 m^2, giving a packing density of over 400 m^2/m^3. A standard rack comprises 16 of these modules, providing a membrane area of 608 m^2 for a rack 3960 mm long by 280 mm wide and 2220 mm high.

Memcor® also produces a package plant for flows up to 0.5 MLD. As with the larger-scale systems, the *Xpress* has separate biotreatment and membrane tanks to assist operation and maintenance of the membrane and, in particular, membrane cleaning in place (CIP).

4.3.18. Sumitomo

Sumitomo Electric Industries is a Japanese conglomerate with activities ranging from financial management, media and real estate to metal products, transport and construction, and minerals. Its life insurance and financial activities (the latter as Sumitomo Mitsui Financial) are autonomous functions, and collectively accounted for over $70bn of turnover in the year to 2009 according to the Fortune 500 list. In the same year the Group's other activities accounted for almost $35bn. The MBR product is one of the range of environmental technologies marketed by the Sumitomo Electric Industries (SEI) Group, which employs around 150,000 people and had a turnover of $24bn in the year to 2008 in its consolidated business. The company's original *POREFLON*™ flat sheet membrane was patented in 1962. A hollow fibre product for ozone dissolution was launched in 2001, and the water filtration product launched two years later in 2003. The membrane module was given Title 22 Certification by the State of California in 2009, and there were a number of reference sites in Japan based on the technology by the end of 2009.

The *POREFLON*™ membrane is one of the few new PTFE products on the market for MBR duty and, as such, the fibre itself has the highest chemical resistance of all the MBR membrane products. Unlike PVDF, the membrane has strong alkaline tolerance: the material can be soaked in 4 wt% NaOH for 10 days at 50 °C and suffer no damage, though no similar test has been published on the actual module. The fibre is 2.3 mm in diameter and produced in pore sizes of 0.1 and 0.2 μm. The modules (Fig. 4.45a) are roughly square in cross-section, with dimensions of $164 \times 154 \times 2410$ mm height, and provide a membrane area of 10 m^2. The stacks (Fig. 4.45b) are offered with 10 or 20 of the modules, with the larger 200 m^2 module having dimensions of $344 \times 1880 \times 2881$ mm height; the metal frame in which it sits is somewhat larger at $840 \times 2280 \times 3900$ mm high. The module SAD_m value recommended by the supplier is 0.3 $Nm^3/(m^2\,h)$.

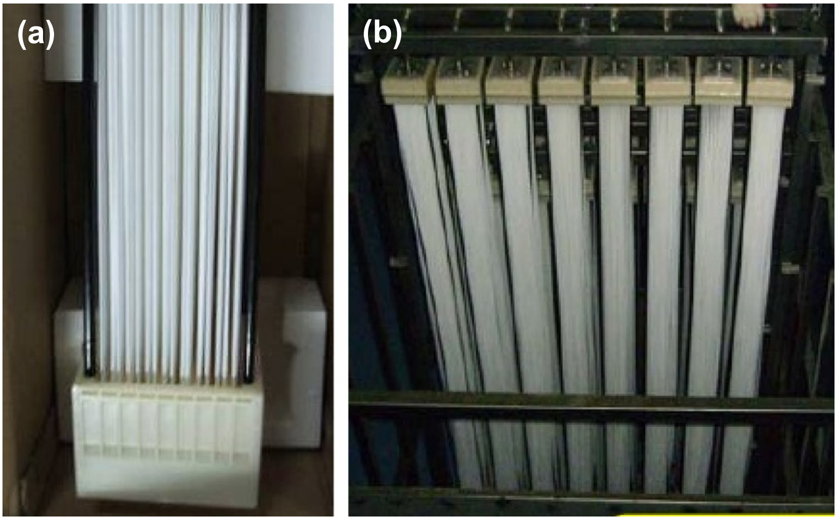

FIG. 4.45 The Sumitomo *POREFLON*™: (a) module and (b) stack.

4.3.19. Superstring

The Superstring MBR Technology Corp. is a small company, which was set up in 2007. As such, the product was still in the early stages of development at the end of the decade, but there were nonetheless two or three reference sites based on the technology by the end of 2009, with the first being installed at the Kong Kee Food Company in New York in 2007.

The 1.25 mm diameter PP *SuperUF* membrane is produced by the TIPS (thermally induced phase separation) method (Section 2.1.2). It has a nominal membrane pore size rated at 0.35 μm according to the bubble point method, though atomic force microscopy measurements indicate a pore size of an order of magnitude lower than this. The membrane elements take the form of narrow panels, 770 × 1140 mm high, which are separated in the module of 10 panels by a distance of 8 mm. Each panel (Fig. 4.46a) provides a membrane area of 1 m^2, and the 10-panel module (Fig. 4.46b) has dimensions of 770 × 330 × 1140 mm high, with a single permeate header at the top of the unit.

4.3.20. Vina

In addition to supplying flat sheet MBR membrane products (Section 4.2.14), the Suzhou Vina Filter Company provide two HF MBR membrane modules, one based on a 0.1-μm pore size PVDF and the other on a 0.2-μm PP. The fibres based on these materials are, respectively, 1.2 and 0.45 mm in outside diameter and formed into cylindrical modules. The *F08* PVDF modules are 1640 mm in overall length, providing 1540 mm effective membrane length, and 160 mm in

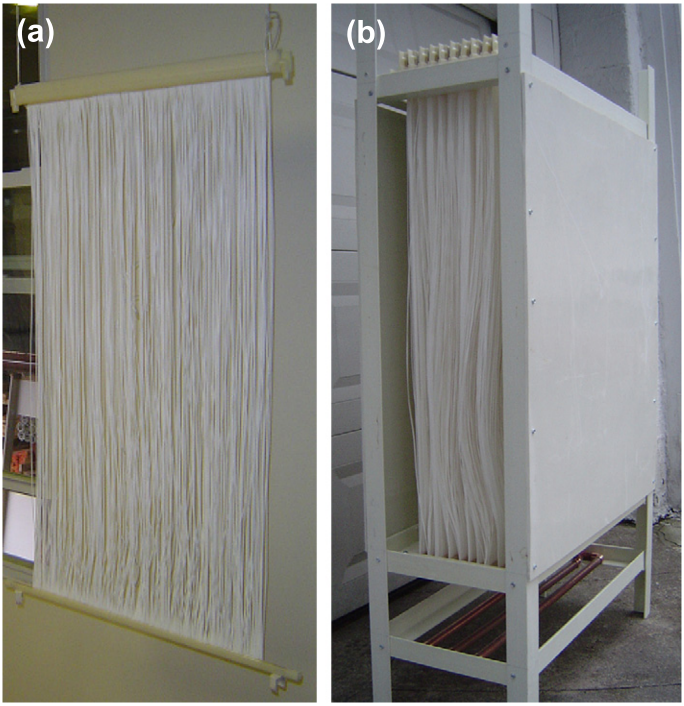

FIG. 4.46 The Superstring: (a) module and (b) stack.

diameter. The smaller PP modules are 800 mm in overall length, 750 mm effective membrane length, and 25 mm in diameter. The surface areas provided by the modules are 15 and 1 m^2 for the PVDF and PP membranes, respectively. The PVDF module comprises a number of small bundles potted at each end (Fig. 4.47a and b), with aeration ports integrated into the module base (Fig. 4.47a).

4.3.21. Zena

Zena SRO was incorporated in 1991 as a research and development company and has been supplying HF membrane modules since then. As well as membranes for separation applications, the company's range of products include gas–liquid contactors, heat exchangers and a photocatalytic degradation technology based on powdered TiO_2. Its submerged HF product recommended for MBR duty is based on a 0.26-mm 0.1 μm pore PP-based *P5* membrane. The module is based on ~25 mm diameter bundles which are 821 mm long, with the active membrane length being somewhat less. The small diameter – one of the lowest of all MBR HF membrane products – means that the packing density at the bundle ends is over 1600 m^2/m^3. The bundles are

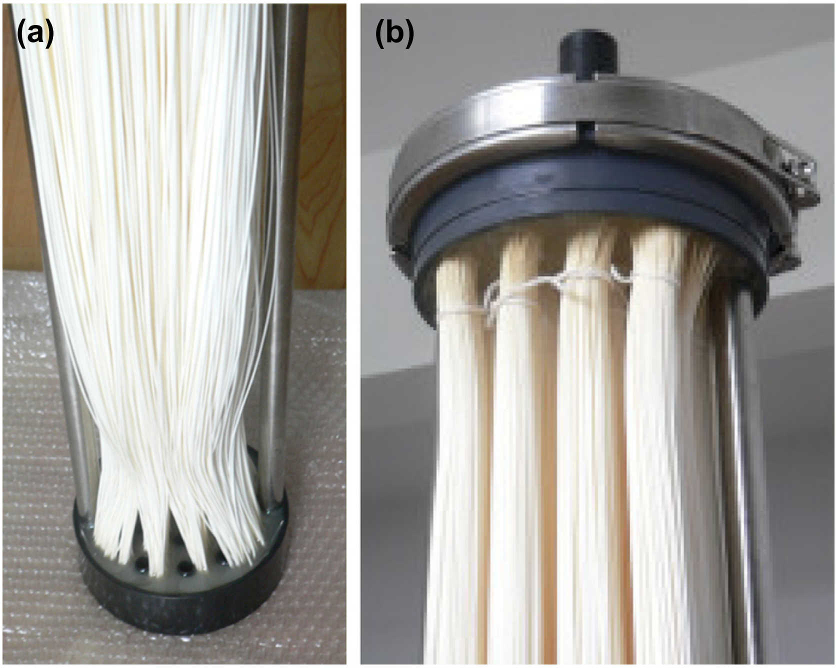

FIG. 4.47 The Vina PVDF module: (a) base and (b) top.

assembled into modules which are provided in various sizes. The 86 m^2 module contains 108 bundles and has dimensions of 784 × 590 × 1010 mm high, and thus an overall packing density of 184 m^2/m^3.

4.3.22. Other HF Products

Other possible MBR HF products for which confirmation and/or comprehensive information could not be obtained by the time this book went to press include Beijing EDI and IWHR, and Kolon and Para of Korea. Kolon have been an established membrane supplier since 1989, and produce a 0.1-μm PS HF submerged membrane (the *Cleanfil-S*®) which would appear to be appropriate for MBR duty. Para produce reinforced HF products in both PVDF and PES. The Beijing Institute of Water Resources and Hydropower Research manufactures UF membrane modules, and Canpure (formerly Beijing EDI Water Treatment Technologies) supplies the *Saveyor SVM* MBR technology which is again based on PVDF. There are additionally a large number of other Chinese products, of which technical detail is generally limited but at least some of which appear to be original products. These are listed in Table C.1 of Appendix C.

There is additionally the specialist MBR membrane supplier MEMOS from Germany, who provide both HF and MT membranes (Section 4.4.3). For

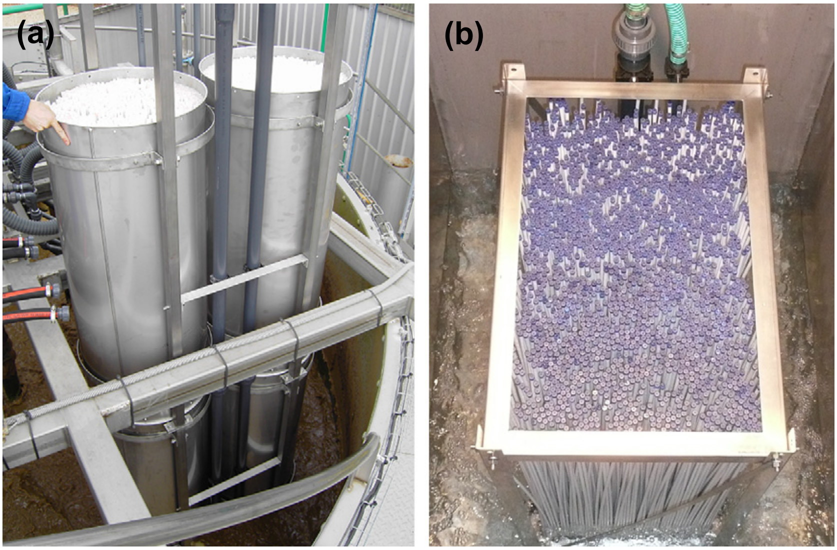

FIG. 4.48 The *MEMSUB* configurations: (a) circular and (b) rectangular.

internal, submerged MBR applications the company produce a membrane module (*MEMSUB*) based on tubular membranes with 8 mm diameter and membrane coating on the outside to operate out-to-in, and are thus in essence wide-bore HF membranes. The membranes are arranged vertically in elements with a bottom permeate collector incorporating the aeration slots and with individually sealed, free-moving membranes. The membrane elements are available in circular (smaller installations, Fig. 4.48a) and rectangular (larger installations, Fig. 4.48b) geometries. Up to three such elements can be stacked.

4.4. SIDESTREAM MBR PRODUCTS

There are a number of multitube (MT) membrane module suppliers providing standard size cylindrical modules, generally 100 mm or 200 mm in diameter, which are then employed by sMBR process suppliers in their proprietary processes. Details of some of the membrane module product suppliers are given below.

4.4.1. Norit

The Dutch-based Norit Group of companies provide water purification equipment and technologies for the municipal and industrial sectors, having begun life in 1918 as a carbon supplier for decolouration in sugar refining. Its current range of products include activated carbon, membranes, pumps, aseptic and hygienic valves, carbon dioxide systems, and quality control equipment. The Group's turnover was in the region of $0.6b in 2008, and the company had around 1700 employees worldwide at that time.

Its membrane activities are covered by Norit X-Flow, which supplies capillary tube (CT) – often referred to as HF, though the flow is in-to-out for this product (Section 2.1.3) – membrane modules for low suspended solids applications such as potable and industrial process waters. It is this group which deals with those applications using its MT product, which was obtained with the acquisition of Stork Friesland in 2000. There are also other parts of the Group which are dedicated to specific sectors or applications, such as beer or point-of-use filtration. Coordination of these activities across the Group is through Norit Process Technology, which is the company within which the MBR activity sits and acts as an integration manager within the Norit Group. Today the total global installed capacity of Norit membrane products is probably in the region of 5000 MLD, with around 150–200 MLD of this figure provided by its existing MBR plants.

The tubular membrane employed for MBR duty is based on a mechanically robust substrate with a two-layer polyester backing for the PVDF membrane. The MT membrane products comprise type *38PRV F4385*, 5.2 mm internal diameter (Fig. 4.49), and type *38GRH F5385*, 8 mm ID (internal diameter), each having a pore size of 0.03 μm. The PVDF MT modules are supplied as both PVC (*38PR* prefix) or glass fibre-reinforced (*38GR*) shells. At 3 m in length, the respective total membrane area and number of lumens is 33 m^2 from 700 lumens (for the 5.2 mm ID) or approximately. 27 m^2 from 365 lumens (8 mm ID). The most established of the Norit sMBRs are based on pumped, horizontally mounted cross-flow modules (Fig. 4.50a). However, the company has also developed the sidestream air-lift system in which the membranes are vertically mounted (Fig. 4.50b). In this system aeration of the modules is combined with liquid pumping at much lower flow rates than those employed for the sidestream system (Table 4.3). The most recent product, launched in 2010, is the *Megablock*, which permits up to 216 vertical modules and is thus suited to larger scale air-lift applications than those usually employing the skid-mounted system shown in Fig. 4.50b.

The MT Norit membrane has been used by a number of process designers for proprietary MBR technologies (Section 4.4.5). Whilst the majority of these relate to relatively small industrial effluent applications, there are an increasing number of larger scale municipal applications based on the air-lift configuration. The latter is recommended for relatively low COD concentrations and high flows (<1 g/L COD and >250 m^3/h), and this includes industrial and municipal effluents, whereas at high COD levels (>5 g/L) and low flows (<100 m^3/h), the pumped system is recommended. This leaves a range of flows and COD levels for which either system may be suited, but the overall COD loading rate over the entire range is normally between 100 and 1000 ppm COD/h.

TABLE 4.3 Norit X-Flow sMBR Operational Parameters

Parameter	Pumped	Air-lift
MLSS, g/L	12–30	8–12
TMP, bar	1–5	0.05–0.3
Flux, LMH	80–200	40–65
Permeability, LMH/bar	40–80	150–600
Footprint, m^3/h capacity per m^2 projected area*	10.8#	7.5**
Footprint, m^2 membrane per m^2 projected area*	108	162
Specific energy demand, kWh/m^3	1.5–4	0.5–0.7
Processing	More simple	More complex
Mode of operation	Continuous	Discontinuous

*Based on a single skid, 1 m × 4 m × 4 m high.
**Based on 55 LMH maximum gross flux, 40–48 LMH net flux: a 'staggered air-lift skid' has a 1.7 × 3.6 m footprint and contains 990 m^2.
#Based on 100 LMH.

4.4.2. Berghof

The Berghof Group from Germany originally began in 1966 as a private research institute (the 'Physikalisch-Technisches Laboratorium Berghof GmbH') with a remit to commercialize research results in the fields of electrochemistry, membrane filtration and plastics technology. They have been

FIG. 4.49 The Norit X-Flow *38PRV F4385* module.

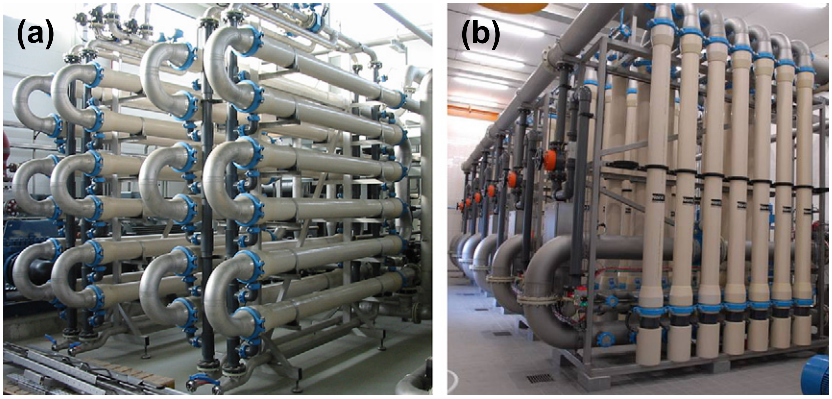

FIG. 4.50 Norit X-Flow modules: (a) pumped and (b) air-lift.

active in MBR technology since the turn of the Millennium, and there were more than 190 MBR projects based on Berghof products as of March 2010.

Berghof Membrane Technology GmbH and Co. KG provide a range of MT module products (Fig. 4.51), up to 4 m long and at internal diameters ranging from 5 to 12.7 mm. Within their MBR product range the *HyperFlux-I8LE* is a 4 m length, 10′ diameter MT module fitted with backflushable 8 mm ID PVDF tubes of 0.03 μm pore size, providing an overall module area of 53.4 m^2. A skid of 10 vertically mounted modules, operated in air-lift mode, thus provides an area of 534 m^2, with a footprint of 3.1 × 1.5 m and a height of 4.9 m. The

FIG. 4.51 Berghof MT membranes and modules.

HyPerm-AE module is fitted with 11.5 mm diameter tubes providing a total membrane area of up to 12.1 m^2 at module lengths of 3 m.

The company offers three MBR technologies, *BioFlow*, *BioPulse*™ and *BioAir DS*™ (Fig. 4.52), employing its MT modules in conjunction with a nominal 5-m-high biotank. The *BioFlow* is based on conventional pumped crossflow, with a feed and recirculation pump in the sidestream designed to maintain a crossflow of 3.5−4.5 m/s. The *BioPulse*™, recommended for less complex and medium-strength wastewater, employs 8 mm diameter tube *HyperFlux LE* modules under more benign conditions of crossflow (1−2 m/s). The module is backflushable to alleviate fouling and clogging and so maintain permeability. The more recent *BioAir DS*™ technology has been developed for the treatment of municipal and other less complex and concentrated wastewaters. It features vertically mounted *HyperFlux-I8LE* modules with an air distributor (*Distair*) fitted at the top of the module to allow air to be injected

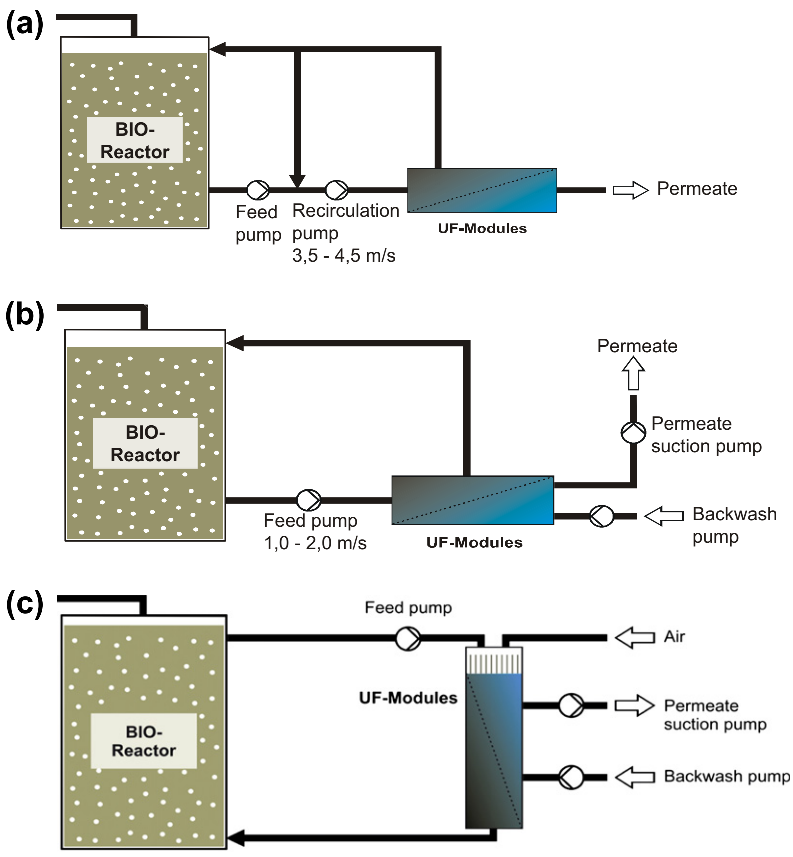

FIG. 4.52 The Berghof MT technologies: (a) *BioFlow*, (b) *BioPulse*™ and (c) *BioAirDS*™.

simultaneously into each individual membrane tube at a fixed and precise rate. This provides an optimized and evenly distributed air/water mixture across the module and skid.

4.4.3. Other Multitube Membrane Products

There exist a large number of multitube/multichannel membrane product suppliers worldwide, with the products predominantly marketed and employed for industrial and, in some cases, municipal potable water supply applications. Examples of some of these products which have been trialled or employed for sidestream MBR duties are given in Fig. 4.53. Three of these products (the Pall *Exekia*®, the Novasep *Kerasep*®/*KLEANSEPR*® and the Veolia *Ceramem*®) are ceramic monoliths. Such membranes would normally be considered uneconomic for MBR technologies, but have nonetheless been used for MBR duties at small scales.

An interesting recent development is the formation of the German company MEMOS Membranes Modules Systems GmbH. MEMOS offers a wide range of tubular ultrafiltration membrane products targeted at MBR applications. Installation of MEMOS membrane modules can be external (cross-flow) or internal (submerged, Section 4.3.22). The company specializes in offering bespoke modules (*MEMTUBE* and *MEMCROSS*) based on tubular membranes

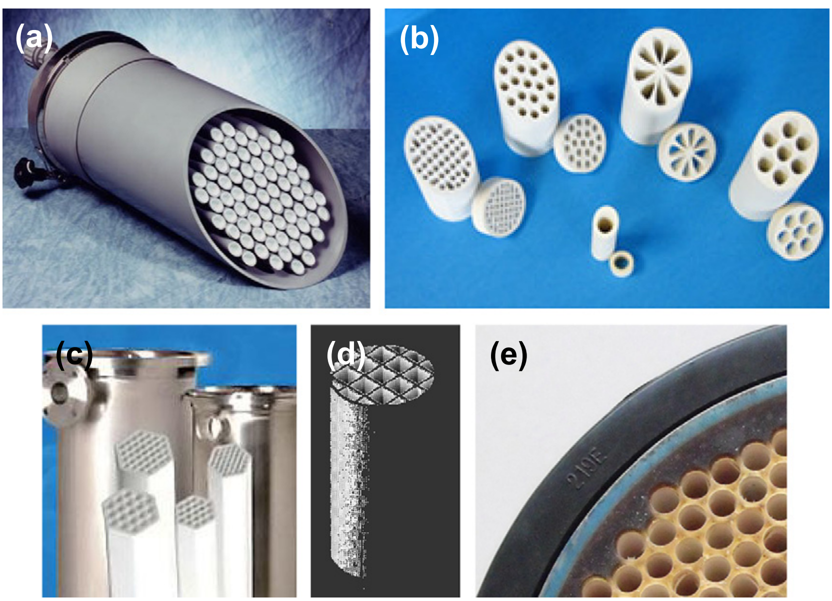

FIG. 4.53 Membrane multube/multichannel products: (a) PCI, (b) Novasep *Kerasep*®/*KLEANSEPR*®, (c) Pall Exekia®, (d) Veolia Ceramem®, and (e) Vina.

of various diameters, from 4 to 24 mm, with lengths of 3 and 3.6 m as standard, but with other sizes available as well. The tubes can be produced with the membrane coating on the inner surface, for external, cross-flow MBR applications. For this application individual membrane tubes are bundled to multi-tubular membrane elements which are installed in stainless steel housings to form complete membrane modules. The membrane modules are offered in sizes of 75−250 mm diameter with membrane areas of up to 53 m^2 (for a 3.6-m length module). The company additionally produces wide-bore HF modules for submerged applications (Fig. 4.48).

4.4.4. Other sMBR Product Suppliers

The French company Novasep provides high-end products for separation applications within the life science industries, and was founded in 1995 as Novasep Process. In 2004 it acquired the company Orelis (formerly Rhodia), which was originally formed as part of the break-up of the Rhone Poulenc group, creating Orelis Environment SAS in January 2009 dedicated to environmental applications of its membrane technologies. As such the company retains a number of the original products, including the FS UF *PLEIADE*® module. This product (Fig. 4.54) has one of the longest histories of use in sMBR applications, although the ceramic module (now part of the Novasep Process range of products) has also been employed. The current *PLEIADE*® element (Fig. 4.54a) comprises a 971 mm by 310 mm polyacrylonitrile-based FS rated at 40 kDa (∼0.015 μm) pore size. The membrane elements have inlet and outlet ports, which allow passage between each cell (i.e. flat sheet channel), and the membrane channel thickness is 3 mm. The membrane holding plates then have dimensions of 1140 mm long and 290 mm wide, and the *MP4*

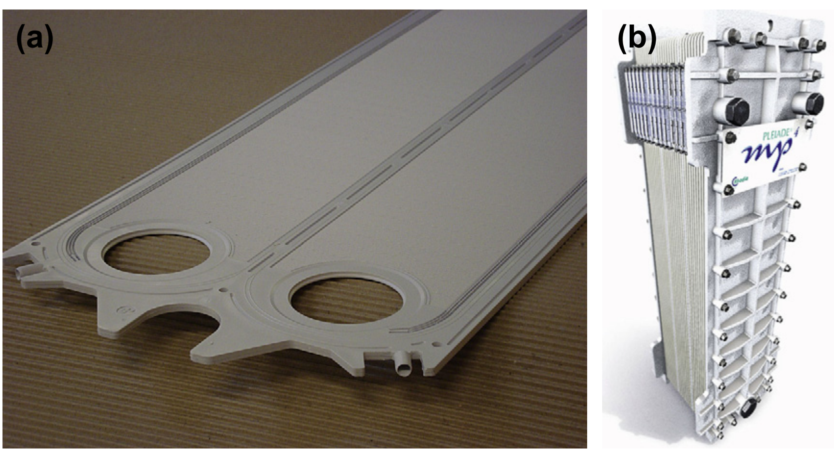

FIG. 4.54 (a) Detail of *PLEIADE*® membrane element; and (b) *MP4* module.

module (Fig. 4.54b) formed from the membrane sheets provides a membrane area of 70 m^2.

There appear to be two small companies each providing an HF-based membrane module configured for sidestream use. The proprietary name of the module provided by the French company Polymem, founded in 1997, is *Immem*. The polysulphone (PS) membranes used are 0.08 μm (300 kDa) pore size and are provided with external diameters between 0.7 and 1.4 mm in modules which are 315 mm in diameter and 1000−1500 mm in length (Fig. 4.55a). The membrane area is thus between 60 and 100 m^2 per module. The Singaporean company Ultra-Flo Pte Ltd, now part of the Mann and Hummel group, is one of the few suppliers of polyacrylonitrile (PAN) HF UF membranes for municipal water and wastewater applications. The filament diameter is 2 mm, and the dimensions of the module (Fig. 4.55b) are 1524 by 203 mm diameter, providing a membrane area of 48 m^2. The packing densities of these sidestream HF products, in terms of m^2 membrane per m^3 module volume, are amongst the highest of all MBR membrane products.

Finally, and more unusually, there appear to be sidestream technologies based on ceramic discs, which include a membrane module product and a technology. The German company Kerafol supply a 0.06 μm pore size, 6 mm thick and 312 mm diameter rotating disc membrane (Fig. 4.56a) providing an area of 0.14 m^2. This disc can be stacked to provide a 48-disc unit operated as a sidestream. The system has been demonstrated by the Fraunhofer IGB at the small village of Heidelburg-Neurott (60 p.e.) in trials which began at the start of 2006 as part of a nationally funded research programme ('DEUS 21'). The

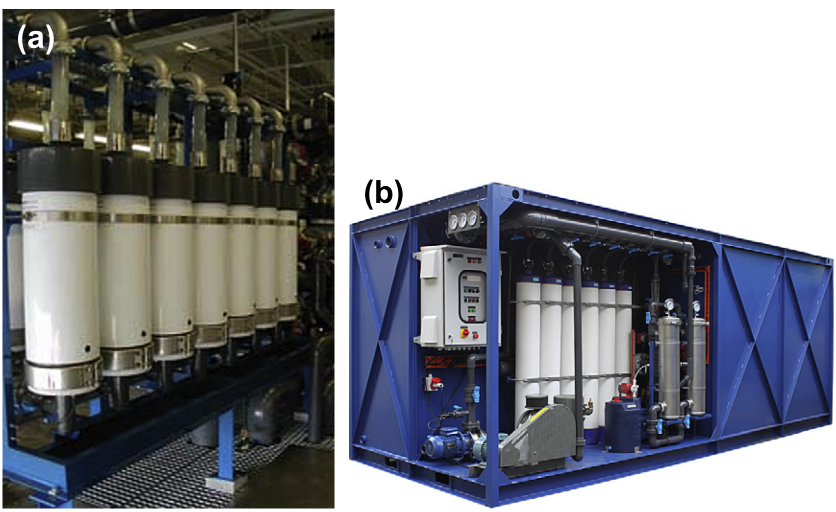

FIG. 4.55 HF sMBR modules: (a) Polymem *Immem* and (b) Ultra-Flo/MANN+HUMMEL MBR-50 unit, fitted with *U860* cartridges.

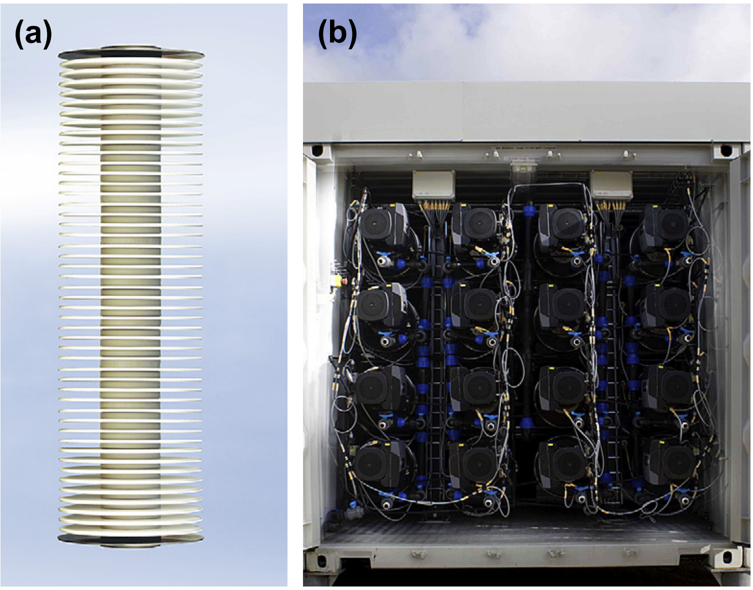

FIG. 4.56 Disc systems: (a) the Kerafol module and (b) a Grundfos *BioBooster* stack of 16 vessels.

BioBooster is produced by the Danish Grundfos Group, which is an international organization with a turnover of \$3.5b and close to 17,900 employees worldwide as of 2008. In this technology both the bioreactor and the membrane discs are inserted in series into a single cylindrical vessel (or reactor; Fig. 4.57), 16 of which can be fitted into a standard 20-foot container (Fig. 4.56b). In this case shear is provided through the use of rotating impellers between the ceramic UF membranes, or possibly rotation of the discs themselves, and atmospheric air is used to sustain the biomass. It is thus one of the most intensive of all the MBR technologies, able to operate at pressures of up to 5 bar and sustain biomass concentrations of up to 40 g/L, and completely modular since both the bioreactor and membrane separation components are incorporated into the same standard-sized tube.

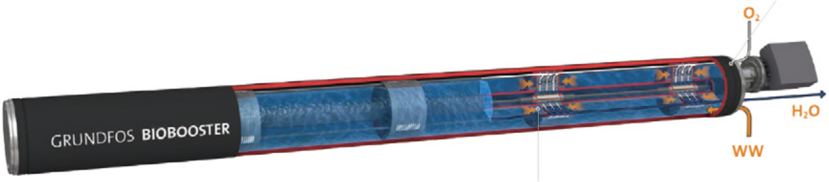

FIG. 4.57 The Grundfos *BioBooster* technology.

4.4.5. Other sMBR Technology Suppliers

As with FS membranes, there are a number of membrane technology suppliers who use existing commercially available membrane products for technologies they have developed for specific applications. This is perhaps more widespread for MT MBR technologies due to standardization of the MT membrane modules themselves; they are provided in standard diameters and lengths and are still quoted in imperial units, 8″ diameter (200 mm) being one of the most common sizes. Three such technology suppliers are outlined below. All these suppliers focus their activities largely on industrial effluent treatment and reuse applications and, in the case of the latter, all are usually able to provide both the sMBR and the downstream NF or RO plant for the complete reuse system. All provide both pumped and air-lift technologies.

Wehrle Environmental, part of the German Wehrle Group which has been operating for almost 150 years, is arguably the most established European company providing MBR technology for landfill leachate and industrial wastewater treatment. The company uses primarily Norit and Berghof membranes and most of the installations since 2000 have been based on standard 8″ (200 mm diameter) $\times$ 3 m tubular membrane modules incorporating 8 mm diameter tubes. The Wehrle *BIOMEMBRAT*® process is an MBR which can be combined with either an atmospheric or pressurized bioreactor (up to 3 bar) depending on the process circumstances, with jet aeration used for the process tank for enhanced oxygen transfer. Bioreactor pressurization offers the advantages of: (a) sludge foaming control, (b) enhanced oxygen dissolution, thus permitting higher organic loadings and/or reduced tank size and (c) reduced risk of stripping of volatile organic matter, and so reducing the size of any air scrubbers which might be required for off-gas treatment. The company has also developed integrated post-treatment processes to provide additional purification. Unit operations have included both activated carbon (AC) and nanofiltration (NF) where the NF concentrate is fed to the AC for organics removal, with the AC effluent (Fig. 4.58).

Aquabio is a UK company with over 10 years' experience in the design and installation of sMBRs and specializes in the application of sMBR technology and tertiary treatment for water reuse. To date, the company has installed 15 full-scale sMBR plants within the food, beverage, biofuels, pulp and paper, landfill leachate, pharmaceuticals and tannery industrial sectors, nine of which involve additional RO treatment to allow water reuse in food applications as well as boiler feed. While the company also provides submerged MBR technologies, Aquabio focuses mainly on MT sMBRs.

The company's three key MT MBR brands are *AMBR*™ (high flux cross-flow) and *AMBR LE*™ (low-energy cross-flow) and *BIOVERT*® (low-energy air-assisted cross-flow), all employing different operating conditions (Table 4.4). The *AMBR*™ uses high cross-flow to achieve high flux rates, and can be operated at high MLSS concentrations. It is aimed at lower flow and higher

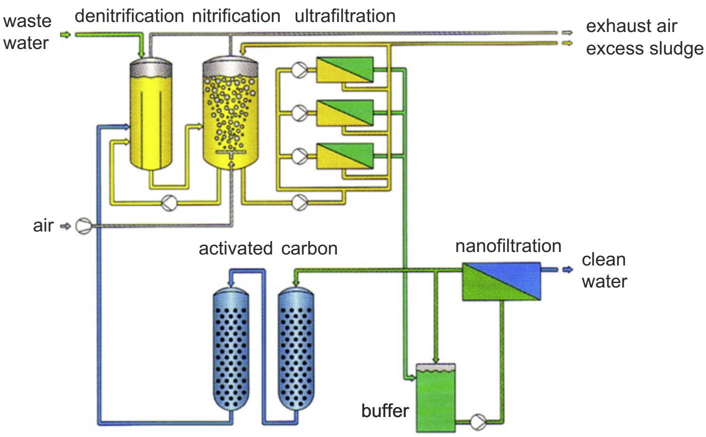

FIG. 4.58 The Wehrle *BIOMEMBRAT*® with integrated post-treatment.

strength industrial wastewaters. The plants are extremely compact, with very low membrane area and low membrane replacement costs. The membrane banks are operated automatically based on the bioreactor sludge level; membrane banks are employed and automatically shut down as required to match the inlet flow, auto-flushing to remove biomass before leaving on standby.

The *AMBR LE*™ is a low-energy design using intermittent backflushing to control membrane fouling and allow a reduced CFV and so a significant saving in energy. A medium to high MLSS concentration is used at medium to

TABLE 4.4 Comparison of Aquabio MBR Technologies

MBR Type	Normal Operating MLSS Range (g/L)	Sustainable Normalized Flux LMH	Filtration Energy Demand, kWh/m^3 Permeate Produced	Configuration/ maintenance
AMBR™	10–20 (Air) 15–35 (Pure oxygen)	80–250	1.8–3.5	'Dry', out of tank. Low level, horizontally mounted membranes
AMBR LE™	10–20	40–120	0.4–1.5	'Dry', out of tank. Low level, horizontally mounted membranes
BIOVERT®	8–15	30–60	0.2–0.5	'Dry', out of tank. Vertically mounted membranes

high flow and low to high strength wastewaters. With the inclusion of variable speed recirculation pumps the system allows for variable flux rates which can be used to optimize energy use relative to the plant load. Plants remain compact, with low membrane area and moderate membrane replacement costs. They operate on a direct link to the level in the bioreactor, automatically adjusting membrane permeate production (and energy use) to suit the inlet flow conditions. The process is suited to applications with inconsistent or variable wastewater flows, or high peak or seasonal loads, or when electricity costs are moderate to high.

The *BIOVERT*® uses vertically mounted modules with air injection to further reduce energy use. This process allows for medium MLSS concentrations and is aimed at higher flows and low-to-medium-strength industrial and municipal wastewaters. The vertically mounted membrane modules allow a low footprint, and the plants operate at lower flux and with a more constant bioreactor level. Controlled air injection is either at the top or base of the membrane tubes. Air scour and turbulence reduces surface fouling and energy use is low. The process is used for applications with consistent wastewater flows and when electricity costs are high.

Three proprietary MT membrane products are employed over the range of technologies, two of these being Norit and Berghof. Recent improvements in membrane module design have led to the installation of 10″ (250 mm dia.) × 4 m modules providing twice the surface area available with the standard 8″ module. In addition this module has an increased backpressure capacity (up to 0.5 bar), allowing modules to be backflushed to control fouling (as used in the *AMBR LE*™ process).

The New Jersey company Dynatec Systems, Inc., using both 5.2 mm and 8.0 mm tubular membrane modules, provides the *Hi-Rate* pumped cross-flow and the *DynaLift* air-lift sMBR sidestream (or 'out-of-basin') membrane process configurations. The company has installations at cereal plants, both sanitary and hazardous landfills for leachate treatment, and many other industrial and commercial installations. The *DynaLift* technology uses 5.2 mm diameter vertically aligned Norit MT modules, assembled in stacks of 6 or 12. The *Hi-Rate* technology employs 5.2 mm or 8 mm tubular membrane modules in series. In 2011, the company expects to deploy an upgrade that simplifies the system, combining the benefits of both the *DynaLift* and *Hi-Rate* technologies.

There is also an established anaerobic sMBR technology. The *BIOREK*® process (du Preex, Norddahl, & Christensen, 2005) has been developed by Bioscan A/S in Denmark primarily for treating animal manure slurries and is an extension of the cross-flow anaerobic thermophilic sMBR originating from the Wier EnVig *ADUF* process. The process provides high-rate thermophilic (50−55 °C) anaerobic digestion with the biomass, high molecular weight organic matter and colloids being retained by 25−100 kDa pore size 12 mm diameter cross-flow PS MT UF membranes. The filtrate, containing low-molecular weight organic molecules, such as volatile fatty acids (VFA), and dissolved inorganic materials, is fed to a liquid–liquid PP or PVDF membrane contactor for ammonia stripping. Ammonia is extracted under the prevailing conditions of high pH and temperature into an acid solution on the permeate side. The extracted filtrate is then fed to a conventional reverse osmosis stage which provides deionized water for steam raising or other operations, and a concentrate stream with a high nutrient content. The biogas from the reactor may be desulphurized, again possibly using a membrane contactor, to recover the sulphur.

4.5. TECHNOLOGY SUMMARY

A review of commercially available MBR membrane products (Table 4.5) reveals there to be at least 42 iMBR FS and HF membrane module suppliers and a further eight or more sMBR product suppliers. There are therefore at least 50 individual suppliers offering perhaps 60 discrete membrane products with respect to configuration and/or membrane material. There are perhaps 3–4 times this number if the different sizes of the individual FS panels, HF modules and MT elements are considered, and many hundreds of stack/cassette products. In the case of the sidestream MT technologies, based on standard-sized modules, suppliers tend to provide three types of systems: classical pumped, low-energy pumped with maintenance cleaning and air-lift.

In producing this summary it must be acknowledged that a large number of products from the Far East, and China in particular, are visible through specialist trade web sites (such as alibaba.com and made-in-china.com), but that their technical description is often limited. A least some of these appear to be original products, albeit possibly with few, if any, reference sites. However, determining the original manufacturer is sometimes challenging.

Polymer materials used for MBR membranes are largely limited to two fluorinated polymers (polyvinylidene difluoride, PVDF; and polytetrafluoroethylene, PTFE), two sulphonated polymers (polyethersulfone, PES; and polysulphone, PS) and two polyolefinic membranes (polypropylene, PP; and polyethylene, PE), with a pore size ranging from 0.01 to 0.4 μm (Fig. 4.59). The combination of good chemical resistance and surface structure (Section 2.1.2) has meant that the sulphonated and fluorinated polymers, and PES and PVDF in particular, dominate in modern MBR membrane materials. PES/PS membranes are mostly in the ultrafiltration (UF) pore size range and make up ~20% of the iMBR membrane materials listed. The PTFE and polyolefinic membranes, on the other hand, are all in the microfiltration range (0.8–0.4 μm). However, polyvinylidene difluoride (PVDF) membranes make up 55% of the total iMBR membrane technologies considered and cover almost the entire pore size range (between 0.04 and 0.4 μm). The only other polymeric membrane materials employed are polyacrylonitrile (PAN), employed in two

TABLE 4.5 Summary of commercial MBR membrane module products (adapted from Santos & Judd, 2010)

Immersed (iMBR)		Sidestream (sMBR)
Flat sheet	Hollow fibre	Multitube/multichannel
A3 – *MaxFlow* ^{DE} Agfa-VITO ^{BE} Alfa Laval – *Hollow Sheet* ^{SE} Brightwater – *MEMBRIGHT*® ^{IRL} Colloide – *SubSnake* ^{NIR} Ecologix – *EcoPlate*™, *EcoSepro* $^{TM\ CN}$ Huber – *VRM*®; *ClearBox*®, *Biomem* ^{DE} Hyflux – *Petaflex* ^{SG} Jiangsu Lantian Peier Memb. Co. Ltd ^{CN} LG Electronics – *Green Membrane* ^{KR} Kubota – *ES/EK* ^{JP} MICRODYN-NADIR – *BioCel*® ^{DE} Pure Envitech Co., Ltd. – *ENVIS* ^{KR} Shanghai Megavision Memb. Engng. and Technol.Co., Ltd ^{CN} Shanghai SINAP Membrane Science & Technology Co., Ltd. ^{CN} Toray – *MEMBRAY*® *TMR* ^{JP} Suzhou Vina Filter Co. – VINAP ^{CN} Weise Water Systems GmbH – *MicroClear*® ^{DE} **Other developing technologies** Inge – *FiSh* ^{DE} IWHR ^{DE}	Asahi Kasei – Microza $^{TM\ JP}$ Beijing Origin Water Technology Co. ^{CN} Canpure – *Canfil* ^{CN} Ecologix – *EcoFlon*™, *EcoFil* $^{TM\ CN}$ ENE Co., Ltd. – *SuperMAK* ^{KR} GE Zenon – *ZeeWeed*® ^{US} Hangzhou H-Filtration Mem. Technol. & Engng Co., Ltd. – *MR* ^{CN} Koch Membrane Systems – *PURON*® ^{US} Korea Membrane Separations – *KSMBR* ^{KR} (Hainan) Litree Purifying Technol. Co. Ltd. – *LH3* ^{CN} MEMOS Membranes Modules Systems – GmbH – *MEMSUB* ^{DE} Memstar Technol. Ltd – *SMM* ^{SG} Micronet Porous Fibers S.L. – *Micronet*® ^{SP} Mitsubishi Rayon Engng. *Sterapore –SUR*™; *SADF* $^{TM\ JP}$ Mohua Technology – *iMEM-25* ^{CN} (Tianjin) Motimo – *Flat Plat FPII* ^{CN} Philos Co. Ltd. ^{KR} SENUO Filtration Technol. Co., Ltd. – *SENUOFIL* ^{CN} Shanghai Dehong Biology Medicine Sci. & Technol. Dev. Co., Ltd. ^{CN} Siemens Water Tech. – *MemPulse* $^{TM\ DE}$ Sumitomo Electric Industries – *POREFLON* $^{TM\ JP}$ Superstring MBR Technol. Corp. – *SuperUF* ^{CN} Suzhou Vina Filter Co. – *F08* ^{CN} Zena SRO – *P5* ^{CZ}	Berghof – HyPerm-AE; HyPerflux ^{DE} Norit X-Flow – F4385, F5385 ^{NL} Orelis Environment –*PLEIADER*®, *KLEANSEPR*® ^{FR} MEMOS – Membrance Modules Systems GmbH – *MEMCROSS* ^{DE} **Hollow fibre** Ultra-flo® ^{SG}/Mann and Hummel ^{DE} Polymem – *IMMEM* ^{FR} **Flat disc ceramic** Kerafol ^{DE} Grundfos – *Biobooster* ^{DK}

AT: Austria; BE: Belgium; CN: China/Taiwan; CZ: Czech Republic; DE: Germany; DK: Denmark; FR: France; IRL: Ireland; JP: Japan; KR: Korea; NIR: Northern Ireland; NL: Netherlands; SE: Sweden; SG: Singapore; SP: Spain; US: United States
See Appendix C for other products from the Far East

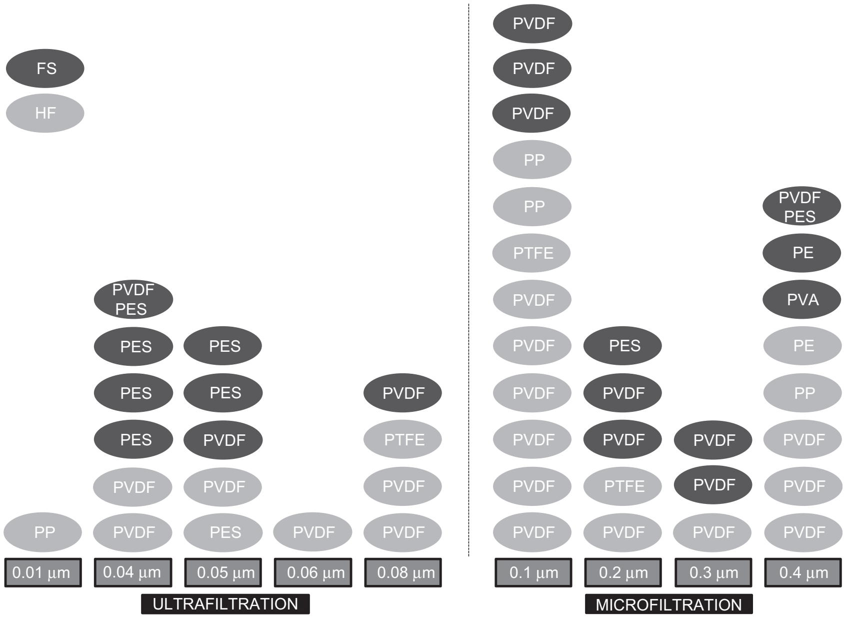

FIG. 4.59 MBR membrane materials and pore size (45 polymeric products) (adapted from Santos, & Judd, 2010).

sidestream products, and polyvinyl alcohol (PVA) in an immersed FS technology. There is a greater diversity of materials used for HF membranes than for FS ones, with those used for FS panels currently being largely limited to PES and PVDF, though the FS membrane market leader (Kubota) employs a chemically modified PE material.

Polyolefinic HF membranes are generally produced by the relatively simple process of dry spinning, which generates slit-like or ovular pores (Fig. 2.4a) therefore having a wider distribution of pore dimension. This, along with the relatively low pore density, tends to make the membrane slightly more susceptible to fouling and thus may necessitate lower flux operation than that of other membrane materials. This can be countered to some extent by producing modules of higher packing density and thus smaller diameter filaments, and this is reflected in the commercial trends (Fig. 4.59). All but one of the polyolefinic HF products are sub-1 mm in filament diameter, and conversely all but one of the non-polyolefinic-based products are above 1 mm in size.

Two practical considerations regarding MBR technologies are their cost and interchangeability. Whilst obtaining unambiguous information on capital costs is challenging, the technical specifications of the commercial iMBR membrane module products allow an assessment of both the module footprint, and thus

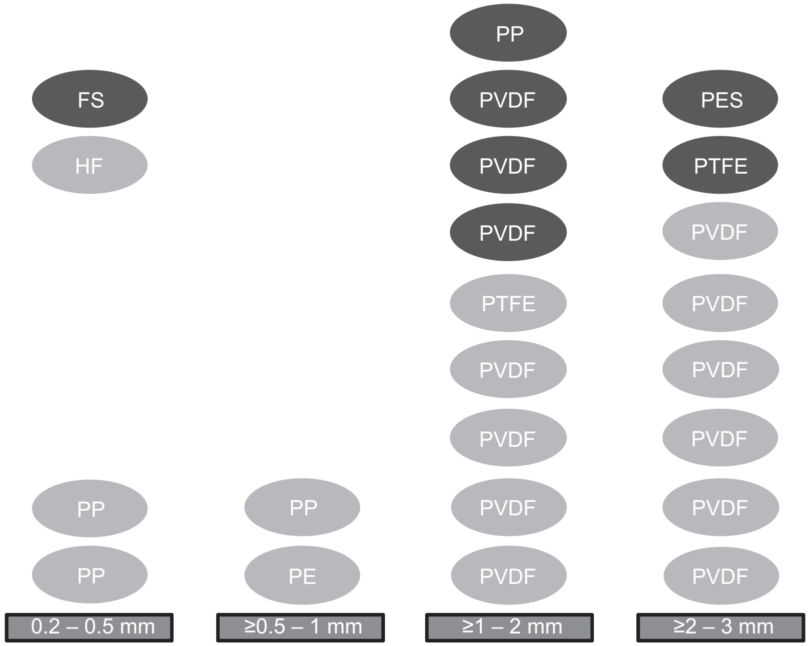

FIG. 4.60 HF filament diameters and material (21 polymeric products) (adapted from Santos, & Judd, 2010).

their interchangeability, and the membrane aeration demand, which provides the greatest contribution to the running costs (Section 3.5.2).

Whilst the multitube sMBR products are largely standardized, the majority being based on a 200-mm diameter module, the same is not true of the immersed products. Module specification data (Appendix C) can be used to determine two key iMBR parameters defining the relative spatial occupation of the membrane modules:

A_m:F The unitless ratio of the module membrane area A_m to its footprint F, where F is the cross-sectional area at the module base; and

φ The module packing density, or the membrane area per unit module volume in m^{-1}; this equates to the A_m:F ratio divided by the module height.

The distribution in values of the above two parameters (Figs 4.59 and 4.60, respectively) then provides an indication of the extent to which the modules can be interchanged, assuming this to be constrained primarily by space occupation. In these two figures the two configurations are categorized according to:

- the number of decks of flat sheet modules (1–3) and
- the geometry of the bundles/elements of hollow fibres in a module (cylindrical or rectangular).

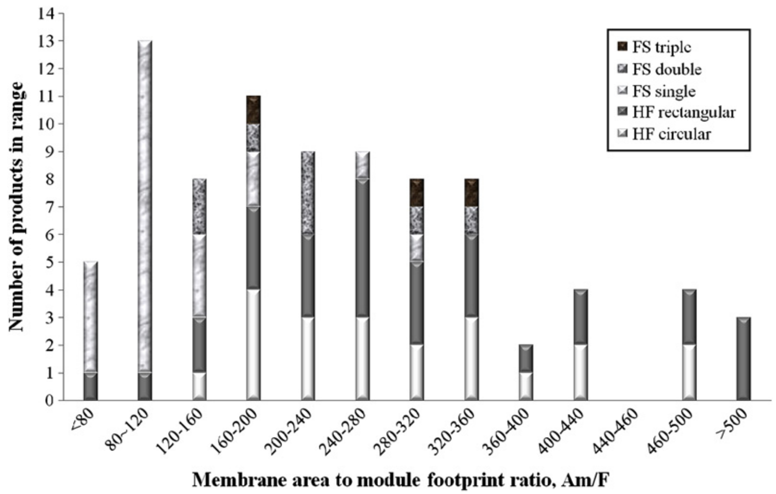

FIG. 4.61 Number of products against ranges of A_m:F ratio values (adapted from Santos, & Judd, 2010).

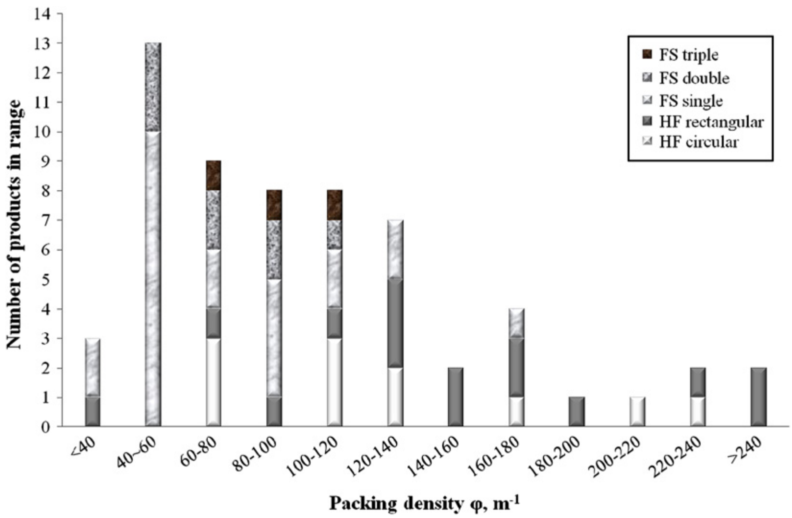

FIG. 4.62 Number of products against ranges of φ values (adapted from Santos, et al., 2010).

All but one of the FS module products are provided as either single- or double-deck, with only one exception where triple-deck stacking is possible. HF systems are normally single deck, though there are three products which are stackable. Figures 4.61 and 4.62 show a wide variation in both A_m:F ratio and φ values across the range of products, with higher values for the HF products. Most FS panels are 4–7 mm in thickness (t), with the exception of some of the newer ultrathin modules, and generally separated by 6–9 mm (δ) in the stack. This means that the maximum possible packing density – the panel packing density φ_{panel} – for an FS stack is given by:

$$\varphi_{panel} = 2000/(t + \delta).$$

According to this correlation, and based on the specifications provided for the FS modules, φ_{panel} across the range of products is between 125 and 267 $m^2\,m^{-3}$, with over three-quarters in the 130–160 range. This compares with a much wider variation in φ overall, with no correlation between φ_{panel} and φ (Fig. 4.63). In the case of the HF products, where filament diameters vary between 0.3 and 2.8 mm across the entire range, there is again no evident correlation between packing density and fibre diameter (Fig. 4.64). A wide range of packing densities (from 40 to 250 $m^2\,m^{-3}$) arises from a very narrow range of fibre size (1.2–1.3 mm) representing almost half of the 25 products presented. Thus, for both the FS and HF products, variations in overall packing density arise mainly from module design and construction aspects, rather than from any spatial constraints imposed by the required average membrane interstitial distance.

It can be envisaged that there exists a range of optimum values of the iMBR module dimensions which are likely to relate to key design operation and maintenance (O&M) parameters such as specific aeration demand of the

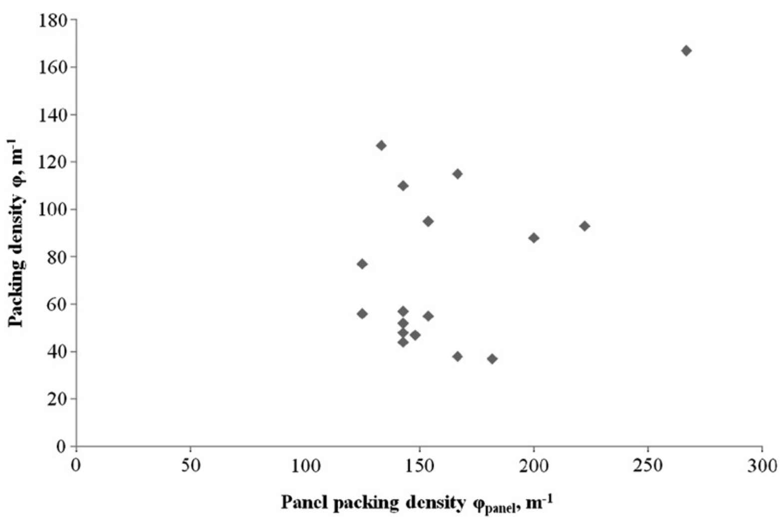

FIG. 4.63 φ vs φ_{panel}, FS products (adapted from Santos et al., 2010).

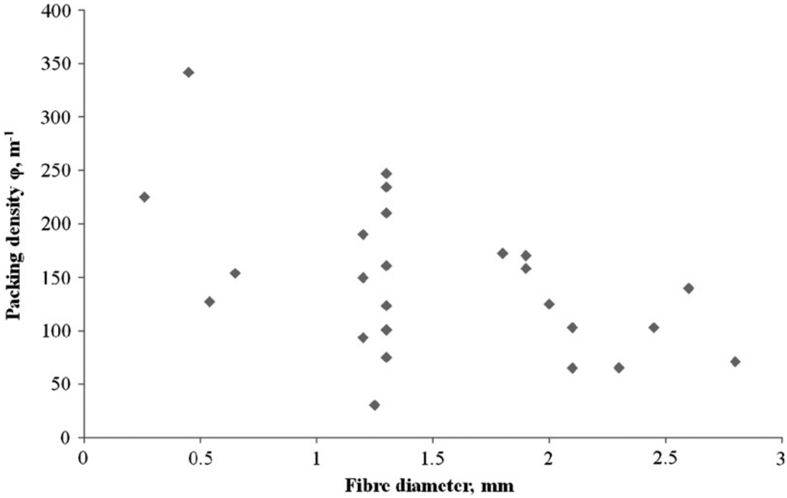

FIG. 4.64 φ vs fibre diameter, HF products (adapted from Santos, & Judd, 2010).

membrane and aeration protocol, gross flux, physical and chemical cleaning protocols, sludge retention time (and thus MLSS) and pre-treatment. Clearly, the maximum value of the key design parameter of membrane packing density is likely to be constrained by fouling and clogging propensity − as well as cleanability − of the membrane. Given the absence of any immutable fundamental correlations between packing density and the key O&M parameters, it is perhaps apposite to review information provided by case studies and practical experience (Chapter 5) to shed light on O&M parameters for the different configurations. These may then provide the basis for MBR design and operation (Chapter 3).

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Chapter 5

Case Studies

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**PURON® is a registered trademark for the Koch submerged membrane module in Germany and other countries.*

5.1. INTRODUCTION

MBR plant operation is largely characterized by hydraulic and purification performance. Purification is normally stated with respect to biochemical oxygen demand (BOD) and/or chemical oxygen demand (COD), total suspended solids (TSS), ammonia (NH_4^+-N), nutrients (nitrogen and phosphorus) and micro-organisms, though the discharge consents may not necessarily specify all of these. Key nutrient-based water quality determinants are the ammonia, total Kjeldahl nitrogen (TKN), total nitrogen (TN), total organic/inorganic nitrogen (TON/TIN) and total phosphorus (TP).

Hydraulic characteristics centre mainly on the flux in L/(m^2 h) (LMH), physical and chemical cleaning cycle times, downtime associated with cleaning and overall conversion. Hydraulic performance may change with flow in megalitres per day (MLD), and hence the difference between the average daily flow (ADF) and peak daily flow (PDF) is important. In the case of immersed systems membrane aeration demand is critical, quantified as the specific aeration demand with respect to the membrane area (SAD_m in $Nm^3/(m^2\ h)$) and permeate production (SAD_p in Nm^3/m^3). For pumped sidestream systems the liquid crossflow velocity (CFV) and permeability – the ratio of the flux to the transmembrane pressure (TMP) – are critical. Cleaning cycle times are normally dictated by the requirement to sustain a reasonable mean permeability for the system, and the absolute permeability value appropriate to an MBR treatment process is dependent on the technology – and more specifically the membrane and process configuration (Chapter 4). Similarly, the membrane aeration demand also varies between technologies, as well as with feedwater characteristics.

In the following sections, over 50 case studies are provided, based on information provided across a wide range of membrane configurations (flat sheet (FS), hollow fibre (HF) and multitube (MT)) and suppliers, consultant/contractors and end users. Information-contributors are listed in the chapter title page. As with Chapter 4, the subject matter is divided according to membrane configuration and supplier. The data are summarized in Section 3.2.2, as a precursor to the following sections on design, operation and maintenance.

5.2. IMMERSED FLAT SHEET (FS) TECHNOLOGIES

5.2.1. Kubota

5.2.1.1. Porlock

Client: Wessex Water
Main contractor: WW MBR
Main consultant: Europumps

Background

Porlock is the site of the first full-scale MBR plant to be installed in the United Kingdom and, indeed, the first of any significance worldwide. Interest in MBRs for municipal wastewater treatment in the United Kingdom arose directly as a result of EU legislation on the quality of treated sewage effluent being discharged to recreational waters, and specifically the Bathing Water Directive. This directive, originally promulgated in 1976 and revised in 2002, stipulated that such waters should meet stringent microbiological guide values of 500/100 mL total coliforms and 100/100 mL faecal coliforms and faecal streptococci. The first membrane plant to be installed for sewage treatment was actually the groundbreaking abiotic plant at Aberporth, a Welsh Water site, in 1994. This plant employs enhanced upward flow clarification with lamellar plates (the *Densadeg* process), followed by polishing with Memcor HF microfiltration (MF) membranes, with upstream and intermediate screening by a 0.5 mm perforated plate (Fig. 5.1). This process has since been superseded by the MBR process whose widespread installation in coastal regions around the United Kingdom followed successful pilot trials of the Kubota membrane bioreactor process in Kingston Seymour in the mid-1990s. The trials subsequently led to the installation of plants at Porlock in February 1998, and the larger plant at Swanage in 1999.

The trials conducted at Kingston Seymour, a Wessex Water site, arose through a collaboration between the UK water utilities Welsh Water, South West Water and Wessex Water. It was largely through these trials that Wessex Water was able to gain considerable know-how for this process. The pilot plant was operated for two years prior to the installation of the full-scale plant at Porlock, and continued for several years thereafter.

At the time at which the pilot trials were taking place, a second pilot trial based at Newton Aycliffe in Northumberland was being conducted on a membrane system developed by the company Renovexx. This latter system was based on a dynamic membrane of aluminium flocculant material formed on an inert woven-cloth porous substrate from washout from a primary tank dosed with alum at 30 ppm as $Al_2(SO_4)_3$. There was thus downtime associated with

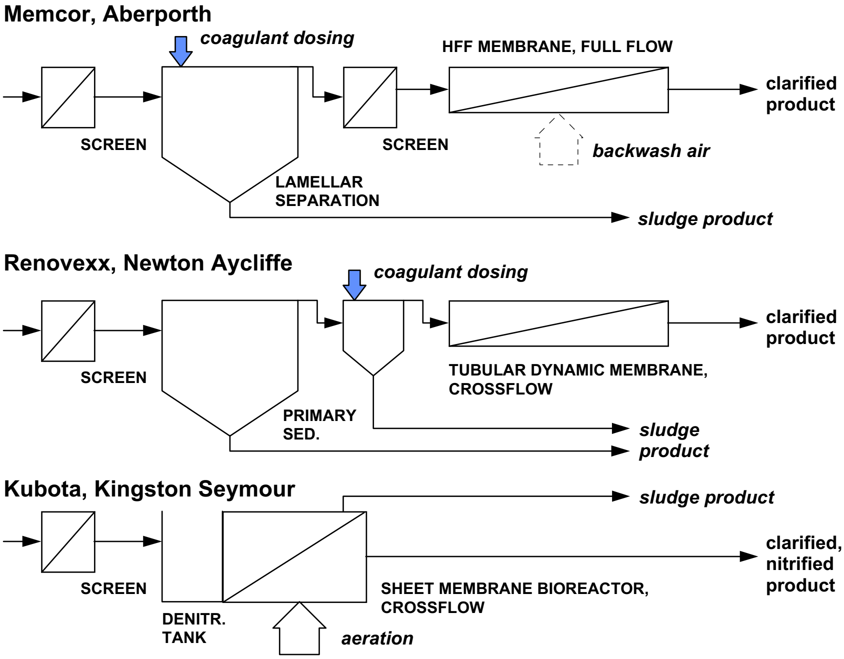

FIG. 5.1 Schematics of the plants at Aberporth, Newton Aycliffe and Kingston Seymour.

the formation and removal of the dynamic membrane, though the actual cost of the woven substrate was relatively low. Because this was a crossflow filtration process, the membrane path length was necessarily very long – about 22.5 m – to obtain a reasonable conversion of 50%. The substrate module (a 'curtain', of which there were 22 in total) comprised a 30 × 25 mm diameter multitube (MT) of a single piece of material, around 1 m high. Whilst low in cost with respect to the membrane itself, the engineering needed to operate the plant successfully was complicated and made the technology overly expensive for all practical purposes.

Data provided at the time (Judd, 1997) for the three plants, the Kubota and Renovexx pilot plants and the full-scale Memcor plant at Aberporth, are given in Table 5.1, and the schematics in Fig. 5.1. It was evident from the trials that, whilst operating at a much lower flux, the MBR technology offered a number of advantages over the other two membrane technologies in that it:

- achieved denitrification,
- achieved significant dissolved organic carbon removal,
- operated on screened sewage, with no further pre-treatment demanded,
- operated at low pressure and
- demanded minimal mechanical cleaning.

TABLE 5.1 Comparison of Performance of the Plants at Aberporth, Newton Aycliffe and Kingston Seymour (Adapted from Judd, 1997)

Parameter	Kubota	Renovexx	Memcor
Location	Kingston Seymour	Newton Aycliffe	Aberporth
Function	TSS, COD/BOD removal and nitrification/denitrification	TSS/BOD, precipitated and coagulated solids removal	TSS/BOD removal and disinfection
Membrane material	Hydrophilicized polyethylene (PE)	Dynamic Al floc on substrate	Polypropylene (PP)
Membrane configuration	Immersed FS	Crossflow MF MT	Full flow HF
Membrane pore size, μm			
Nominal	0.4	10 (for substrate)	0.2–0.3
Operating	<0.1	Nomin. <1 μm	0.2–0.3
Feedwater specification	Screened raw sewage	Screened, settled sewage	Coagulated, settled sewage
TMP, bar	0.1	2 Inlet, 1 bar outlet	0.2
Flux, LMH	22	140 Capacity	87 Capacity

Production rate, MLD	0.13	4 Capacity	3.75 Capacity
Membrane module	0.8 m^2 panel	25 mm dia., 22.5 m × 30 multitube, 53 m^2	0.55 mm dia., 10 m^2
Total number of modules	300	22	180
Opern./regen. cycle, min	Continuous	180/25*	45/2
Cleaning			
Mechanical	Water jet, 1/yr	Brush/water jet	Compressed air
Chemical	2/yr		Every 3–4 d
Tot. membrane area, m^2	240	1166	2700 (duty/stdby)*
Membrane life, years	~7	1.5–2 (w. brush cleaning)	~5
Membrane cost, £/m^2	80**	30–40	35
Capital cost, £m/MLD plant	0.6–0.8/1	0.5/4	nd/3.75

**2:1 Duty:standby, hence 1800 m^2 duty. nd, Not disclosed.*
***Projected cost: the cost of the Kubota membrane at the start of the 1995 Kingston Seymour trial was >£150/m^2 including housing. The cost in 2005 for a large installation would be likely to be <£40/m^2.*

Whilst the projected capital costs of the MBR plant were higher because of the lower flux, this was more than offset by the additional capability offered by the MBR. It is noteworthy that whilst Aberporth remains the only UK site where microfiltration of sewage is employed abiotically, there were at least 44 MBRs for sewage treatment operating in the United Kingdom by June 2005 and around double that number by the end of 2009, most of these being based on Kubota technology.

Plant Design and Operation

The stipulations regarding the Porlock plant were that it should be a small-footprint plant, and able to disinfect the sewage whilst imposing little visible impact. The plant therefore had to be housed in a stone-faced building identical in appearance to an adjacent farmhouse (Fig. 5.2). It comprises four aeration tanks (Fig. 5.3), 89 m^3 in liquid volume size and having dimensions 3.3 m wide × 7.4 m long × 4.5 m deep. The actual liquid depth is ~3.8 m, the dimension allowing for any possible expansion to the seven modules per tank. The tank volume allows for the volume displaced by the membrane units, about 0.6 m^3 per unit. Each tank contains six membrane units with 150 panels per unit, giving a total membrane panel area of 2880 m^2 provided by 3600 membrane panels at 0.8 m^2 membrane area per panel. Pre-treatment comprises a 3-mm perforated screen, this having replaced the original 2 mm wedge wire screen during the first year of operation.

At a peak flow of 1.9 MLD, the plant operates at a flux of 27 LMH. However, the average flux is generally below 20 LMH and the TMP usually between 0.02 and 0.11 bar under gravity-feed conditions (governed by the tank depth and pipework hydraulic losses). This means that the permeability is normally between 200 and 500 LMH per bar, but can increase when the fouling and clogging propensity is low. The membrane module is aerated at an SAD_m rate of 0.75 $Nm^3\,h^{-1}$ per m^2 membrane area (the standard coarse bubble aeration rate for the Kubota system), which means that SAD_p, the Nm^3 volume of air of m^3 permeate product, is 32. The membrane is relaxed for 30–60 min daily as a result of low overnight flows, and cleaning in place with 0.5 wt% hypochlorite is undertaken every eight to nine months. Diffusers are flushed manually without air scour. The plant operates at SRTs of 30–60 days, producing 0.38–0.5 kg sludge per kg of BOD, and 3–6.5 m^3/day of 2% dry

FIG. 5.2 The MBR plant at Porlock: (a) profile and (b) plan.

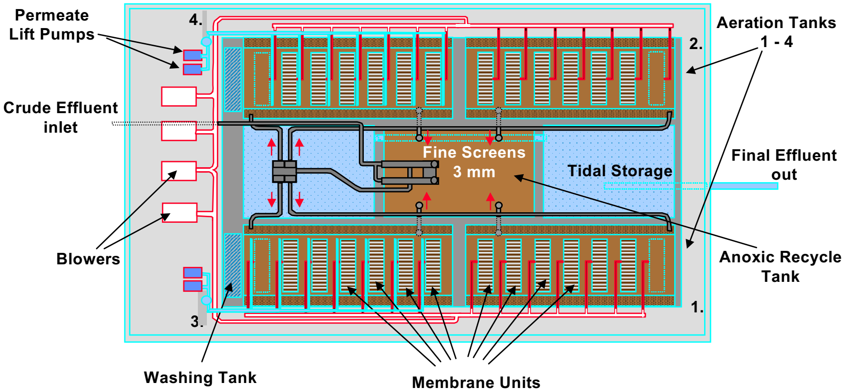

FIG. 5.3 A schematic of the MBR plant at Porlock.

solids sludge. The recommended mixed liquor suspended solids (MLSS) concentration range is 12–18 g/L, but it has been operated as high as 30 g/L and is regularly over 20 g/L.

The plant is of a simple and robust, possibly conservative, design and construction and has generally proved very reliable. Membrane replacement due to actual membrane failure for this plant during the first eight years of operation was minimal (Kennedy & Churchouse, 2005), with the membranes only having been substantially replenished after 12 years; the plant was still operating with ~40% of its original panels in March 2010. A consideration of all UK and US membrane plants based on this technology reveals membrane replacement per annum (p.a.) generally to be very low (Table 5.2) – generally less than 1% p.a. and always <3%. Permeate water quality has always been well within consent, and the plant produces very low odour levels (<2 mg m^{-3} H_2S). Porlock was originally costed on the basis of a membrane life of seven years, a milestone that has already been passed. Periodic fouling events have been encountered at this plant, and some of these have been linked with seawater intrusion. Shock loads of salinity are known to cause increases in soluble COD levels in the mixed liquor, and this has been shown to increase fouling propensity (Reid, Liu, & Judd, 2006).

5.2.1.2. Swanage

Client: Wessex Water
Main contractor: TJ Brent and WABAG
Main consultant: Wessex MBR

The MBR option was chosen for Swanage primarily due to footprint limitation. Disinfection of the effluent was required to meet the EU Bathing Water Directive (BWD). There was some confidence in the technology derived from

TABLE 5.2 Total Number of Membrane Failures by Age of Original Panel (Kennedy & Churchouse, 2005)

Within	No. of Installed Panels of Age	Panel Failures in Year	% p.a.	Cumulative %
Year 1 (i.e. 0–1)	141,061	192	0.14	0.14
Year 2 (i.e. 1–2)	115,985	598	0.52	0.65
Year 3 (i.e. 2–3)	89,714	690	0.77	1.42
Year 4 (i.e. 3–4)	74,606	1388	1.86	3.28
Year 5 (i.e. 4–5)	36,247	33	0.09	3.37
Year 6 (i.e. 5–6)	32,772	95	0.29	3.66
Year 7 (i.e. 6–7)	14,938	20	0.13	3.80
Year 8 (i.e. 7–8)	4200	0*	0	3.80
Year 9/10 (i.e. 8–10)	386			

Note: data as at June 2005, 45 operational plants (including *all* 40 UK & Ireland, both municipal and industrial). Data exclude ~300 panels repaired by client following accidental damage.
**No failures reported in this year as at June 2005.*

extensive testing at Kingston Seymour and experience gained operating the Porlock plant. Operation started in 1999 and, at a peak daily flow of 12.7 MLD, the plant was the largest MBR installation in the world at the time.

The plant is one of the least visible large-scale sewage treatment works. It has been completely landscaped into the Dorset coastline (Fig. 5.4), a considerable feat of civil engineering incurring a commensurately high cost. Treatment comprises grit removal, fine screening down to 2 mm using a rotating band screen, followed by the MBR, which has six aeration tanks of 3.3 × 22.5 × 5 m dimensions. It was originally fitted with 132 single-deck units (22 per tank) with 150 panels per unit providing a total membrane area of 15,840 m^2 from 19,800 panels, and operated under similar aeration conditions and MLSS levels as that of Porlock (0.75 $Nm^3/h\,m^2$ SAD_m and 12,000–18,000 mg/L MLSS).

The original design flux of the plant at Swanage was 34 LMH. However, it rarely reached that capacity in practice and was beset with a number of problems:

Clogging of membrane channels and aerators. The screens (2 mm rotating band screen) did not prevent hairs from entering the plant and causing problems. As stated in Section 3.6.2, these hairs, if discrete and free flowing, are not onerous to the process. However, they have a strong tendency to agglomerate (or reconstitute) in the membrane tanks along with cellulosic materials to form

FIG. 5.4 View of Swanage sewage treatment works: (a) from the sea and (b) from the air.

long 'braids'. These formed 'mats' at the entrances to the channels and, ultimately of even greater concern, on the aerators. The aerators are routinely flushed through to remove any sludge collecting inside them (Section 3.6.4). However, this does not necessarily remove braids wrapped around the aerators.

General sludging and fouling. Although the plant typically operated at fluxes between 15 and 20 LMH and TMPs of 0.05–0.12 bar, it was designed for a flux of 0.8 m/day PDF – substantially higher than the design flux of 0.6 m/day (25 LMH) for Porlock. The plants are similar in design concept, with multiple aeration tanks fed from a single distributor. The high imposed fluxes led to fouling and sludging between the panels. Repeated labour-intense external manual cleans were required to maintain the required flow capacity. This inevitably led to a significant proportion of panels becoming torn and damaged.

Wessex Water have since adapted the tanks to allow double-deck units to be fitted. The refurbished plant is fitted with 18 *EK* units of 300 panels in each tank, hence 5400 panels per tank compared with 3300 for the old single-deck system. Thus in the refurbished plant the number of panels has increased from 19,800 to 32,400 in total, with space for four more units per tank. The fitting of double-deck units necessitated raising the walls on one of the six tanks, the remainder having sufficient freeboard to allow 1.1 m of hydrostatic head after the modifications. The head is limited to maintain a constant pressure differential across the plant. This has had the effect of reducing flow through the plant, which could cause backing up in the central feed tank, though this does not appear to have caused a problem in practice. Since the tanks are not operated independently, fouling of membranes in one tank does not always necessitate cleaning, since the flow is simply re-directed to another tank. Provided the fouling is ephemeral and the permeability recovered, flows can be readjusted to recover the homogeneous distribution between the six tanks. Other modifications are the uprating of the blowers from 45 kW to 90 kW and the screening of the return activated sludge (RAS). The modified plant has been operating for two years without any undue problems.

5.2.1.3. Other Wessex Water Plants

The most recent Kubota MBR plants installed for sewage treatment in the United Kingdom have an automated diffuser maintenance programme, whereby they are periodically flushed with water and air-scoured. It is generally recognized by the operator that these plants require careful maintenance to suppress clogging (or sludging), since the filling of the channels with sludge represents a very significant constraint to the viable operation of these plant. To this end, more conservative peak fluxes of 27 LMH appear to be appropriate to suppress fouling, coupled with rigorously cleaned aerators to maintain the aeration rate, and thus air scour, in the membrane flow channels. The specific energy demand figures for the Wessex plants range from 1 kWh/m^3 to 2 kWh/m^3.

5.2.1.4. Daldowie

Client: Scottish Water
Contractor: SMW
Membrane designer: MBR Technology (now part of Ovivo)

The Daldowie Sludge Treatment Centre processes 50,000 tonnes of dried sewage sludge fuel per annum from the Greater Glasgow region with more than 1.5n population equivalent (p.e.). A public finance initiative (PFI) concession was awarded to SMW Ltd by Scottish Water to build the treatment centre and operate it for 25 years, with SMW Ltd paid per tonne of dry fuel produced. The facility was commissioned in December 2001 and was fully operational in Autumn 2002.

The facility receives sludge from a variety of sources in the Greater Glasgow region. Most of the sludge is co-settled and the average sludge volume received at the centre is typically 6500 m^3/day at between 1.7% and 2.3% dry solids (DS). Normal sludge reception is to two reception tanks and two buffer tanks, providing 25,000 m^3 of normal storage on site. This is supplemented by further emergency storage tanks giving a total on-site storage of 36,000 m^3/day. Typically there are three to four days' storage on site. Prior to entering the sludge process building, the sludge is pre-screened to 5 mm to remove any gross debris.

The sludge processing centre consists of six process lines each comprising duty/standby centrifuges followed by a drum dryer. The centrifuges dry the solids to 26% DS, with 90–92% DS achieved after the dryers. On average the centre tankers away between 130 and 150 tonnes of dried sludge a day. The sludge has a gross calorific value of 16,600 kJ/kg. The centrate and the evaporate liquors from the sludge processing centre pass through lamellar settling tanks, installed to provide protection should solids breakthrough occur in the liquors from the sludge processing centre, which incorporate 3 mm static perforated plate screens. This then passes to the MBR plant prior to discharge into the river Clyde, the discharge consent being 20:75:12 BOD:TSS:NH_3-N.

Sludge liquor is the aqueous fraction of sewage sludge which has undergone dewatering by processes such as belt pressing, rotary drum vacuum filtration and centrifugation following conditioning with coagulant and/or polymeric flocculant reagents. It may contain up to 25% of the total nitrogen load in the original sludge and contribute as little as 2% of the total influent flow. It is thus highly concentrated in ammonia, as well as in dissolved organic matter. Composition and flow are extremely variable and dependent on the upstream sludge handling and treatment processes. Whilst the most economical option is to return the liquor to the head of works, this is not always possible either due to logistical limitations or due to the excessive load it would place on the existing sewage treatment process. In the case of Daldowie, a number of options were available for consideration at the time (Table 5.3).

As is often the case for this duty, the choice of technology at Daldowie was seen as being between a conventional activated sludge process (CASP) and an MBR. However, the small footprint of the MBR, together with its ability to

TABLE 5.3 Summary of Treatment Options (Modified from Jeavons, Stokes, Upton, & Bingley, 1998)

Method	Advantages	Disadvantages
Magnesium ammonium phosphate precipitation	Low capital costs Total nitrogen removal Instantaneous start-up	High operating costs Difficult to control Centrifuge required High sludge production No large-scale experience
Ammonia stripping	Total nitrogen removal Limited experience of similar technology Instantaneous start-up	Disposal of high ammonia liquid waste No large-scale experience in UK water industry
Biological	Tried and tested technology Simple to control No problem wastes Surplus sludge boosts nitrifier population in ASP	Nitrates returned to ASP Start-up not instantaneous

accept the large fluctuations in organic loading arising from periodical releases of poor quality effluent from the centrifuges, meant that the MBR was selected. The buffering capacity offered by the MBR means that it can operate at an MLSS of 12−18 g/L rather than the 5 g/L of a CASP or sequencing batch reactor (SBR), and tolerate significant increases in organic and hydraulic loading for up to two to three days.

The main effluent streams are centrifuge centrate and dryer condensate, although other site liquors are also treated. The feedwater after the lamellas typically has a mean composition of 200 mgN/L as NH_3-N and 1500 mg/L as TSS. Mean feed COD levels are in the range of 2500−4000 mg/L. The effluent from the MBR plant typically has BOD $<$ 3 mg/L, TSS $<$ 6 mg/L and NH_3-N $<$ 3 mg/L, representing removal efficiencies greater than 95%. The maximum design flow is 12.8 MLD (yielding a minimum HRT of 15 h), obtained with six dryers running without a standby.

Flows pass through lamella separators and a 3-mm screen into a central denitrification/recycle tank. From there, flows are distributed into four combined membrane and aeration tanks, each of 2000 m^3 volume (Figs 5.5 and 5.6), and a sludge recycle stream returned from the tanks to the denitrification section. In each of the aeration tanks there are 1710 diffusers and 32 $\times$ 200-panel (*J200*) membrane units, giving a combined 25,600 panels in the plant (20,480 m^2). The flows thus equate to a mean design flux of 18 LMH and a maximum of 22 LMH.

A great deal of time and effort has been spent identifying the most appropriate polymer and its dose for chemical conditioning in the centrifuges without detriment to membrane permeability in the downstream MBR. Despite the problems associated with optimizing the polymer dosing, the plant has maintained good performance in terms of COD and ammonia removal. Moreover,

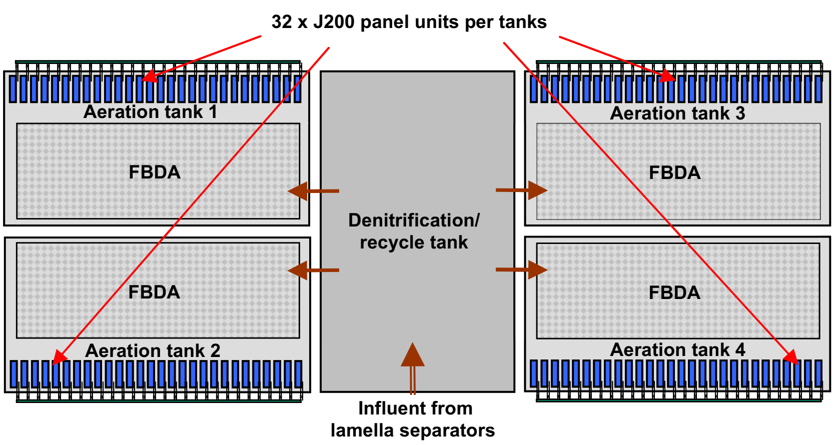

FIG. 5.5 Schematic layout of the Daldowie sludge liquor treatment plant.

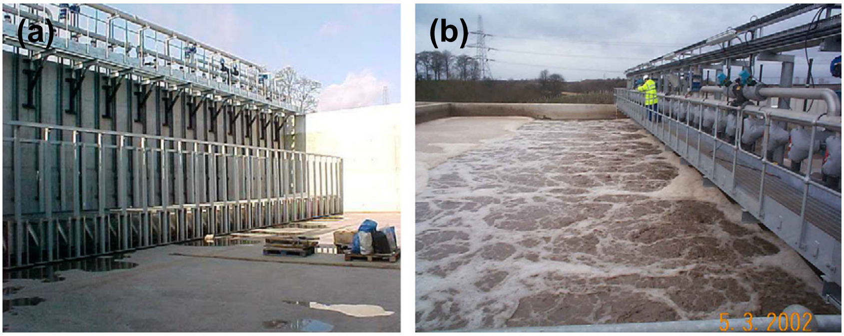

FIG. 5.6 The Daldowie sludge liquor treatment plant: (a) under construction and (b) operational. The photograph shows the plant with the membrane units and manifolding to the right.

the membranes have proven very robust; no more than 40 of the 25,600 panels were replaced in the first three years of operation.

In 2006 routine cleans in place (CIPs) with hypochlorite and citric acid were being completed every four weeks or so. Furthermore, operational experience had highlighted that the plant was unable to treat the peak design flow while one tank was out of service. SMW contracted Eimco Water Technologies (now Ovivo) to review the design of the plant and increase the membrane surface area to enable the maximum design flow and load to be treated with one tank out of service. Following the design review it was decided to increase the membrane surface area by placing a second deck of membranes on 16 of the 32 membrane units in each tank. The upgraded plant now contains 48 $\times$ 200-panel membrane units, giving a combined 38,400 panels in the plant (30,720 m^2). The original design fluxes of 18 LMH and 22 LMH were maintained as design parameters for the upgraded plant with one tank out of service. In addition, a number of the original membranes were removed from the tanks and tested to determine if a cleaning protocol could be implemented to recover the performance of the original membranes. This identified that an external hypochlorite and acid clean would recover the performance of the membranes to 'as new' levels. These external cleans were completed at the same time as the installation of the additional membrane units. The time taken to clean the membranes and install the new units into each tank was less than three days. The performance of the restored original membranes was similar to that of the new panels, with CIPs completed typically once every six weeks.

One other unusual operational problem encountered during an early period of operation was an abrupt increase in differential pressures accompanied by sudden foaming. The foam on the top of the tanks was green-grey in colour, whereas the sludge remained brown. Samples of the foam revealed it to consist almost entirely of non-settling discrete chlorella algae of about 15 μm diameter. The algae were neutrally buoyant and non-flocculating and appeared to pass through the centrifuge as a result. The incident was attributed to the sludge plant processing waterworks sludge from a site where an algal bloom had been experienced. Whilst differential pressures had initially doubled, the foam gradually subsided over a number of days and differential pressure gradually recovered without intervention over a one to two week period.

5.2.1.5. Running Springs

Client: Running Springs Water District
Main contractor: Frost
Main consultant: Engineering Resources and Enviroquip (now Ovivo)

The 2.3 MLD ADF (4.5 MLD PDF) Water Recycling Plant at Running Springs was the first double-deck (*EK*) membrane module installation in the USA. The plant is located at Big Bear National Park in California and is operated and owned by the Running Springs Water District. The installation resulted from a change in the discharge permit granted to the existing activated sludge process plant, designed for BOD removal only, in which limits for nitrogen and phosphorus were tightened. The operators originally attempted to meet the new consents by extending the sludge age to increase the MLSS concentration in the aeration tanks. However, since the existing rectangular secondary clarifiers could not handle the increased loading other options had to be explored. Between April and September 2003, the existing plant was converted to an Enviroquip MBR system by retrofitting with Kubota membranes coupled with the *SymBio*® process, a proprietary technology designed to achieve simultaneous nitrification/denitrification (SNdN) and partial enhanced biological phosphorus removal in the same zone.

The plant (Fig. 5.7) is an example of Enviroquip's *UNR*™ (ultimate nutrient removal) treatment strategy which is SNdN combined with the MBR. The two membrane tanks, each containing 8 *EK300* units of 240 m^2 each (total of 3840),

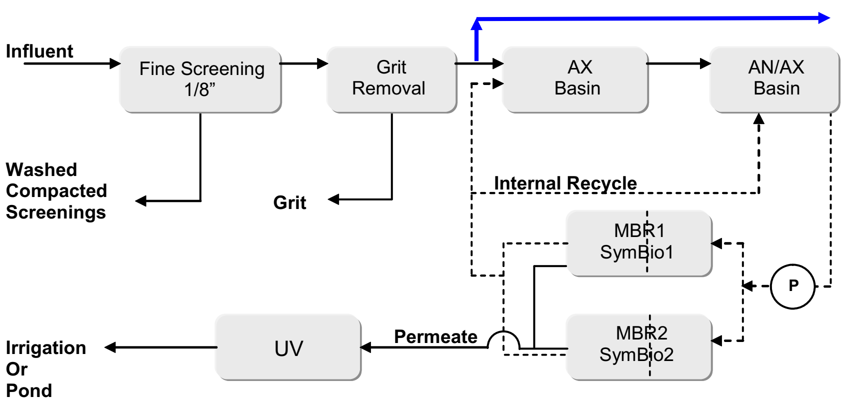

FIG. 5.7 Running Springs plant, schematic.

are placed downstream of sequential anoxic (AX) and anaerobic/anoxic (AN/AX) basins, converted from existing clarifiers. The anoxic treatment is preceded by screening, originally a 3 mm step screen which was subsequently replaced with a 3.2 mm centre flow band screen, followed by degritting in an aerated grit chamber. The aerobic part of the *UNR*™ process is operated at a low dissolved oxygen concentration using the *SymBio*® process to promote SNdN. The *SymBio*® zone is fitted with Sanitaire fine bubble aeration equipment and is fed air at a controlled rate based on the measured DO and biological potential activity (BPA) – a parameter based on the level of the energy transfer co-enzyme nicotinamide adenine dinucleotide (NADH). Submersible pumps continuously recycle mixed liquor to the aerated reactors, partitioned into MBRs and *SymBio*® zones, at roughly four times design flow. The retained sludge is returned to the anoxic basin for denitrification. Waste activated sludge (WAS) is pressed prior to disposal, and the filtrate returned to the head of the plant.

The membranes are aerated at a mean rate of 0.56 $Nm^3/(h\ m^2)$. The plant is typically set to relax for 1–2 min out of each 10–20 min filtration cycle, and cleaning with 8 g/L hypochlorite is typically twice yearly (2–6 p.a. since 2008 pending extended peak flow conditions outside of design criteria).

To accommodate inflow and infiltration during snowmelt conditions, an equalization (EQ) basin was added to the process. Whereas in the original design membrane performance was optimized by automatically matching permeate flow to hydraulic demand based on varying liquid levels in the AN/AX basin, control is now based on a larger, dedicated EQ basin. Equalized influent level can still be used to determine the mode of operation of the MBR, which can be anything from zero to peak flow. Using the real-time NADH and DO signals, the *SymBio*® process controller automatically adjusts the speed of a dedicated blower to keep the MLSS DO concentration between 0.2 and 0.8 mg/L to ensure that SNdN is optimized. From April 2003 to August 2003, the plant was operated based on DO measurements and conventional treatment strategy. Following optimization based on the *SymBio*® process during September 2003, effluent TN and TP levels decreased from 14.4 to 3.9 mg/L and from 7.4 mg/L to 4.9 mg/L, respectively.

Recent permeability testing confirmed clean water performance of the membranes to be within 95% of factory values (75 gfd/psi or 1850 LMH/bar). Stable, high membrane permeability coupled with periodic polymer (Perma-Care *MPE50*) dosing enables the plant to increase throughput, although winter flows still exceeded achievable capacity. Continued extended, cold weather (7–10 °C water temperature) events led the District to install the EQ basin and, in addition, eight *RW-300* units equipped with the new *B2-515* panels are to be added in 2010 to bring the total installed membrane area to 7320 m^2 and thus the peak design flux to 25.6 LMH. In anticipation of the plant upgrade, turbo fans have replaced lower output positive displacement blowers to increase air scouring capacity. According to plant staff, the new blowers draw roughly 50%

less energy than the old positive displacement machines and overall the energy efficiency of the plant continues to improve.

The total energy demand of the entire plant, including solids handling, heating, ventilation and air conditioning (HVAC) and other electrical loads such as chlorine generation, varies seasonally between 1.3 and 3 kWh/m^3. Assuming 40−50% of energy demand can be attributed to other plant loads around the plant, the MBR system consumption was roughly 0.7 kWh/m^3 in February 2010 when flows are at their highest. The trend is similar to that of other MBR systems with fixed membrane air flows, where normalized energy efficiency increases as flow increases during the winter months.

5.2.1.6. Dundee

Client: Village of Dundee, MI
Primary contractor: Barton Malow
Primary consultants: ARCADIS and Enviroquip (now Ovivo)

The MBR plant at Dundee (Fig. 5.8), MI in the USA was installed as a retrofit to an existing SBR plant requiring upgrading to achieve higher flows and meet more stringent seasonal consents (of 1.5 mg/L BOD and 0.5 mg/L ammonia during the summer months). Conversion of the SBR tanks to process tankage combined with the construction of four dedicated membrane tanks has allowed doubling of the hydraulic capacity to 5.7 MLD, along with improved and reliable effluent water quality (Table 5.4).

The feedwater is screened to 3 mm and degritted before entering a single anoxic zone (for alkalinity recovery) prior to flowing under gravity to the aerobic zone, which is fitted with fine bubble diffusers. The mixed liquor is then conveyed to a common channel that feeds four membrane tanks, each

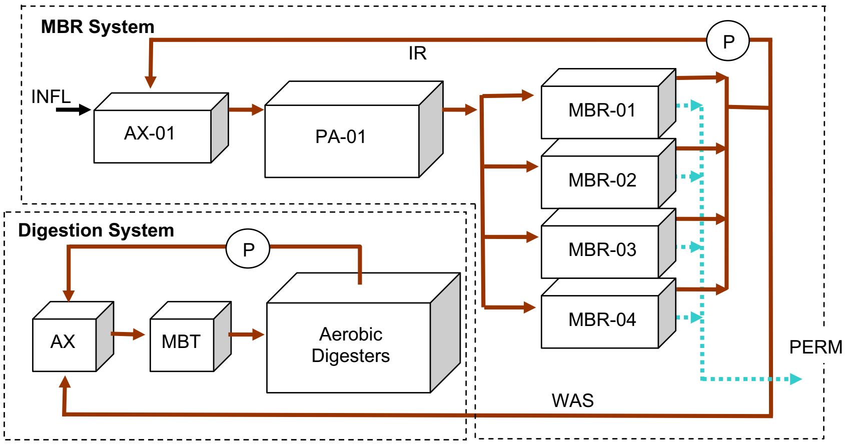

FIG. 5.8 A schematic of the plant at Dundee, MI.

TABLE 5.4 SBR vs MBR Effluent Quality Comparison, mg/L

Parameter	SBR	MBR (2006)	MBR (2007)
Carbonaceous BOD_5, $CBOD_5$	11.9	0.1–1.3, 1.0 on ave	<1.0
TSS	15.9	0.3–1.0, 0.7 on ave	<0.4
Ammonia nitrogen (as N), N-NH_4	6.0	0.7 on ave*	<0.2
Total nitrogen (an N)**, TN	–	1.0–8.0	NA
Total phosphorus (as P), TP	0.77	0.04–0.21, 0.13 ave	<0.12

**September excursion of 2.9 mg/L taken out due to aeration issues.*
***The plant is not optimized for TN removal and the concentration varies seasonally.*

equipped with 11 *EK-400* (14,080 m^2 membrane area double-deck) units with space for two more units. The sludge from the membrane zones flows into a common return channel where it is recycled (pumped) back to the anoxic zone at the head of the plant. Filtered effluent (permeate) is pumped into a baffled contact tank where alum, dosed at a mean concentration of around 40 mg/L at the average daily flow, is added for phosphorus removal. WAS is diverted from the common recycle channel into the digestion system where sufficient oxygen and contact time is provided to meet regulatory requirements before the treated solids are applied to land during biannual wasting events.

The retrofit, completed in 2005 at a total constructed cost of $6.55m, has resulted in a significant improvement in water quality, and solids handling costs have dropped by roughly 35%. However, the power costs almost doubled as a result of the upgrade, and have led the process supplier to investigate the benefits offered by varying the membrane aeration according to the hydraulic loading, so-called proportional aeration (Stone & Livingston, 2008).

Mean specific energy demand for the site has varied between 0.66 kWh/m^3 at peak flows of 5.67 MLD and 2.69 kWh/m^3 at 2.0 MLD. The overall average for the plant during January (2006, 2007), the peak flow period, was 0.95 kWh/m^3, of which <0.3 kWh/m^3 related to the MBR zone. Under average flow conditions, however, the contribution to the energy demand from the membrane blowers increases significantly (Fig. 5.9). The decrease in energy efficiency with decreasing flow reflects the impact of the aeration control limitations during turndown. It has been suggested that the aeration rate be adjusted by using three different setpoints for the blower, to supplement the setpoints for the liquid flow. This would allow operation of the plant at the 'ideal' energy demand experienced at peak flows. It has been suggested by the supplier that the use of proportional aeration in conjunction with improved distribution could bring the energy demand down below 0.84 kWh/m^3 without impacting

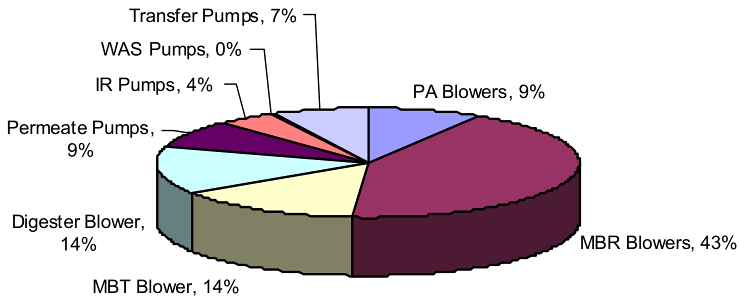

FIG. 5.9 Energy demand distribution at average dry weather flow at Dundee, MI.

performance which, based on historical costs, could translate into >\$50,000/yr (Stone & Livingston, 2008).

Notwithstanding the potential improvements offered by proportional aeration, the plant is one of the most energy efficient of its kind. It generally operates at discrete instantaneous net flux setpoints of 20 and 32 LMH. For nearly two years, the plant routinely operated between these fluxes at a sustained air scour intensity of approximately 0.27 $Nm^3/(m^2\ h)$, or roughly half of the typical value for the installed membrane equipment (*EK-400*), with only one clogging incident in one of the tanks. Maintenance cleaning was performed on average six times per annum for two of the membrane tanks and less than three times p.a. for the remaining two tanks, the difference relating to a single process upset affecting one of the tanks in May 2007 in which sludging took place within some of the units. This was recognized as being due to non-ideal hydraulics and was corrected through improvement of the flow splitting between the four tanks. Cleaning has been through soaking in >5000 mg/L hypochlorite for >2 h periodically followed by a 300 mg/L mineral acid soak. Recovery cleans have only been employed twice, one each on two of the tanks, requiring the units to be removed from the tank.

5.2.1.7. WwTP Sabadell-Riu Sec (Barcelona)

End User: Sabadell Municipal Council
Public Tender: Catalonian Water Agency (ACA)/Sabadell Municipal Council
Constructor Consortium: UTE Dragados-Drace-SAV-DAM
MBR supplier: HERA AMASA

The original facility at Sabadell was built at the beginning of the 1990s and initially planned for 30 MLD with primary treatment and conventional ASP. The plant required upgrading to achieve a 35 MLD capacity (200,000 p.e.) and a higher effluent quality with respect to nutrient removal. MBR technology was selected primarily due to the space constraints of the site, which is bounded by a main road, the local airport and a protected historical building, and the fact that construction of the new plant could take place without requiring shutdown of the existing process. The FS configuration was selected due to concern over clogging effects by textile fibres from received industrial effluents from local industries, the low maintenance requirement and operational simplicity, and the experience of the MBR supplier. HERA AMASA have provided the engineering, design, supply, installation and commissioning of Kubota-based MBR plant since 2002, with a number of reference plants in Spain prior to the Sabadell project. Installations include the first Spanish municipal MBR plant – Riells I Viabrea WwTP for Catalonian Water Agency ACA in 2003 with an average daily flow (ADF) of 2.4 MLD. The company has installed 12 Kubota WwTPs for municipal wastewater and golf resort applications, the overall capacity of these being 16.8 MLD.

The work, which began in 2007, required conversion of one of the two existing biological reactors to an MBR and the construction of a second new MBR of the same design (Fig. 5.10). Two additional buildings were needed to house peripheral equipment such as blowers, pumps and control instrumentation, along with crane bridges fitted with the reactors to allow the removal of the membrane modules for external cleaning or replacement, if required.

Following primary treatment using the existing coarse screen and sedimentation tanks the clarified water is passed to two MBR reactors which are

FIG. 5.10 A plan view of the plant at Sabadell under construction; the boundary with the historical building is at the top of the picture.

protected by 1 mm screens. Each reactor comprises four lanes, each fitted with an anoxic, an aerobic and a membrane chamber. The membrane chambers each contain 2×12 *EK400* units, and thus 192 double-deck units in total (61,440 m^2 total membrane area). The plant is equipped with 16 permeate suction pumps, rated at 170 m^3/h at 7.5 m hydrostatic head of pressure, controlled by 16 frequency converters. Permeation is provided under negative pressure on a cycle of 9:1 min operation:relaxation. A CIP is actuated when the pressure increases to −0.2 bar, normally over a period of around four months, by backflushing with 5 g/L sodium hypochlorite solution which incurs a downtime of 2 h; the CIP requires no tank draining or module removal. Membrane aeration by coarse bubble diffusers is at a SAD_m of 0.53 $Nm^3/(m^2\,h)$.

At the peak flow rate of 2620 m^3/h the flux incurred is 43 LMH, compared to that of 23.7 LMH associated with the design flow of 35 MLD. An average HRT of 9.1 h is provided by the 13,086 m^3 capacity tank, of which 4368 m^3 is occupied by the membranes, 4864 m^3 provides aerobic treatment and the remainder is the anoxic zone. The MLSS concentration is around 10 and 12 g/L in the aerobic and membrane tanks, respectively. Discharged WAS is dewatered on site by centrifuges. Sludge recirculation is via four submerged pumps, each with a capacity of 1100 m^3/h (and thus a recirculation ratio of ~3 at 35 MLD flow).

At June 2010 the MBR was undergoing commissioning with six of the eight lines in operation at a mean flux of 25 LMH, slightly higher than their design flux, at an overall specific energy demand of 0.8−1.0 kWh/m^3, this latter figure being a conservative estimate. The plant has been subject to a few foaming events, the cause of which has not been determined; however, the permeability has generally been above 125 and is often at 500 LMH/bar. The permeate quality is reliably high, with mean BOD_5, N-NH_4 and TN concentrations of 5, 1 and 22 mg/L, respectively.

5.2.1.8. *Illovo Sugar*

Client: Illovo Sugar
Primary contractor: Aqua Engineering, Johannesburg, South Africa
Primary consultant: Aquator South Africa

The plant at Sezela, Durban in South Africa treats 25% of the chemical process effluent arising from a sugar industry by-products plant, and is situated alongside a conventional ASP, which treats up to 2 MLD of mixed sewage/drain effluent from the adjacent Sezela sugar mill. The conventional plant provides a key function of generating WAS which is used to seed the MBR; this is essential since the MBR operates for only eight to nine months of the year. The MBR plant itself has a hydraulic design capacity of 1.2 MLD but in practice achieves no more than 1 MLD due to limited fine bubble diffuser aeration (FBDA) availability. This means that loading is limited to 18 tonnes of COD per day, or 3.8 kg/(m^3 day) compared with a design of 4.5. It is intended to increase the oxygen transfer by increasing the size of the fine bubble diffuser heads by 20%.

Selection of MBR technology was due to the presence of a trace toxin in the process effluent which inhibits conventional aerobic and anaerobic treatments. The high MLSS offered by MBR overcomes this limitation. The MBR plant was conceived, designed and installed at a time when electricity prices in South Africa were artificially low. It comprises a single 29 m diameter, 7 m deep and 4300 m^3 capacity cylindrical tank (Fig. 5.11) fitted with 12 *EK400* Kubota membrane modules providing a total membrane area of 3840 m^2 from 4800 panels. Membrane scouring is provided from a 61.5-kW blower rated at 2880 Nm^3/h at 500 mbar, yielding an SAD_m of 0.75 Nm^3/h per m^2. This compares with the considerably greater aeration rate demanded for process air provided by two 224 kW blowers rated at 7060 Nm^3/h at 740 mbar. The unusually deep tank was selected to reduce its footprint, and demanded that the membranes were mounted 2 m from the tank base.

The MBR is challenged with a feed of 17.5−18 g/L COD at a pH of 2.5 and with very low TSS, thus demanding minimal screening (0.5 mm wedge wire). The COD is primarily made up of readily assimilable organic matter in the form of ∼1% acetic acid, ∼0.1% formic acid and intermittent furfural concentrations of up to 100 mg/L (Kennedy & Young, 2007). The flow of feedwater into the tank is controlled by monitoring the pH of the mixed liquor, which is maintained above a value of 6 by impeding the flow to the plant when the value decreases below this set point. Whilst the capacity of the plant is limited microbiologically, it has been demonstrated that fluxes in excess of 25 LMH are readily achievable at this plant at the operating MLSS concentration between 12 and 18 g/L (attained at SRT values of 30−60 days). The long sludge ages provide correspondingly very low sludge yields as low as 0.06 kg/(m^3 d).

FIG. 5.11 The Illovo Sugar plant: (a) the biotank and (b) connections from the upper and lower decks of the membrane units in the biotank.

The plant operates at unusually high temperatures of around 52 °C, produced by the feed temperature of 36–38 °C coupled with the exothermic reactor conditions. The MBR was not originally designed for operation under such thermolytic conditions and, indeed, the membrane panels themselves are only warranteed up to an operating temperature of 40 °C. However, the plant has been operating since 2004 and, in the first six years of operation, required no more than nine replacement panels. These panels, all from the top deck, suffered weld failures. Other minimal damage to the membranes, and specifically surface scratching, has not demanded replacement due to the 'self-sealing' properties of the membrane (i.e. the reformation of the gel layer). In all, membrane replacement from the first six years of operation has been at around 1.5%.

The plant operates on the sugar mill effluent for only eight to nine months of the year determined by harvesting of the sugar cane, demanding start-up of the plant during the month of April. The elevated temperature of the effluent requires a very specific start-up protocol, since out of the sugar processing season the plant is shut down and needs reseeding. The plant is seeded with municipal sludge from the conventional ASP, accumulating to a solids concentration of around 8 g/L. The reactor is then fed with the by-products process effluent and the temperature gradually increased from the feedwater temperature to its steady operating temperature of 52 °C, during which the conditions change from mesophilic to thermophilic. This operation is complicated by the presence of an as yet unidentified toxin which suppresses bioactivity at the lower solids concentrations prevailing during start-up. However, the operators have found that daily dosing of the tank with around 50–100 m^3 of WAS from the CASP (screened to 0.5 mm) is sufficient to suppress the action of the toxin, presumably through adsorption, and also assists the transition from mesophilic to thermophilic conditions at around 50 °C. It appears to be the combination of the high operating temperature, WAS dosing and the high MLSS concentration which maintain both the required bioactivity – ameliorating toxic shocks – and membrane permeability; there is a distinct deterioration in permeability below an MLSS of 10 g/L. Once the MBR is up to operating temperature and MLSS, the addition of seed WAS is suspended other than if there is a clear indication of reduced bioactivity in the MBR.

The plant has operated very reliably with little or no manual intervention outside of start-up. It was operated without any chemical cleaning for the first year, when fed solely with the low TDS by-products process effluent. Following an in situ clean during the first annual shutdown, CIPs during the second and subsequent operating seasons were undertaken approximately every three to four months (i.e. 2–3 times every operating year). Additional chemical cleaning was required as a result of introducing an experimental process change, whereby a high proportion of the sugar mill's mixed drain/sewage effluent was treated to relieve the overloaded CASP. This then demanded a CIP with dilute HCl every two to three months, the Mill effluent having a very high scaling tendency. Chemical cleaning comprises hydrochloric acid at 2–4 wt% concentration, employed primarily for struvite removal followed by sodium hypochlorite.

The plant has served to demonstrate the feasibility of the MBR technology to the end user, both with reference to effluent treatability and plant operability. The feed is highly biodegradable with negligible TSS, and COD removal is 95–96% resulting in a permeate COD of 700–800 mg/L. The permeate is, however, discoloured due to the long sludge age and high operating temperature. Cane wax, a natural polymer from sugar cane present at concentrations of around 150 mg/L, does not appear to foul the membrane. However, soaring South African electricity prices, having doubled since 2007 and set to double again by 2012, mean that the cost of aerobic processing on this scale (over 70 tonnes/day of COD) is prohibitive, and the full-scale plant for treating the whole flow is likely to include anaerobic treatment. Toxin-resistant anaerobic technologies are being researched at the time of writing. Moreover, because the region is subject to periodic droughts, water recycling of the treated effluent is also under consideration. This will probably involve upgrading the existing MBR with additional membranes to increase its hydraulic capacity and using it as a polishing step downstream of an upstream anaerobic process.

5.2.1.9. Al Ansab

Client: Oman Wastewater Services (Haya Water)
Primary contractor: Galfar Engineering and Contracting
Primary consultants: Khatib and Alami, Metcalf and Eddy

The plant at Al Ansab in Oman is the largest MBR plant in the world, with a current ADF of 55 MLD (77 MLD PDF), with capacity for future expansion to 84 MLD. The plant is built on the site of the existing 12 MLD plant and can house up to 304 double-deck *EK* stacks, offering a membrane area of 97,280 m^2 from the 121,600 panels. The plant is currently fed with sewage which is tankered in. This is fed through a grease trap and degritter prior to screening with 3 mm drum screens before entering the biotank. The plant has been fitted with 240 *EK* units across eight 1080 m^3 tanks, each double-deck unit containing 400 panels (320 m^2), and thus 76,800 m^2 overall. At the ADF this corresponds to a flux of 0.72 (30 LMH), afforded by the consistently high wastewater temperatures. Relaxation takes place for 5 min in every hour. Scouring air can be adjusted according to the membrane permeability, with sufficient capacity to provide up to 1.1 Nm^3/h of air per m^2 membrane.

5.2.1.10. Anaerobic Kubota Systems: Original Developments

Kubota MBR technology has been applied to fermentation since the turn of the Millennium, following pioneering demonstration trials on municipal waste dating back to 1996. These installations have generally been either for

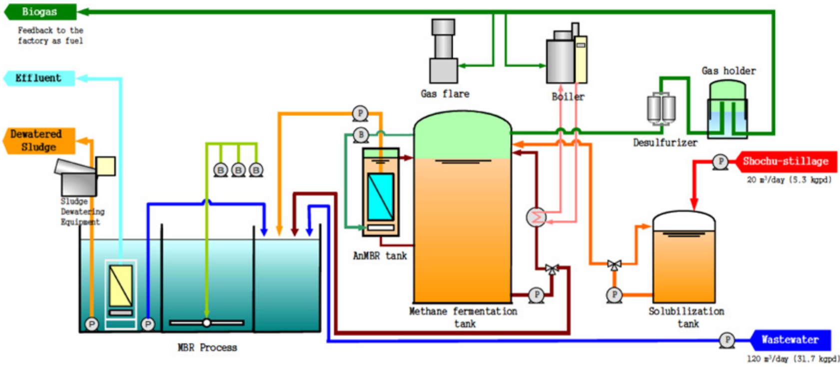

FIG. 5.12 A complete *shochu* combined wastewater and stillage treatment scheme.

municipal waste, wastewater sludge or industrial food and beverage waste. By the end of 2006 there were more than a dozen such systems in Japan alone with loadings ranging from 0.2 to 60 tonnes/day, the largest being for the stillage concentrate stream from *shochu* production. These installations generally comprise a solubilization tank, the fermentation reactor and the membrane separator/biomass concentration unit, and can be coupled with an aerobic MBR (Fig. 5.12). Biogas from the fermentation reactor is used to scour the aniMBR membrane, which extracts the ammonia-rich permeate from the liquor and feeds this into the aerobic iMBR process treating the bulk wastewater stream. Removal of the ammonia is essential to prevent inhibition of the fermentation process.

5.2.1.11. *Anaerobic Kubota Systems: Ken's Foods*

Design/Build firm: ADI Systems Inc.

Possibly one of the most significant anaerobic immersed MBR (aniMBR) plants installed outside of Japan is the one recently provided by the Canadian company ADI Systems Inc. The installation at Ken's Foods, a salad dressing and barbecue sauce producer in Marlborough, MA, was commissioned in July 2008. This plant is the first of its kind in North America and possibly the largest of its type in the world, treating an average flow rate of 325 m^3/day, peaking at over 500 m^3/day. The plant was completed as a design/build project to upgrade the existing treatment train which consisted of a proprietary low-rate anaerobic *ADI-BVF*® (bulk volume fermenter) reactor and *ADI-SBR*® (sequencing batch reactor).

The existing anaerobic technology had worked well for several years treating wastewater of high TSS and FOG concentration. However, Ken's Foods wanted to expand production that would result in a 60% increase in wastewater flow and loading. Conversion of the existing system to an aniMBR through the addition of submerged membranes and sludge recycle allowed for expansion of the production plant. The aniMBR system was chosen for its economics, space-saving advantages, and ability to produce consistently high quality anaerobic effluent.

Raw wastewater with characteristics of 34,000 mg/L COD (18,000 mg/L BOD), 12,000 mg/L TSS and 1500 mg/L FOG is fed to the main anaerobic reactor (former *ADI-BVF®* reactor) via an equalization tank following alkalinity addition. The mixed liquor from the anaerobic reactor flows by gravity and is split evenly to feed four membrane tanks (Fig. 5.13), each containing seven 200-panel single-deck Kubota *ES200* units (and hence 1120 m^2 of membrane surface area per tank), and fitted with a retractable geomembrane structural cover connected to the biogas membrane scour system.

Membrane scour is provided by three biogas blowers (two duty and one standby), which supply biogas for continuously scouring the membranes. The biogas scour flow is used to sustain a net average membrane flux rate of 4 LMH at MLSS concentrations ranging from 20 to 45 g/L. Permeability recovery, invoked when the TMP reaches 0.1 bar, is through in situ membrane soaking for 2 h in a 10% citric acid solution. No such cleaning was required during the first 20 months of operation, although a scheduled maintenance clean has been performed on a six to eight week rotation (hence every six months for each individual tank) for three of the four membrane tanks. The membranes in Tank 4 have never been cleaned to date; the operational TMP for this tank was approximately 0.01 bar after 20 months of continuous operation.

The average treated effluent BOD, COD and TSS concentrations for the first 20 months of operation were 16,200 and <2 mg/L, respectively, corresponding to greater than 99.5% removal for each of these three determinants. Biogas produced in the aniMBR system is conveyed to a dual-fuel boiler for heating the treatment reactors to an operating temperature of 35 °C. Excess biogas is used for

FIG. 5.13 The covered aniMBR tanks at Ken's Foods.

hot water supply and building heating. Methane production has been 0.35 m^3 CH_4 per kg COD removed, and the biogas contains approximately 60% methane.

The enhanced COD and TSS removal via methanogenic conversion has significantly reduced operating costs relating to waste sludge discharge/de-watering and aeration requirements for the existing SBR. Further operational cost savings have resulted from the elimination/reduction in macronutrients, chlorine, anti-foam reagent, and polymer addition for the existing wastewater treatment system employed prior to commissioning of the new aniMBR.

5.2.2. Brightwater

5.2.2.1. Coill Dubh, Ireland

Client: Kildare County Council
Primary contractor: Brightwater Engineering Ltd/MDY construction
Primary consultant: Brightwater Engineering Ltd

Coill Dubh is a small village about a one-hour drive west of Dublin. The existing works consisted of basic concrete primary/septic tanks followed by peat filter beds. Effluent was directly discharged from these peat beds to a small local river. The works was in a poor condition and, on completion of the MBR scheme, was demolished.

The new works (Fig. 5.14) was provided by a developer as part of his planning consent for a housing estate close to the site. This works was to treat the sewage from this new estate plus the existing village. At the tender and design stage in 2001–2003, little information was available on the influent, other than it was to be mainly domestic in nature, contain storm flows, have a maximum flow to treatment of 50 m^3/h and contain loads of 2000 p.e. The treated effluent consent was BOD, 10 mg/L; SS, 10 mg/L; N-NH_4, 5 mg/L; and total phosphorus (TP), 1 mg/L. Though this standard could be achieved with

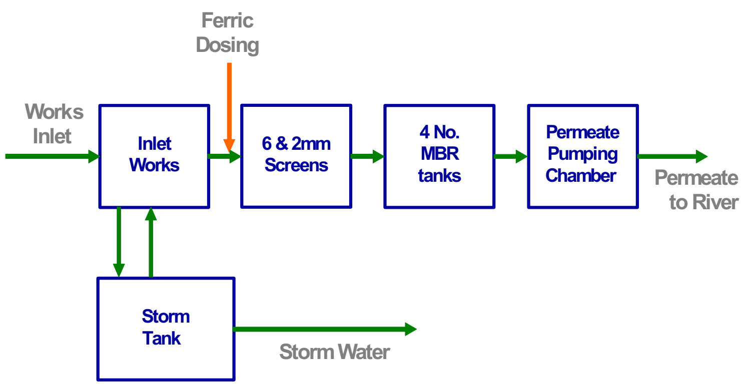

FIG. 5.14 Schematic of Coill Dubh WwTW.

a conventional treatment process, the small footprint and relatively shallow tanks afforded by the MBR process led to this technology being selected; the new works was to be built on a peat bog with a very high water table, making deep excavations difficult and expensive. The simplicity of the *MEMBRIGHT*® design made this option competitive and attractive.

No piloting was undertaken as this influent was considered as 'normal' domestic sewerage based on an inland catchment with an observable existing treatment process. The contract was led by the house developer MDY Construction. MDY also supplied the civil element to the contract based on supplied designs. The process, mechanical and electrical design element of the contract was delivered by Bord na Móna Environmental (BnM) and supplied by Brightwater Engineering. M&E installation was supplied by BnM subcontractors, while process commissioning, verification and pre-handover operational support was supplied by Brightwater. The plant was commissioned in June 2004.

The works (Fig. 5.14) consists of an inlet flume, followed by an overflow sump from where flows in excess of 50 m^3/h are diverted to a storm tank with 2 h retention at flows between 50 and 100 m^3/h. Excess flows from the storm tank are discharged from it directly to the stream. Storm water is pumped back to the inlet sump once works inlet flows have reduced. Flows to treatment are dosed with ferric for phosphorus control and pumped to a 6 mm free-standing wedge wire rotary screen. Flows drop from the screen into a stilling zone located in a settling tank that also acts as the works sludge and scum storage. From the stilling zone, flows pass under a fat-retaining submerged weir and onto a 3-mm drum screen. Flows are then diverted via a common manifold to four tanks, each containing six membrane modules. Each module provides a total effective surface area of 92 m^2 from 50 panels. Aeration is supplied to the modules only by duty/duty blowers. Treated effluent/permeate gravitates by siphoning from each reactor to a common collection sump via flow measurement on each reactor. Permeate is then pumped under level control to outfall, or can be recycled back to the inlet of the works via a divert valve.

Excess sludge is removed manually from each reactor via a single progressive cavity pump and discharged into the inlet sludge storage tank. Fat, scum and grit are also collected in this tank, and the accumulated waste periodically removed from site. Also supplied on-site are a control/mess building, ferric storage and a standby generator. A remote monitoring system provides Brightwater with a feedback of operational data.

The MBR operates at a maximum TMP of 300 mbar and a gross flux of 27 LMH with relaxation for 5 min every hour. Aeration is supplied by an aeration lateral system located at the base of each plate providing coarse bubble aeration for both membrane scour and process aeration at a rate of up to 118 Nm^3/h per module, or a SAD_m of 1.28 $Nm^3/(m^2\,h)$. No additional fine bubble aeration system is required for the combined sewer loads supplied to Coill Dubh. Although chemical cleaning is suggested at 12-month intervals, none has been applied over the first 18 months of operation; the performance of the membranes is maintained with the air scour alone. Cleaning is to be conducted ex situ by sluicing, followed by a 4−12 h soak in either hypochlorite or citric acid, depending on the assessed fouling. Flushing of the aeration system can be conducted at the same time. The design sludge age is 21 days, though this often increases to ∼30 days such that the design MLSS concentration of 12 g/L increases to 15 g/L. The HRT is 11 h at full flow.

The process at Coill Dubh was built as a low-tech approach to membrane technology and has performed better than anticipated despite periods of operational stress. The permeability remains acceptable at flux rates between 22 and 27 LMH and the permeate water contaminant levels have consistently been below most of the parameter limits of detection (i.e. 5 mg/L BOD, 5 mg/L SS, 0.5 mg/L N-NH_4 and 0.05 mg/L TP). While the effluent quality exceeds that required for discharge, the simplicity of the system and the robustness of the membranes have made this plant very simple to operate, despite fat and rag levels far in excess of those anticipated. The continued operation without the requirement for chemical cleaning makes the system even more attractive. A preventative inspection/clean is currently being discussed with the client, who is now Kildare County Council.

5.2.2.2. Other Brightwater Plant

Another of Brightwater's MBR plants (450 p.e. or 0.276 MLD flow design capacity), operated by Bord na Móna on behalf of the local council, is located at Halfway WwTW near Cork in Ireland (Fig. 5.15). This plant is fitted with

FIG. 5.15 The plant at Halfway. The screen is located mid/top picture, with the inlet pump chamber behind it. The anoxic tank is below the screen, and the MBR is the covered tank with the swan-neck air feed lines. The sludge tank is located behind the MBR and to the right of the anoxic tank. The permeate pump chamber is behind the sludge tank.

a 3 mm screw wedge wire screen and has an anoxic reactor (45 min HRT) for TN removal, the consented outlet concentration being 5 mg/L TN, and ferric dosing for phosphorus removal. The plant is fitted with six membrane modules in a single aeration reactor, providing a total area of 552 m^2 with a design flux of 21 LMH. All process aeration is supplied by the scour air system, as for Coill Dubh. Mixed liquors are recycled from the aeration reactor to the baffled and mixed anoxic reactor to enable nitrate removal. Screened and ferric-dosed influent is also fed into the anoxic reactor at this point so as to provide a carbon source for denitrification.

5.2.3. Colloide

5.2.3.1. Feenagh

Client: Limerick County Council
Primary contractor: Colloide
Primary consultant: Murnane Consulting Engineers

Feenagh is a small village in Co. Limerick in Ireland. Colloide were employed to design and build the complete treatment plant including all civil works for a 300 p.e. (60 m^3/day) treatment plant for the village, the plant (Fig. 5.16) having a discharge consent of 10 mg/L BOD and 10 mg/L SS. The plant, which started up in October 2009, had to be designed for significant diurnal fluctuations in flow and also allow for a low flow at initial start-up, with a modular design to expedite future extension. It comprises an inlet pumping station, a 3-mm spiral screen, primary settlement (12 h on average), and equalization of 20 h on average followed by aerobic treatment and membrane filtration in an 80 m^3 tank. Treatment tanks are of glass-reinforced plastic (GRP), and both the membrane assemblies and the control system were built and tested prior to delivery and installation on-site to reduce the project time. The two membrane streams each have one unit of 12 modules, providing a membrane area of 240 m^2. The mean net flux is 18 LMH, with the membranes operated on a cycle of 6 min filtration and 1.5 min relaxation. Membrane aeration is at

FIG. 5.16 (a) Installation of the tanks at Feenagh and (b) the membrane modules.

0–0.5 $Nm^3/(m^2\,h)$, with a TMP of up to 0.3 bar. A six monthly 2-h clean with sodium 100 mg/L NaOCl is applied at the site.

As of the end of 2009, there have been no membrane failures at any of the 24 Colloide plants in Ireland, the first of which was installed in 2003.

5.2.4. Huber

5.2.4.1. Arenas de Iguña

Client: EDAR Arenas de Iguña
Primary contractor: Huber Technology Espana
Primary consultant: EDAR Arenas de Iguña

The small North Spanish town of Arenas de Iguña is situated approximately 30 km south of the coastal town of Santander. Prior to the installation of the MBR plant, the town had no sewage treatment plant, despite the drinking water for the region being abstracted from the same river not far beyond the point from where the wastewater was being discharged. The plant (Fig. 5.17) was installed in August 2006 primarily to provide disinfection, as well as BOD, ammonia and nutrient removal. As of end 2009, the plant was the largest Huber installation in Spain with a design capacity of 4 MLD (20,000 p.e.) and a peak hourly flow of 416 m^3/h (corresponding to a peak loading factor of 2.2).

The feedwater passes through a 3-mm perforated plate screen and a grit channel before entering a 733 m^3 equalization tank (average HRT of 14 min). The plant is designed for biological nutrient removal with anoxic, anaerobic and aerobic tank volumes of 330, 330 and 2111 m^3, providing mean HRTs of 2, 2 and 12.7 h, respectively, with a recycle ratio of ~3. The four membrane trains/tanks, of 300 m^3 total volume, each house a *VRM® 30/448* unit

FIG. 5.17 The Arenas de Iguña site.

FIG. 5.18 The Huber *VRM* unit being installed at Arenas de Iguña.

(Fig. 5.18) comprising 448 modules of 6 m^2; a single module comprises eight 0.75 m^2 panels (Section 4.2.7), such that the total membrane area provided at the site is 10,752 m^2.

The plant is operated at an MLSS concentration of around 4 g/L in the aerobic tank, 6 g/L in the membrane tank, and at a net flux of 14 LMH (cf. a design flux of 17 LMH) based on a cycle of 9 min filtration and 1 min relaxation. Membrane scouring is employed at a SAD_m of 0.25 $m^3/(m^2\,h)$, giving a mean SAD_p of around 18. The TMP ranges from 25 to 250 mbar, corresponding to a permeability of 56–560 LMH/bar. The plant has been shown to sustain fluxes of 30 LMH at the membrane aeration rate provided, corresponding to a peak SAD_p of 8.3. It has received one 6 h recovery clean with 250 mg/L hypochlorite in three years of operation, with no maintenance cleaning.

The product water is free of pathogens and has BOD_5, COD, ammonia, TN and TP levels of <5, <20, <1, <10 and <1 mg/L, respectively.

5.2.4.2. Hutthurm

Client: Markt Hutthurm
Primary contractor: Huber SE
Primary consultant: GFM, Beratende Ingenieure

The MBR plant at the Hutthurm WwTP, in the Bavarian region of Germany, was installed in July 2008 to upgrade the existing works and thus protect the environmentally sensitive River Ilz to which the plant discharges. The site receives a large proportion of its organic load from local breweries, providing around 1320 kg of BOD per day. The MBR was chosen ahead of the alternative option (of tertiary treatment by sand filtration followed by UV disinfection) because of the challenge imposed both by the high loading and the geological and spatial limitations of the site: space at the works was very limited, the ground rocky and the groundwater table high. The plant has a design capacity of 3.4 MLD with storm flows of 220 m^3/h and, as at the end of 2009, was the largest MBR plant in Bavaria.

Huber have provided the main process technologies for the site, including the screens and the mechanical thickener for the WAS. The wastewater is screened to 5 mm using a Huber *ROTAMAT®* wedge wire drum screen, degritted and then fine screened to 1 mm though a second *ROTAMAT®* screen via a 1000 m^3 balancing tank (and hence 7.1 h HRT at 3.36 MLD). The MBR system comprises two aeration basins, 1300 m^3 in total volume (9.3 h HRT), and four filtration tanks of 250 m^3 total volume. Three of these are currently fitted with three *VRM® 30/544* units, each having a membrane surface area of 3264 m^2, with the fourth tank to be used for future plant expansion.

The plant is operated at an MLSS concentration of around 10 g/L in the aerobic tank, 12 in the membrane tank, and at a net flux of around 20 LMH – somewhat higher than its design flux of 15.7 LMH – based on a cycle of 9 min filtration and 1 min relaxation. Membrane scouring is employed at an SAD_m of 0.245 $m^3/(m^2\,h)$, giving a mean SAD_p of around 16. The TMP ranges from 20 to 400 mbar, with a corresponding permeability of 100–780 LMH/bar. The plant can sustain fluxes of 25 LMH at the design SAD_m, yielding a peak SAD_p of 9.8. The membrane is cleaned with 3 g/L hydrogen peroxide on a quarterly basis, incurring a downtime of 24 h per clean.

The product water is essentially free of total coliforms (<1 CFU per 100 mL), has <20 mg/L COD, <10 mg/L TN and has non-detectable ammonia and BOD_5 levels.

5.2.4.3. Hans Kupfer & Sohn, Heilsbronn

Primary contractor: KG Nellingen
Primary consultant: IB Dr. Resch, Weißenburg

Hans Kupfer & Sohn GmbH based in Heilsbronn, Bavaria is one of Germany's largest meat processing companies. The MBR plant at their site was installed due to increased capacity and clarification requirements at the site. The plant has a flow capacity of 1.6 MLD and 100 m^3/h peak hourly flow, and was commissioned in January 2008.

The plant (Fig. 5.19) combines mechanical, physicochemical and biological wastewater treatments, with additional sludge treatment, all technologies being provided by Huber. The wastewater at 25–35 °C passes through a Huber *RakeMax®* bar screen (size 6300/952 with 15 mm bar spacing) and then two parallel *ROTAMAT®* rotary drum 1 mm fine screens (*Ro2* units), with supplementary high pressure cleaning. After intermediate 1500 m^3 buffer

FIG. 5.19 The WwTW at Hans Kupfer & Sohn.

storage the wastewater is fed to a dissolved air flotation (DAF) plant, *HDF* size 10, with polyaluminium chloride (PACl) dosing, for fat and grease removal. The water then passes to the 900 m^3 aerobic tank (13.5 h HRT at the 1.6 MLD flow capacity) with the sludge flow recirculated to the 210 m^3 capacity membrane tank containing three *VRM® 30/400* units providing a total membrane surface of 7200 m^2. The waste MBR sludge is thickened using a *ROTAMAT®* disc thickener (*RoS2S* size 1) with polymer dosing.

The plant is operated at an MLSS concentration of around 5 g/L in the aerobic tank and 7 g/L in the membrane tank, and at a net flux of 8 LMH, cf. a design flux of 12 LMH, with a cycle of 9 min filtration and 1 min relaxation. Membrane scouring is employed at a SAD_m of 0.25 $m^3/(m^2\ h)$, giving a mean SAD_p of ~42. The TMP ranges from 20 to 400 mbar, with a corresponding permeability of 100−780 LMH/bar. It seems likely that the plant can operate under more challenging hydraulic conditions. The membrane is cleaned once a year with 500 mg/L hypochlorite and 500 mg/L citric acid, with a downtime of 24 h per clean.

The feed and effluent characteristics are given in Table 5.5. The plant has worked well with a high operating stability and the required effluent quality standards for direct discharge. The reduction in sludge volume has led to reduced disposal costs. The success of the installation contributed to Kupfer receiving the 2009 Bavarian Environmental Award from the Bavarian Landesstiftung.

5.2.5. LG Electronics

5.2.5.1. Municipal Wastewater Pilot Plant

Primary contractor: LG Electronics/GS Neotek/KOReD
Primary consultant: LG Electronics/GS Neotek/KOReD

There is a small 0.1 MLD ADF LG membrane-based MBR pilot plant in Seoul treating water for non-potable reuse, with post-treatment of the MBR permeate

TABLE 5.5 Feed and Treated Effluent Quality for the Plant at Kupfer

Parameter	Concentration		% Removal
	Influent	Effluent	
COD	1600	<30.0	98.0
BOD	1100	<3	99.7
TN	70	<5	95.5
NH_4-N	Not meas.	<1	–
NO_2-N	Not meas.	0.1	–
NO_3-N	Not meas.	2.0	–
TP	25 mg/L	0.2	99.2

by reverse osmosis. Pre-treatment takes the form of flow equalization for 1 h, clarification for 3 h and fine screening with a 0.5 mm drum screen. Biological nutrient removal is provided by tank volumes of 18, 6 and 20 m^3 for the anoxic, anaerobic and aerobic tanks, respectively, which, coupled with the 11.9 m^3 membrane tank, give a total HRT of 13 h. The MLSS concentration in the aerobic and membrane tanks is 6.5 and 9.5 g/L, respectively.

The membrane tank is fitted with two 150-panel modules providing a total membrane surface area of 300 m^2 and thus a mean operational flux of 23.7 LMH. Operation is on a cycle of 11 min filtration and 1 min relaxation, and with a SAD_m of 0.6 $Nm^3/(m^2\,h)$, and thus a SAD_p of 25. Maintenance cleans are employed six times a month, and comprise a 3-h soak in 0.5 wt% NaOCl. This is supplemented with cleans in 0.1 wt% NaOH twice a year. The MBR operates at a specific energy demand of 2.1, 4.2 and 0.1 kWh/m^3 permeate for the membrane scouring, total aeration and sludge transfer, respectively. The plant achieves 99%, 98%, 80% and 70% removal of BOD, ammonia and TN and TP down to 1.1, 0.2, 4.4 and 0.85 mg/L, respectively.

5.2.6. Shanghai Sinap

5.2.6.1. Songjiang District Landfill Leachate Treatment Plant

The small 150 m^3/day plant in the Songjiang District receives water of around 14,000 mg/L COD, 3200 mg/L BOD and 1800 mg/L ammonia at between 21 and 37 °C, discharging the treated effluent (Table 5.6) to a common sewer. The feed is screened to 3 mm and pH adjusted before entering the biotank where the MLSS is held at 11.2 g/L (13.3 g/L in the membrane tank). The total biotank volume is 2030 m^3, including a 1350 m^3 anoxic and 547 m^3

TABLE 5.6 Water Quality, Songjiang

Parameter	Concentration	% Removal
TSS	114	96
BOD_5	160	95
COD	415	97
Ammonia*	56	97
TKN*	97	95

*as N.

aerobic/membrane tank. The membrane tank is fitted with 720 m^2 of membrane area provided from eight 60-panel units. The membranes operate at a mean flux of 9.3 LMH, a 5-min filtration cycle including 60 s of relaxation, TMPs between 0.1 and 0.4 bar and a SAD_m of 2.1 $Nm^3/(m^2\,h)$ (and so an SAD_p of more than 200). Recovery cleans every 20–60 days employ HCl to remove phosphate fouling.

5.2.7. Toray

5.2.7.1. Heenvliet

Client: Waterschap Hollandse Delta
Main consultant: Witteveen + Bos
Main contractor: Keppel Seghers

The plant at Heenvliet is actually a demonstration plant which forms part of the STOWA (Dutch National Programme), and comprises an MBR coupled with a CASP (Table 5.7). It can be used both as a conventional MBR (Fig. 5.20a), thus accepting the screened sewage directly, or coupled with the aeration tank from the existing 7 MLD capacity CASP (Fig. 5.20b). The membranes permeate the entire flow during periods of low flow, but during storm flows the secondary clarifier becomes active. This allows the MBR to be operated at full capacity, thus maximizing its utilization, whilst also reducing the hydraulic load on the secondary clarifier during storm flows and so maintaining high product water quality throughout. The MBR plant was commissioned at the start of 2006 by the Dutch Waterboard Hollandse Delta and is designed to treat 2.4 MLD daily flow (0.91 MLD dry weather flow from 13,000 p.e.); above this flow the secondary clarifier is operated. The MBR provides biological nutrient removal through the provision of anoxic and anaerobic zones (Table 5.7), with a recycle ratio of 3 from the membrane to the anoxic zone.

TABLE 5.7 Operating and Performance Parameters for the Hybrid Plant at Heenvliet

Parameter	CAS	MBR
Design dry weather flow, L/s	31	10.5
Design full flow to treatment, L/s	80.5	28
Design loads, kg/day		
COD	986	340
BOD	365	126
N (as TKN)	88	30
Tank volumes, m^3		
Anoxic	590	54
Anaerobic	–	76
Aerobic	1170	101
Membrane	–	176
Product water quality		
TSS, mg/L	6.7	<4
BOD_5, mg/L	3.6	1.2
COD, mg/L	34	24.6
Ammonia, mg-N/L	2.65	0.13
TKN, mg-N/L	4.41	1.01
Total nitrogen, mg-N/L	7.59	3.06
Total phosphorus, mg-P/L	2.69	2.23
Fecal coliform bacteria, MPN/100 mL	1253	52
Cu (μg/L)	<2	2.4
Zn (μg/L)	20	16.4
As (μg/L)	4.8	4.1

When operating as a conventional plant, wastewater is fed to the plant via a 6 mm coarse screen followed by a Jones and Attwood 3 mm fine screen, with around 4 h of equalization provided. The MBR plant comprises two parallel membrane tanks, each equipped with four Toray *TRM140* double-deck units. Each unit offers a membrane area of 526 m^2 from 376 panels, giving a total

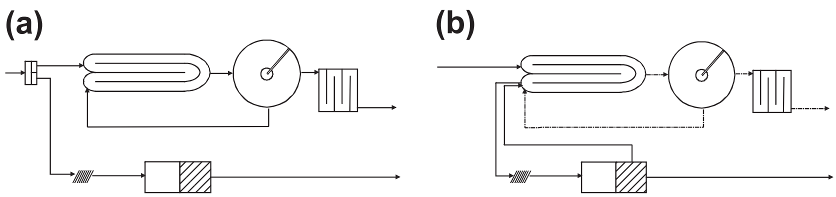

FIG. 5.20 The Heenvliet plant, with the MBR configured: (a) conventionally and (b) in hybrid mode.

membrane area of 4200 m^2. The plant operates at a target MLSS of 10 g/L in the aerobic zone and 14 g/L in the membrane tank, maintained by sludge wasting at 500 kg DS/day.

Scouring air is delivered at a rate of 0.29 $m^3/(m^2\,h)$, which at the normal net operating flux of 20 LMH yields a SAD_m of 14. The flux is sustained through relaxation for 1 min following 9 min of filtration, and the maximum TMP is 0.2 bar (minimum permeability of ~150 LMH/bar). Fluxes of 24 LMH have been attained by the plant, though the nature of the coupling with the CASP means that high flux operation is unnecessary. Recovery cleaning of the membranes is done roughly four times a year through soaking in 2–6 g/L hypochlorite followed by 1–3 wt% citric acid for at least 3 h each. Clogging has been evident at this site, but the permeability is generally recovered by the chemical clean. Membrane aeration incurs a mean energy demand of 0.37 kWh/m^3 permeate.

Heenvliet is an important European demonstration plant (Fig. 5.21), and has formed the focus of a number of research projects from which many papers have resulted (Evenblij, Geilvoet, van der Graaf, & van der Roest, 2005; Verrecht, Judd, Guglielmi, Brepols, & Mulder, 2008; Moreau, Ferreira, & van Nieuwenhuijzen, 2009; Geilvoet, 2010). In addition to assessing the viability of hybrid operation, the plant provides an important indication of the relative efficacy of the CAS and MBR for removing key contaminants. A comparison of the permeate and clarifier water quality reveals there to be only marginal differences in many ostensibly soluble micropollutants such as Cu, Zn and As – an observation corroborated by a recent review of the subject (Santos & Judd, 2010). However, significant differences in performance arise for other contaminants – most notably BOD, ammonia and total coliforms.

5.2.7.2. Fuji Photographic Production Plant, Tilburg

Main consultant: Keppel Seghers
Main contractor: Keppel Seghers

The plant at Tilburg in the Netherlands (Fig. 5.22) is a photographic paper production plant, owned by Fuji, discharging waste high in dissolved organic matter (Table 5.8). This project was initiated in 1998 when a number of drivers combined to make the existing management scheme untenable. First, the

FIG. 5.21 Heenvliet WwTW: (a) from the air, (b) from the side and (c) inside the membrane tanks.

FIG. 5.22 Overview of plant at Tilburg. Front right: MBR plant building; front left: building housing RO, dewatering unit, storage of chemicals and permeate and control room.

TABLE 5.8 Effluent Characteristics, Photographic Production Plant

Parameter	Unit	Design value
Mean flow (peak)	MLD	0.84 (1.080)
Temperature	°C	37
pH		7–8
COD concentration (load)	mg/L (kg/day)	1800 (1512)
BOD concentration (load)	mg/L (kg/day)	850 (714)
BOD/COD	–	0.7
TSS concentration (load)	mg/L (kg/day)	48 (40)
Total solids concentration (load)	mg/L (kg/day)	79 (94)
TKN	mg/L	35
NO_3-N (NO_2-N)	mg/L	30 (25)
Total P concentration (load)	mg/L (kg/day)	5 (4.2)
BOD/P ratio	–	170
Ag concentration (load)	mg/L (kg/day)	47 (40)

province where the factory was based was advocating reduced groundwater usage. Trace chemicals discharged in the wastewater, although not blacklisted at the time, were viewed as being onerous. Costs levied by the municipality for wastewater treatment were increasing. Also, the temperature of the water discharged in the summer months was close to the permitted limit of 30 °C. There was also a supplementary cost associated with silver recovery from one of the sludge by-products, this process being outsourced to a specialist company.

A cost benefit analysis revealed that for an estimated capital investment of €2.3m for the recovery plant, savings in sludge treatment, energy and chemicals (amounting to €0.6m/a) would provide a payback of <4 years. Additional savings arising from reuse of reverse osmosis (RO) permeate from the recycling process were estimated to reduce the payback time further to two-and-a-half to three years. It was on this basis that piloting was initiated in 1998 using a 50-L/h sMBR pilot plant followed by a 500-L/h plant in 1999, coupled with an RO pilot plant with a 400-L/h capacity. A three-month pilot trial based on an iMBR pilot plant, estimated to improve further the cost benefit by ~€100k/a in energy costs, was successfully conducted in 2002 by the consultants for the project (Keppel Seghers) and led to the installation of the full-scale plant.

The plant comprises an MBR-RO process with a 0.75 mm pre-screen. Sludge flows under gravity from the membrane to the aeration basin with pre-denitrification (Fig. 5.23). Aeration is achieved with disc aerators powered by one or two air compressors which are frequency-controlled to attain the target reactor DO level. Membrane filtration is with 12 continuously aerated Toray stacks of 135 m^2 area, each fitted with 100 panels, providing 1620 m^2 membrane area in total. Excess sludge at ~18 g/L from the membrane basin is pumped to a peristaltic decanter where the sludge and silver are separated from the water, the supernatant being returned to the denitrification basin. The dewatered product of 20–24% is ultimately processed to extract the silver (300–400 tonnes annually).

The ultrafiltration (UF) unit is designed for a maximum throughput of 45 m^3/h and a nominal flow of 35 m^3/h (0.840 MLD), equating to a flux of 21.6 LMH (27.8 LMH peak). Membranes are projected to operate at a TMP of 15 mbar and thus a mean permeability of around 1500 LMH/bar at 20 °C. At the SRT and HRT of 24 days and 14 h, respectively, the MLSS concentration is 15 g/L, of which ~12 g/L is estimated to be biomass. Membranes are operated on a cycle of 8 min on/2 min relaxation, with chemical cleaning for 4 h with up to 6 g/L NaOCl planned on a yearly cycle. Membranes are aerated at 0.4 Nm^3/h per m^2 membrane, providing a SAD_p of 18.5 Nm^3 air per m^3. This compares with fine bubble aeration rates of 11–22 Nm^3 air per m^3 permeate for biotreatment.

The permeate from the UF membranes is pumped to the RO system, housed in an adjacent building (Fig. 5.22), via two heat exchangers to recover heat for part of the boiler feed water. The inlet design temperature of the RO is set between 20 °C and 25 °C. The RO permeate is delivered to the cooling towers as make-up water. The three-stage (a 4/5-2-1 array with six elements per module) RO unit is fitted with Toray *SU-720* modules and normally operates between 11 and 14 bar with a recovery of up to 85%. The first stage has a spare line in case an additional pressure vessel is needed. If maximum flow conditions prevail over a long period, this vessel may be fitted with six extra elements

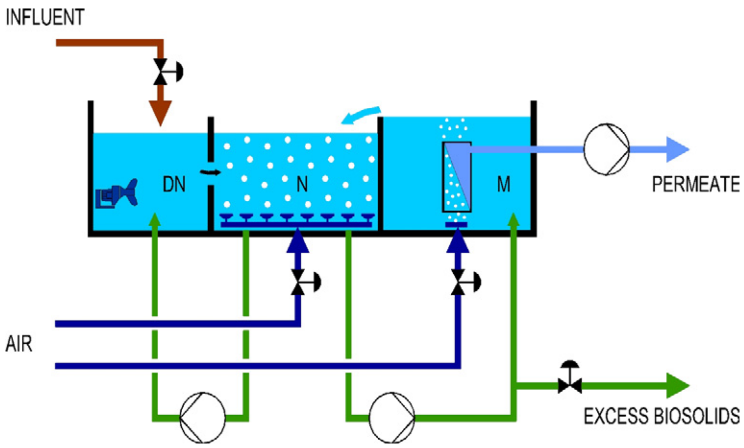

FIG. 5.23 The Tilburg MBR plant, schematic.

to reduce the hydraulic load. The RO permeate is stored in a buffer tank prior to transporting to the ion exchange filter feed tank. The brine is discharged into sewer since it contains only inert salts.

The plant is guaranteed to meet effluent concentration limits of <51 mg/L COD (>97% removal), undetectable TSS and 5 mg/L TN. Thus far, some minor teething problems have been experienced since commissioning. These have included foaming, linked to wider than expected variation in feedwater quality, and concrete erosion. Further improvements, and specifically automated control, have been implemented.

5.2.7.3. Other Toray Plant

Toray are active in many regions of the world, and particularly so in the Middle East. The company has been awarded, by Wetico (Water & Environment Technologies Company Ltd), the contract for a 60 MLD municipal MBR plant for the city of Najran in the Kingdom of Saudi Arabia. Under the contract, Toray is to supply 320 MBR modules with a total membrane area of 89,600 m^2, with commissioning of the plant expected sometime in 2011. This will represent the world's largest Toray MBR plant.

The largest Toray plant in the world as of the end of 2009 was the Yas Island WwTP, serving the Formula One race track and the surrounding tourist district. The plant, for which the primary consultant and contractor are Halcrow and Metito respectively, is to have an ultimate capacity of 57 MLD provided by three trains of four tanks. As of March 2010 two of the three streams had been fitted with membranes, the third planned for future expansion. Each tank has space for twenty-four 200-panel double-deck units, and hence 26,880 m^2 membrane area is provided per train. The modules are aerated at a rate of 0.43 $Nm^3/(m^2\,h)$ and the peak design flux is 30 LMH. Pre-treatment comprises a 12 mm bar screen and then a 1.5 mm Serco drum mesh screen followed by a spiral aerated grit separator. The biotreatment includes an anoxic zone and, as is the case with almost all wastewater treatment plants in the region, the permeate water is used for irrigation.

As of March 2009 the plant was operating at well below capacity due to the limited flow to the works. The peak salinity levels of the feedwater has also been very high at around 4000 mg/L – four times the design value, but this appears not to have impeded the steady-state microbiological performance of the plant.

5.2.8. Weise Water Systems GmbH

5.2.8.1. Holiday Inn, Selbourne, United Kingdom

Client: Holiday Inn
Primary contractor: Weise Water Systems GmbH
Primary consultant: EnSo International Ltd

Weise primarily provides submerged UF cassette modules and small packaged plants on a service contract basis. An example of the use of the company's

Microclear® technology is at an installation at Selborne (Fig. 5.24), in Hampshire (United Kingdom) by the environmental solution provider EnSo International Ltd. The plant treats effluent from the 'Holiday Inn' Hotel, near Winchester, and has a capacity of 76 m^3/day (2.4 L/s peak, corresponding to a peak loading factor of 2.7). An MBR was specified based on the close location of the Hotel to a drinking water sourcing area. Installation was completed in December 2009.

Incoming wastewater from the kitchen enters a dual-section 18 m^3 settlement/storage tank (Fig. 5.24) where gross solids are settled out before the kitchen wastewater enters the grease trap under gravity. The effluent from the grease trap is combined with other wastewater from the site and enters a similar settlement tank of three sections and 36 m^3 volume. Settled sewage then enters the main 36 m^3 aeration tank, which provides aerobic treatment and flow balancing, prior to being pumped to the membrane tank. Each of the two membrane sections is equipped with a filter housing type *MA04-30* containing 105 m^2 membrane surface per filter housing fitted with 15 *MCXL* membrane units of 7 m^2 each. The MBR sections are equipped with a sludge pump. Filtered effluent (permeate) is pumped into a permeate header tank, from which the final effluent flows under gravity into the sample chamber. The electrical controls, blowers and associated pipework are housed within a kiosk.

The plant operates at an average flux of 15 L/m^2 h from 210 m^2 total area, with only one projected chemical cleaning per year. The BOD_5 and NH_4-N loadings are 30 and 3.3 kg/day, with corresponding feed concentrations of 396 and 43 mg/L, and the discharge consent is 5 mg/L for both BOD and NH_4-N.

FIG. 5.24 The storage and biotreatment tanks at Selborne, UK.

5.2.9. Other Membrane Module and Technology Suppliers

5.2.9.1. Oerlemans Foods Netherlands B.V

Primary contractor: Triqua B.V.
Primary consultant: Triqua B.V.
Membrane technology: A3

Oerlemans Foods Netherlands B.V. produce freshly frozen vegetables, fruit and potato products for international food service, retail and industrial markets. MBR technology was selected to meet the challenge presented by highly variable COD and BOD loads from process and run-off waters produced at the Broekhuizenvorst site, in the south of the Netherlands. Triqua B.V. secured the project in 2008 on Build Own Operate Transfer (BOOT) conditions based on a 10-year term.

The plant (Fig. 5.25) treats 5−10 m^3/h of effluent, i.e. 0.24 MLD PDF. Pretreatment at the plant comprises duty and standby rotating drum screens, with both operating during periods of high flow. The drum screen filtrate flows under gravity to an 800 m^3 mixed buffer tank before being pumped to a lamella plate separator. The settled wastewater overflows under gravity into the bioreactor, the primary sludge being periodically pumped to a storage tank at a mean rate of 9 m^3/day (1.2% DS) equating to an SRT of around 80−90 days. The MBR aeration tank has a working volume of 800 m^3, and is aerated by fine bubble diffusers. The submerged membrane system is located in a separate membrane tank of 89 m^3. The membrane system consists of three frames of 2×3 double-deck A3 membrane modules of 70 m^2 each

FIG. 5.25 The MBR installation at Oerlemans.

(Section 4.2.2), giving a total membrane area of 1260 m^2. The membranes are aerated at a rate of 0.36 $Nm^3/(m^2\,h)$ to maintain the mean flux of ~6 LMH based on a cycle of 8−9 min filtration and 1−2 min relaxation.

In 2009, the measured influent Total-COD concentrations varied between 1201 and 14,150 mg/L and TKN between 2 and 328 mg/L. However, the effluent quality was constant: TCOD concentrations varied between 5.5 and 82 mg/L and TKN between 7.6 and 12 mg/L. Average COD and TKN removal over 2009 were 99.3% and 95.9%, respectively, with the flux stable at 5−10 LMH. External cleaning of the membranes was carried out in 2009 only once. Full recovery of membrane flux was through treatment with enzymatic cleaning agent to remove fat from the membranes. During 2009 cleaning the membranes in place with sodium hypochlorite was carried out only twice.

The plant has operated successfully for one-and-a-half years at highly variable COD and TKN loads with near-constant treated effluent. Permeability reduction due to accumulation of biomass and fats on the membranes has been shown to be reversible through chemical cleaning.

5.2.9.2. Hitachi

Hitachi have been installing containerized plants based on a 200 m^2, 200-element cassette (Section 4.2.17) since 1997, and since 2006 in the United Arab Emirates. The Japanese plants, not necessarily containerized, range in capacity from 0.05 to 2.4 MLD and are mainly employed for industrial effluent treatment. The UAE plants are predominantly installed at construction sites, used for treating wastewater from the worker camps. Ten such plants, based on units of 250, 500 and 750 m^3, were installed in Dubai alone between October 2006 and March 2009; the combined total capacity of the 34 units was 16.5 MLD.

5.3. IMMERSED HOLLOW FIBRE (HF) TECHNOLOGIES

5.3.1. GE Zenon

5.3.1.1. Perthes en Gatinais

The plant at Perthes en Gatinais is one of the earliest examples of a Zenon-based MBR sewage treatment plant, commercialized as the *Biosep*™ process in the early 1990s by Veolia, and the first such plant in Europe. It was commissioned in 1999 and built and operated by Veolia Water (formerly Vivendi). The plant treats the effluent from four small communities in the Paris region, and is designed to meet a discharge content of 40 mg/L COD, 5 mg/L BOD and TSS, 10 and 2.5 mg/L TN and TP, respectively, and 5 log removal TCs.

The flow and loads for the plant are given in Table 5.9. The complete process treatment scheme (Fig. 5.26) includes pre-screening to 1 mm, biotreatment, membrane filtration and sludge dewatering. The MBR is based on a 500-m^3 cylindrical aeration tank (Fig. 5.27) with the sludge depth of 5−6 m,

TABLE 5.9 Characteristics of Plant and Effluent, Perthes en Gatinais Plant

Parameter	Units	Value
Population equivalent	p.e.	4500
Dry weather daily flow rate	m^3/day	900
Rain weather flow	m^3/day	1440
Peak hourly flow	m^3/h	140
Maximum flow rate at inlet Biosep™	m^3/h	85
COD load	kg/day	675
TSS load	kg/day	315
Total N-load	kg/day	67.5
Total P-load	kg/day	18
Temperature range	°C	10–20

the level variation to provide buffering during rainfall. This equates to a hydraulic retention time of around 13 h, and the plant operates at an SRT of ~25 days. The MLSS concentration in the aeration tank is approximately 15 g/L. The membrane plant comprises seven cassettes, each fitted with eight *ZW500* modules, providing a total area of 2604 m^2 (372 m^2 per cassette). At the hydraulic loading given, the flux through the membrane is around 18 LMH at average flow and 29 LMH at peak flow.

Coarse bubble aeration is applied at around 1 $m^3/(m^2\,h)$. The permeability is routinely between 70 and 120 LMH bar^{-1} over a range of 0.2–0.5 bar TMP. Regular flushing, for 45 s every 10 min, together with a weekly CIP with hypochlorite enables the required steady-state flux rates to be maintained. External recovery cleaning with sodium hypochlorite is conducted by transferring the membrane cassette to an adjacent cleaning tank where both cleaning

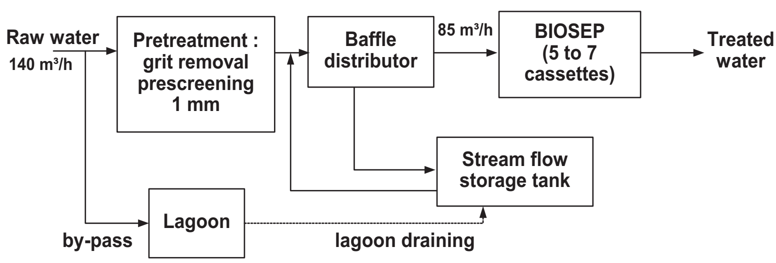

FIG. 5.26 Process flowsheet, Perthes en Gatinais plant.

FIG. 5.27 Aeration tank and covers, Perthes en Gatinais plant.

chemical reagents (acid and sodium hypochlorite) and cleaning durations can be employed more flexibly.

Phosphate removal is achieved by dosing with ferric chloride. The WAS generated is around 0.5 kg DS/kg COD. The dewatering unit consists of a sludge thickener with a belt press. To achieve a 16 wt% DS after the belt press, a polymer dose of 6 kg/tonne DS is required. After dewatering, the solids content of the sludge product is increased and the sludge is further stabilized with lime at a rate of 0.52 kg CaO/kg DS. The end product, generated at a rate of around 270 kg/day, has a dry solids concentration of 25 wt% and is distributed to local farmers for land application.

5.3.1.2. Nordkanal

Client: Erftverband
Primary contractor: Krüger Wabag GmbH

The Nordkanal plant in Germany, owned and operated by the Erftverband (Erft Association), treats wastewater from the nearby towns of Kaarst, Korschenbroich and Neuss. It was installed and commissioned in January 2004 following the success of the group's first MBR plant at Rödingen, a smaller plant which was commissioned in 1999 (Table 5.10). The plant is the largest in Europe, with a population equivalent of 80,000 and an installed capacity of 48 MLD.

The site has five buildings which respectively house the sludge mechanical dewatering process, the fine screens, the coarse screen, the membrane bioreactor and the process controls. Additional installations include lidded sludge

TABLE 5.10 Comparison of MBR Plants at Rödingen and Nordkanal

Parameter	Rödingen	Nordkanal
Design		
Capacity, p.e.	3000	80,000
Bioreactor volume, m^3	400	9200
Membrane area, m^2	5280	84,480
Sludge concentration, g/L	12−18	12
Operation		
Sludge loading, kg BOD/(kg MLSS d)	0.04	<0.05
Hydraulic retention time, h	3.6	4.6
Sludge retention time, d	25	25
Peak flux, LMH	28	25

and sludge liquor holding tanks, a grit chamber and the denitrification tanks, the two latter operations being open to the atmosphere (Fig. 5.28). Water is pumped from the original wastewater treatment plant 2.5 km east of the site to the 6 mm step screens. It is then fed via an aerated grit chamber to two Huber rotary drum 1.5 mm mesh-grid fine screens, changed from the 0.5 mm fine screens originally installed, which each provide a capacity of 24 MLD. There is a standby 1 mm fine screen which comes on-line in case of a mechanical breakdown of the former two. Screenings from the fine screens are discharged into the sludge dewatering process and subsequently disposed of by incineration off-site. The screened water is transferred to the MBR.

FIG. 5.28 Aerial view of the Nordkanal site near Kaarst.

Biotreatment comprises four tanks, each fitted with two membrane trains with an upstream anoxic (denitrification) zone of between 2600 and 3600 m^3 and an aerobic (nitrification) zone of 5600 m^3 including the membrane zone of around 1500 m^3, giving a total tank capacity of 9200 m^3. The anoxic zone receives sludge from the subsequent membrane aeration tank at a recycle ratio of 4:1. The eight membrane trains are each fitted with 24 *ZW500c* membrane cassettes of 440 m^2 membrane area each (20 m^2 per module; 22 modules per cassette), such that the total membrane area is 84,480 m^2 (24 units of 440 m^2 in eight trains). At the dry weather flow of 24.5 MLD the HRT is around 9 h. The aeration tanks receive supplementary mechanical agitation from impeller blade stirrers to maintain biomass suspension. Sludge wasted from the tank is dewatered by centrifuge to around 25 wt%. There is also a simultaneous precipitation for phosphorus removal.

The plant is operated at an SRT of around 25 days, which maintains the mixed liquor at between 10 and 15 g/L, with the average of around 12 g/L. The membranes are operated at a mean net flux of 23.7 LMH and a mean permeability of 150−200 LMH/bar. Intermittent coarse bubble aeration is provided on a 10 s on/10 s off cycle, giving a SAD_m of 0.43 Nm^3 per m^2 h and a SAD_p of 17 m^3 air per m^3 at the mean net flux. Physical cleaning comprises a backflush every 400 s for 50 s at a flux of around 38 LMH. The number of membranes on-line is adjusted according to the flow, but this is controlled in such a way as to ensure that no membrane train is off-line for more than a total of 70 min. A 500-mg/L hypochlorite (adjusted to pH 10) maintenance chemical clean in place (CIP) supplemented by a 0.2 wt% clean with citric acid (pH 2.5−3) is conducted weekly. Cleaning is conducted on individual tanks, such that the installed capacity is reduced by only 12.5% when cleaning is instigated. Recovery cleans, when required, employ 1000 mg/L hypochlorite and the membranes are cleaned externally (COP).

The total specific power demand for all operations, including pre-treatment and sludge processing, is 0.9 kWh/m^3 on average, the precise value varying between 0.4 and 1.7 kWh/m^3 depending on the flow. On average, the membrane air scouring contributes 49% of the energy demand. Process aeration in the bioreactor (12.1%), biomass circulation (1.3%), bioreactor mixing (11.5%) and permeate suction (2.7%) each demand significantly less. The remainder (23.2%) is consumed by pumping stations, pre-treatment, sludge dewatering and miscellaneous process units. An average of 0.7 kWh/m^3 has been calculated for all the conventional sewage treatment plants operated by Erftverband: the MBR is thus ~30% higher in specific energy demand. The Nordkanal plant, which incurred a total capital cost of €25m, has been successfully operating for seven years without major incidents, and details of the plant design and operation have been widely presented and published (Brepols, Drensla, Janot, Trimborn, & Engelhardt, 2008; Verrecht et al., 2008; Brepols, Schäfer, & Engelhardt, 2009, 2010; Blšťáková, Engelhardt, Drensla, & Bodík, 2009; Brepols, 2010).

5.3.1.3. Brescia

Client: a2a S.p.A (former ASM Brescia S.p.A.)
Primary contractor: Zenon Environmental (now GE)
Primary consultant: Zenon Environmental (now GE)

The Verziano conventional WwTP at Brescia originally had three trains, each consisting of primary clarification, biological oxidation, secondary clarification and final chlorination (Fig. 5.29). Pre-treatment of the incoming untreated wastewater prior to the three trains was undertaken by coarse and fine screening and with sand filtration for oil and grease removal. Trains A and C were identical in design capacity, treating 24 MLD each, while Train B had an original capacity of 12 MLD. Shortly after assuming operations at the Verziano WwTP in July 1995, the water company ASM Brescia undertook a series of evaluations and identified problems relating to additional discharges. The most significant problem was plant capacity, which was about two-thirds of that required. In response, several important projects were undertaken over the subsequent five-year period (1995–2000) to remedy these issues.

The principal remedial step was the upgrading of Train B, the oldest of the three trains, from 12 MLD to 38 MLD annual average flow capacity by converting the activated sludge treatment system to an MBR. High effluent quality and limited available footprint were the key driving forces leading to the decision to employ MBR technology for the plant upgrade. The need for improved effluent quality resulted from new Italian legislation (Community Directive 91/271/CEE and D.Lgs. 152/99), requiring mandatory nitrification/denitrification and effluent TSS limits lowered from 80 mg/L to 35 mg/L (Table 5.11). In addition, ASM had extremely limited available land to extend the plant footprint. Estimates for conventional processes treating the same capacity and incoming loads showed that the footprint required would be twice that of an MBR for the biological reactor and roughly six times greater for the final sedimentation. The decision to pursue the MBR solution was also made easier by prior experience with the same membrane technology at full scale from the

FIG. 5.29 The original plant at Brescia.

TABLE 5.11 Discharge Design Criteria, Brescia Plant

Parameter	Units	Regulation Limit	Design Value
COD	mg/L	125	125
BOD	mg/L	25	25
TSS	mg/L	35	5
Total N	mg/L	10	10
Total P	mg/L	2 (since 2009)	10
Escherichia	UFC/100 mL	–	10 on 80% of the samples 100 on 100% of the samples

summer of 1999 onwards for the plant's leachate treatment system, with positive results. The upgrade was initiated in 2001 and completed in October 2002. The MBR system was expanded to 42 MLD capacity through the installation of additional membrane cassettes after about one year of operation.

The MBR system, which is protected by a fine screen upstream, consists of a denitrification tank followed by one nitrification tank which precedes four independent filtration trains. Each train is capable of producing up to 438 m^3/h of permeate, yielding an overall installed capacity of 42 MLD from a total membrane area of 73,442 m^2 (Table 5.12) for which the design flux is 24 LMH. In each train permeate is withdrawn from two membrane tanks (subtrains), each tank containing 20 cassettes of 22 modules of 20.8 m^2 each. Fine bubble aeration takes place at between 6000 and 11,000 Nm^3/h depending on the organic loading. An MLSS concentration of 6–9 g/L in the membrane

TABLE 5.12 Membrane Process Components

Item	Value
No. trains	4
No. subtrains per train	2
No. cassettes per subtrain	20
No. modules per cassette	22
Area per module, m^2	20.8
Total surface area, m^2	73,442

compartment is maintained by an SRT which varies between 11 and 24 days with an average of 17 days and an HRT of 8–9 h.

The plant is equipped with fully automated operation and maintenance (O&M) cleaning. It is operated with a filtration cycle of 800 s, with 715 s production and 85 s relaxation: no backflushing is employed at this plant. The plant originally operated with cyclic aeration, with the aeration alternating between adjacent banks of 10 cassettes. Each subtrain has two separate air headers serving 10 cassettes (one bank) each; in total there are four membrane banks per train for a total of 40 cassettes. In the '10/10' aeration mode of operation, employed from start-up until July 2009, two banks per train were aerated at constant airflow for 10 s in a 20 s cycle, yielding a SAD_m of 0.253 $Nm^3/(m^2\,h)$. In July 2009 '10/30' (*Eco aeration*) was implemented and airflow was reduced by a further 50% ($SAD_m = 0.127\ Nm^3/(m^2\,h)$) by aerating only one bank of membranes per train at any time for 10 s over a 40 s cycle. The reduced aeration was made possible by the implementation of an on-line fouling controller based on resistance in series (Ginzburg, Peeters, & Pawloski, 2008), which involves calculation of the total resistance within the permeation cycle by interpolation of data (Figs 5.30 and 5.31) to evaluate initial (membrane) and cake resistances. Cyclic aeration, 10/10 or 10/30, is then actuated on the basis of the fouling rate. From May 2009 until February 2010 the membrane operation energy demand has been recorded for both 10/10 (May 2009–July 2009) and 10/30 (August 2009–February 2010) cyclic modes. Results (Fig. 5.32) indicate a 25% reduction in the membrane operation specific energy demand from switching from 10/10 to 10/30 aeration.

Maintenance cleaning in situ is conducted weekly using 500 mg/L NaOCl at pH 8–8.5 and with 2 g/L citric acid. The citric clean was introduced following the commencement of ferric addition in January 2009 due to an imposed more stringent TP effluent requirement. The chemical is back-pulsed slowly through the membrane, whilst still immersed in the mixed liquor, for a period of 45 min.

To date, there have been no operational problems with this plant or membrane replacement after almost eight years. Normalized permeability for

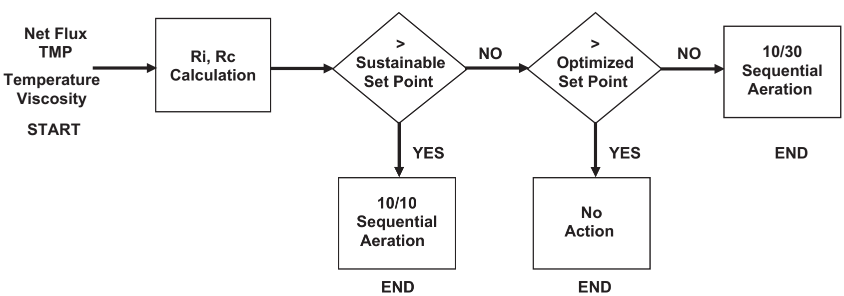

FIG. 5.30 On-line fouling monitoring concept for aeration control at Brescia.

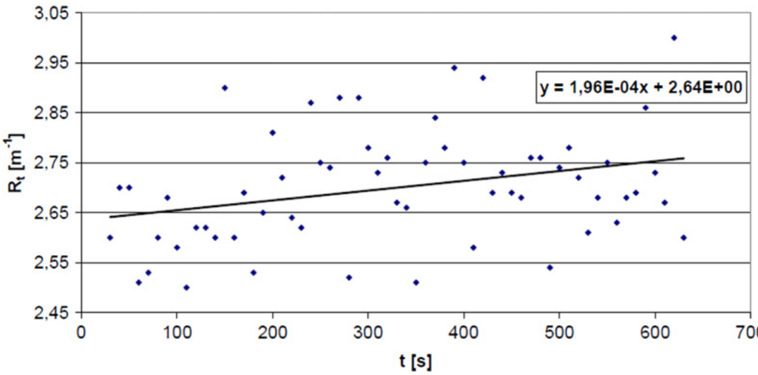

FIG. 5.31 Example of total resistances interpolation for Brescia Train #4, net flux = 21.3 LMH, MLSS = 8.6 g/L, T = 14 °C.

all trains varies between 130 and 160 LMH/bar and net flux from 20 to 24 LMH. Effluent quality has met or exceeded expectations for SS, BOD, ammonia-N and TN. The high-quality MBR effluent is blended with effluent from the conventional Trains A and C, allowing the overall plant to meet present discharge standards. The MBR effluent quality is also sufficiently high quality to be considered for reuse as irrigation water.

5.3.1.4. Traverse City

Client: Traverse City, MI, USA
Primary contractor: CH2M HILL
Primary consultant: CH2M HILL

Traverse City is a community of about 15,000, well known for the natural beauty surrounding the city which contributes to a well-developed round-the-year tourism and recreation industry. The wastewater treatment works (Fig. 5.33) also serves parts of Grand Traverse and Leelanau Counties, such that the overall population equivalent is close to 50,000. MBR technology was selected for the site following extensive public consultation which revealed widespread interest in providing effluent of as high a quality as would be reasonably possible. The public was also decidedly uninterested in increasing the footprint of the existing works, which is bounded on two sides by Boardman Lake, on another side by the Pere Marquette railroad, and on the fourth side by a recently improved city park. The conversion of the plant to an MBR was part of a project which took place over a two-and-a-half year period at a cost of $31m, increasing the plant's capacity by 42%, with respect to BOD loading, while increasing effluent quality. The 32 MLD maximum monthly flow (64 MLD peak hourly flow) plant began producing MBR permeate in July 2004.

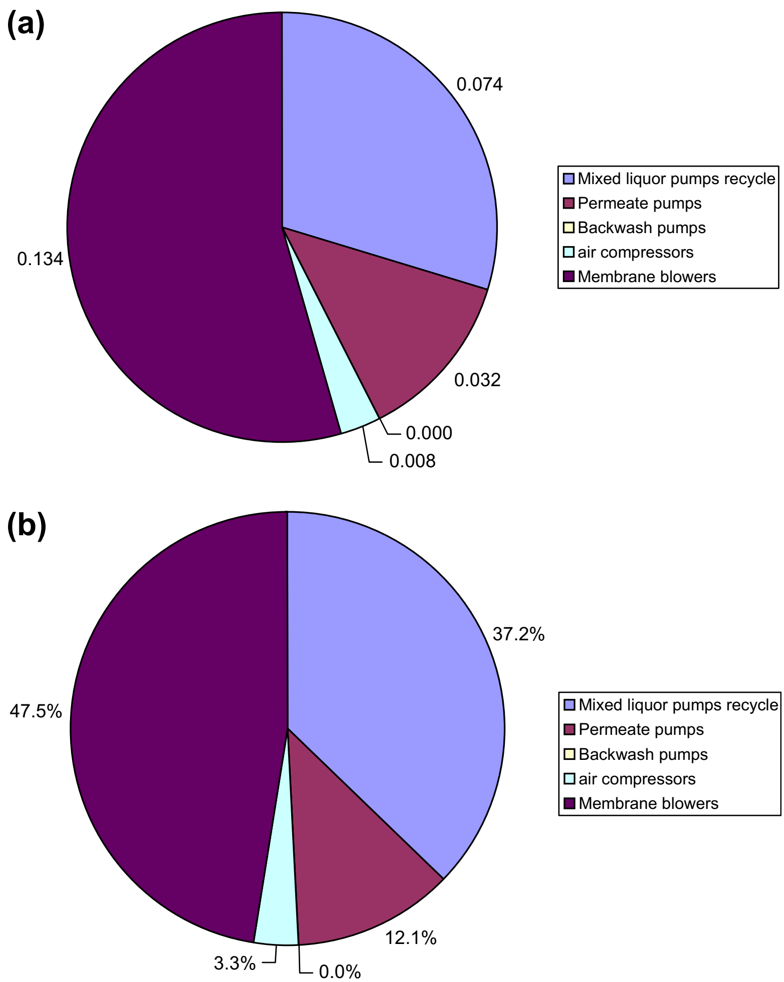

FIG. 5.32 Membrane operation energy demand at Brescia: (a) 10/10 aeration, May–Jul 2009, 0.248 kWh/m^3 total and (b) 10/30 aeration >93% of the time Jul 2009–Feb 2010, 0.187 kWh/m^3 total.

Influent to the works consists of wastewater from Traverse City and six surrounding townships, and treated effluent from Grand Traverse County Septage Treatment Facility. It passes through a 6 mm screen, then detritor grit removal systems. The screenings and grit are disposed of in a sanitary landfill. Wastewater then flows through primary sedimentation tanks (PSTs). Sludge from the PSTs is directed to anaerobic digesters. A portion of the PST underflow, however, can be returned to the plant influent to facilitate liberation of volatile fatty acid (VFA) generated under the anaerobic conditions prevailing in

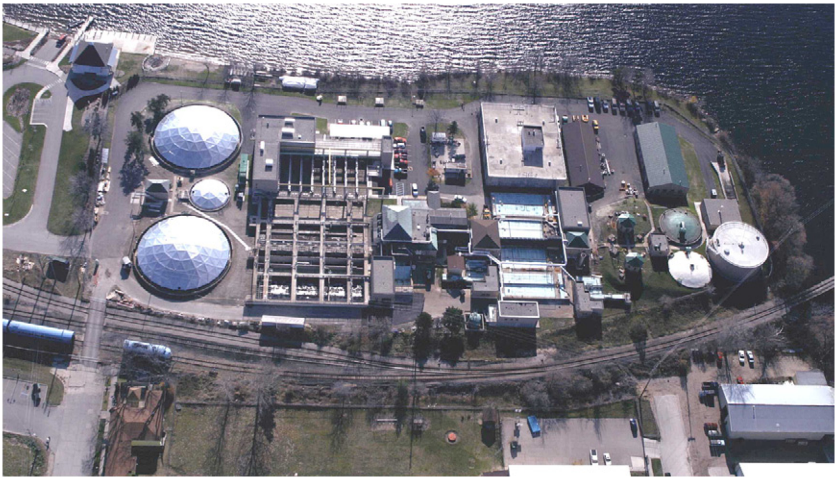

FIG. 5.33 Traverse City WwTW.

the PST sludge blanket. The PST thus can also serve as a fermenter to promote biological phosphorus removal in the downstream aerobic process.

Following screening to 2 mm using perforated plate screens, the primary effluent is transferred by screw pumps to the bioreactor basins. These comprise anaerobic, anoxic and aerobic zones with recirculation of mixed liquor from the aerobic to the anoxic zones, at a recycle ratio of 1, and from the anoxic to the anaerobic zones, also at a recycle ratio of 1. Ferric chloride can be added to the mixed liquor to precipitate phosphate as necessary to supplement its biological removal and reduce P levels in the plant effluent. Mixed liquor flows from the aeration basins to the membrane vessels and is recycled from the downstream end of the membrane vessels back to the aerated zone of the aeration basins at a recycle ratio of 4. Recycle flows for the plant are summarized in Fig. 5.34.

The membrane vessels (Fig. 5.35) comprise eight trains each containing 13 cassettes of 32 *ZW500c* modules, thereby giving a total installed membrane area of 68,091 m^2. The membranes are typically operated on a standard 10 s on/30 s off cyclic aeration mode at a specific aeration demand of 0.18 $m^3/(m^2\,h)$,

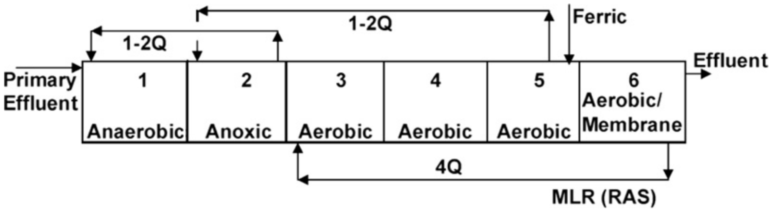

FIG. 5.34 Recycle flows for the Traverse City plant (Crawford, Daigger, & Erdal, 2006).

FIG. 5.35 The MBR at Traverse City WwTW, showing the permeate and air headers.

with relaxation for 40 s in every 12 min of operation. Maintenance cleaning using 200 mg/L hypochlorite is carried out every 10 days, supplemented by a clean with 2000 mg/L citric acid every 20 days. This schedule maintains a net average flux of 19.2 LMH at the MLSS concentration of 4−5 g/L (6−10 g/L in the membrane basins). The permeate flows through an ultraviolet disinfection system, supplied by Infilco Degremont, and is discharged as effluent to the Boardman River.

The plant has on-site solids processing, with half of the holding tank volume provided by conversion of the now out-of-service secondary clarifiers. WAS is taken from the point where the RAS exits the membrane vessels. It is dosed with polymer coagulant and concentrated in a gravity belt concentrator. Ferric chloride can be applied to the WAS to fix the phosphorus chemically, thus preventing its reintroduction into the wastewater through sidestream flows. It can be dosed either upstream or downstream of the concentrator. The concentrated waste activated and primary sludges are anaerobically digested, the sludge residue dosed with polymer and the solids further concentrated in sieve drum concentrators prior to storage in tanks with a total capacity exceeding 11,000 m^3. Around 15,000 m^3 sludge (800 tonnes dry solids) is generated annually, and is sub-surface injected into local farmland.

The target effluent quality parameters set were 4 mg/L BOD, 4 mg/L TSS, 1 mg/L N-NH_4 and 0.5 mg/L TP, significantly lower than the statutory requirements of 25, 30, 11 and 1 for these respective determinants. The actual effluent quality achieved during plant performance testing has nonetheless exceeded these standards, with non-detectable BOD and TSS and less than 0.5 mg/L N-NH_4 and TP levels of 0.05–0.4 mg/L.

5.3.1.5. Peoria

Client: City of Peoria, AZ, USA
Primary contractor: Sundt Construction, Inc.
Primary consultant: Black & Veatch

The water reclamation facility (WRF) at Butler Drive in Peoria, Arizona was constructed on a greenfield site (Fig. 5.36). It has a capacity of 47 MLD maximum monthly flow and a 91-MLD peak hourly flow, equating to a peak loading factor of 2.4 over the average daily flow of 38 MLD. MBR technology was selected to provide effluent of sufficiently high quality (monthly average concentrations of <0.5 NTU, <5 mg/L cBOD, <8 mg/L TN and non-detectable FCs) for aquifer recharge. The consultant for the project was Black & Veatch and the plant began operation in June 2008.

Wastewater is screened to 13 mm, using a Duperon Flex Rake, at the remote pumping station and transferred to the aeration tank. There is no buffer tank, with some equalization provided through an additional 0.6 m depth to the aeration and membrane tanks. This provides around 1.5 h of buffering at the average daily flow, with the majority of the peaking processed through the membranes. Following degritting the water is screened through 2 mm punch-hole plates with an EIMCO/Brackett and Green internally fed travelling band screen before entering the bioreactor basins. The bioreactor has anoxic and aerobic zones, of 3000 and 14,000 m^3, respectively, with a recycle ratio of 3 for the transfer of mixed liquor from the membrane to the anoxic zone. The MLSS is approximately 8 g/L in the aerobic zone and 10–12 g/L in the membrane zone. The WAS, discharged at a rate of around 12 tonnes/day, is dewatered on-site, consuming an average of 700 L of ferric chloride (for H_2S control) and 340 L of polymer. About 65 tonnes/day of 18–20 wt% DS sludge is generated from the site.

The membrane basins, 1400 m^3 in total volume, house the 60 cassettes containing 48 *ZW500d* modules (31.6 m^2 and 1517 m^2 per module and per

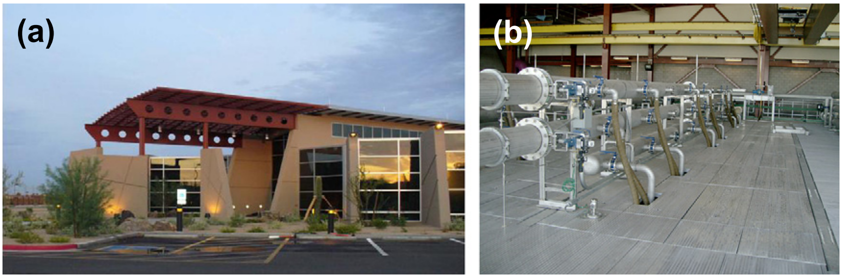

FIG. 5.36 The WRF at Peoria: (a) plant entrance and (b) membrane tanks.

cassette, respectively) distributed between 10 trains (and hence six cassettes per train). The membranes operate on a filtration cycle of 12 min, including 45 s of relaxation. Scouring air is typically applied on a cycle of 10 s on/30 s off at an overall SAD_m of approximately 0.2 $Nm^3/(m^2\ h)$ which, at the design flux of 17.4 LMH, yields a SAD_p of approximately 11.5. The plant has thus far operated at a flux of 20 LMH, due to the current flow of wastewater. The maximum operational TMP is rated for 0.55 bar, but typically operates at 0.03−0.10 bar. Membranes are maintenance-cleaned weekly with 200 mg/L hypochlorite, incurring a 1 h downtime, and the plant has not used citric acid maintenance cleans. The current planning is for an anticipated yearly recovery clean, once each with 1 g/L sodium hypochlorite and 2 g/L citric acid. About 30 kg of NaOCl is used per maintenance clean.

The plant routinely exceeds the product water quality objectives, with less than 3 mg/L TN. The mean membrane and total aeration energy demand for the plant are around 0.34 and 0.55 kWh/m^3, respectively.

5.3.1.6. Ulu Pandan

Client: Public Utilities Board
Primary contractor: Dayen Environmental Limited, Singapore
Primary consultant: PUB Consultants Private Limited

The MBR plant installed on the site of the Ulu Pandan Water Reclamation Plant (WRP) in Singapore forms part of the water reuse technology demonstration programme undertaken by the Singaporean Public Utilities Board (Fig 5.37). The programme had led to the pilot-scale demonstration of a number of MBR technologies from early 2003 onwards (Section 3.2.1.3), the results of which were sufficiently encouraging to sanction the 23 MLD demonstration plant, which was commissioned in December 2006.

This MBR (Fig 5.37) is retrofitted to the existing WRP and is fed from the settled sewage tank, which receives wastewater of roughly 90% domestic and 10% industrial origin (Table 5.13). Two of the existing biological trains were converted to provide anoxic and aeration zones (volume ratio 2:3), to which the settled sewage flows under gravity along with the flow from the membrane tank. The mixed liquor is pumped from the aerobic to the membrane tank via

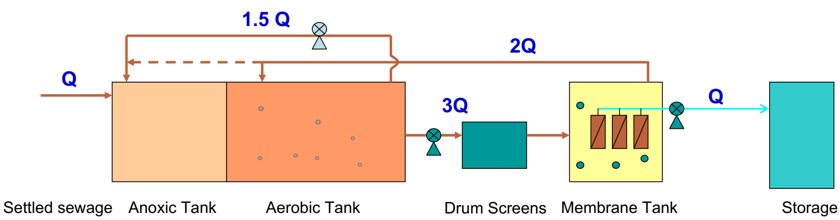

FIG. 5.37 The Ulu Pandan MBR demonstration plant.

TABLE 5.13 Settled Sewage and Product Water Characteristics at Ulu Pandan

Parameter	Units	Average	Range	Permeate
BOD_5	mg/L	138	111–171	
COD	mg/L	292	236–420	
TOC	mg/L			4.8
TSS	mg/L	105	89–120	
Turbidity	NTU			0.02
TKN	mg/L	47.6	36.7–61.8	
NH_4-N	mg/L	32	20.5–46.6	
NO_3-N	mg/L			6.3
Alkalinity	mg L^{-1}as $CaCO_3$	166	95–202	
Coliforms	CFU 100 ml^{-1}			<1
Mixed Liquor Temperature	°C	30	28–32	
Total phosphate as P	mg/L	6.7	5.2–8.1	3.3
pH			6–8	

2 mm mesh drum screens. The bioreactors operate at a maximum MLSS of 10 g/L, a minimum sludge age of 10 days, an HRT of 6 h at the constant 23 MLD flow and a minimum *F*:*M* ratio of 0.1 kg BOD/(kg MLVSS d). The membrane tank has five trains each with five *ZW500c* cassettes providing a total membrane area of 37,920 m^2. At the net design flux of 25 LMH the maximum MLSS in the membrane tank is 12 g/L. The MBR system performance is controlled and monitored on-line by a SCADA (supervisory control and data acquisition) system, and there are regular laboratory analyses of water quality.

Membrane operation comprises 12 min of filtration followed by 30 s of backflushing, after every 10 cycles, or relaxation for the remaining cycles. Maintenance cleaning is twice weekly with 200 mg/L sodium hypochlorite. Recovery cleaning with 500 mg/L hypochlorite followed by 1000 mg/L citric acid has been required only four times in the first three years of operation; the overall membrane fouling rate at the plant is less than 0.15 bar per year.

The operators have devoted much effort to optimizing the MBR for energy efficiency with respect to MLSS level, MLSS recirculation rate, and process and membrane aeration (Tao et al., 2009). This was achieved partly by installing variable frequency drives for the blowers and pumps. An overall baseline energy demand of 0.59 kWh/m^3 was then found to be attainable at 10 g/L MLSS, with the process and membrane blowers each contributing 32% to the overall energy demand and sludge transfer another 17%. Halving the MLSS level was found to reduce the energy demand by a further 7% to 0.549 kWh/m^3. Both reducing the recirculation ratio, from the baseline value of 2, and redirecting the RAS stream to the aerobic region of the tank were found to be counterproductive.

On the other hand, the incorporation of on-line ammonia nitrogen and TOC meters to monitor biotreatment efficacy, rather than using the 1.5 mg/L set point for dissolved oxygen, permitted the gradual reduction of process aeration to 60% of the design value without compromising the product quality. This led to energy consumption further reduced to 0.475 kWh/m. Finally, reduction of the membrane scouring air from a cycle of 10 s on:10 s off to 10 s on:30 s off was found to have the greatest impact on the overall energy demand, reducing it to 0.4 kWh/m^3. Such intermittent aeration is compromised in the case of Ulu Pandan by there being five trains: 10:30 *Eco aeration* relies on the number of trains being divisible by four, such that four trains are aerated 25% of the time by switching the scouring air between them sequentially. One train was therefore maintained with 10:10 aeration.

The optimization of operating conditions at the Ulu Pandan plant has had no detrimental effect on product water quality or membrane fouling, and demonstrates the extent to which the energy demand for an MBR plant can be reduced whilst still maintaining production.

5.3.1.7. Basic American Foods

Primary contractor: Zenon Environmental Corp.
Primary consultant: Donohue & Associates

The MBR plant at Basic American Foods in Blackfoot, Idaho treats effluent from a potato processing plant, the site of possibly the first potato dehydration facility in the USA. An MBR was selected to enable the wastewater to meet the water quality objectives demanded by its reuse for crop irrigation, principally nitrogen removal. The plant is designed to treat a maximum daily flow of 6.4 MLD and was commissioned in December 2002.

Screening of the wastewater is through two pairs of shaker tables, one pair rated 3.4 and the other 1.7 mm. The effluent is then passed through a primary clarifier before entering the bioreactor, which comprises a 9100 m^3 aerobic zone and a 4550 m^3 anoxic zone which provide a combined HRT of over 50 h at peak flow. The MLSS concentration in the aerobic and membrane tanks is 8−12 and 10−15 g/L, respectively, with a 10:1 recycle ratio for the anoxic zone. The biological treatment process is not supplemented with additives but the groundwater is naturally high in alkalinity, which assists in nitrification. The wastewater temperature ranges from 20 to 40 °C.

The membrane filtration system comprises three trains each containing 10 cassettes of 22 *ZW500c* modules, providing a total membrane surface area of

15,330 m^2. The operation cycle is 12 min filtration followed by 35 s of backwash to sustain a design flux of 17.5 LMH. Membrane aeration at an instantaneous rate of 3300 Nm^3/h is applied with a cyclic pattern of 10 s on/20 s off to each membrane train. Maintenance cleaning is carried out weekly by applying 100 mg/L of sodium hypochlorite for 60 min. In the seven years of operation, recovery cleaning has been limited to five cleaning events with 0.2 wt% citric acid.

The plant has operated consistently well, achieving an effluent COD concentration of <15 mg/L (95% removal) and <5 mg/L TN.

5.3.2. Asahi Kasei

5.3.2.1. WenYuHe (Wenyu river)

Client: Beijing Water Affairs Bureau of Shunyi District
Primary contractor: Beijing Origin Water
Primary consultant: Beijing Institute of Water

The plant on the Wenyu River (Figs 5.38 and 5.39) in the Shunyi district is one of the largest MBR plants in the world at a capacity of 100 MLD ADF in the summer, decreasing to 25 MLD in the winter when the feed temperature drops to near-freezing. It was constructed by Beijing Origin Water, originally to provide recycled water to the Chaobai River. The plant is, in effect, a river purification system: it is required to produce effluent of sufficient quality to meet Class III standard of the statutory environmental quality standards for surface water (GB3838/2002). Water quality standards for China are based on five classes of water, ranging from Class I for a 'National Nature Reserve' to Class V for irrigation; Class III surface water is that designated as a centralized drinking water source, as well as for fishing and bathing (Table 5.14). The MBR option was chosen primarily because of the high quality effluent generated, combined with the low footprint incurred. Installation of the plant was preceded by 10 months of sampling and flow-charting by BOW, and the plant went on-line in October 2007.

Pre-treatment comprises a 20-mm bar screen followed by a 700,000 m^3 equalization tank, rebuilt from a natural basin close to the river, and then a 3-mm fine screen upstream of the inlet pumps. There follows classical clarification with dosing to 20–30 mg/L with PACl in the sedimentation tank before the water enters the MBR. Four parallel channels are provided following clarification, each channel consisting of an anoxic tank and aeration tank (Fig. 5.38b), with recirculation at a ratio of 2–4 depending on the influent ammoniacal nitrogen concentration. The membrane modules (Fig. 5.38c) are immersed directly in the nitrification tank, where further PACl dosing takes place at 10–20 mg/L. The MLSS concentration is between 4.5 and 6 g/L at the SRT of 15–40 days in the summer and 80–100 days in the winter, depending on the loading. The low winter temperatures mean that the nitrification can decrease to 80% in the winter (Yu et al., 2009).

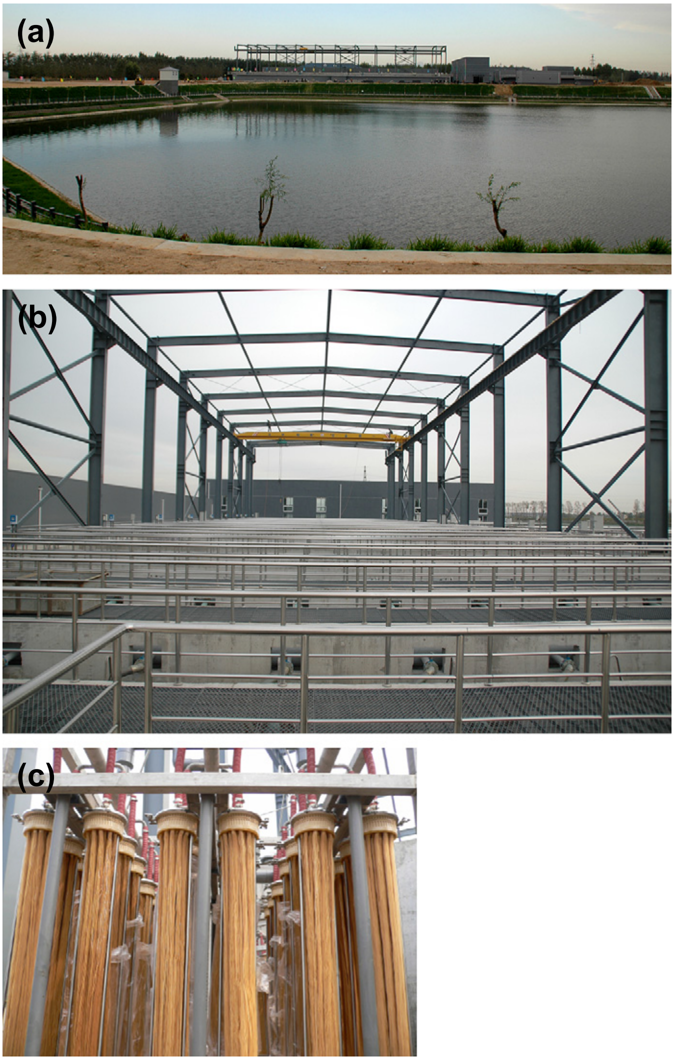

FIG. 5.38 The Wenyu River plant: (a) a view from across the river, (b) membrane tanks and (c) the membrane modules.

The total membrane area of 180,000 m^2 is provided by 20 trains of 10 stacks containing 36 modules of 25 m^2 area. The TMP is maintained between 0.1 and 0.3 bar for much of the time, with the flux reaching 20 LMH in the summer compared to the lowest attainable flux of 5 LMH in the winter. Filtration is on a 10 min cycle, with 9 min filtration and 60 s backflushing at about 24 LMH. The TMP is between 0.1 and 0.3 bar, and the SAD_m 0.2–0.28 $Nm^3/(m^2\,h)$. At the unusually low loadings at which the plant operates (Table 5.14) this

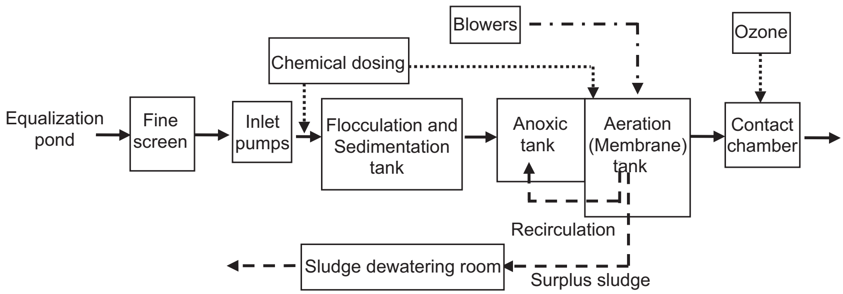

FIG. 5.39 Schematic of the Wenyu River plant (Yu et al., 2009).

provides sufficient air for biotreatment as well as membrane scouring, such that the overall specific energy demand is one of the lowest for any MBR plant at only 0.274 kWh/m^3 − almost 90% of which is for aeration. The membrane undergoes a weekly maintenance clean with 500 mg/L NaOCl solution, supplemented with a monthly clean with 3000 mg/L NaOCl and a quarterly clean with 1.5 wt% citric acid, with a downtime of 2−3 h in each case. Annual 24 h recovery cleans using 3000 mg/L NaOCl and 2 wt% citric acid can also be employed when necessary.

The plant provides a valuable illustration of the impact of feedwater temperature on design flow. The flow capacity is determined largely by the water temperature, since the peak loading is very small because the normal diurnal changes in flow which apply to municipal sewage do not apply. At the minimum mixed liquor temperature of 3 °C in the winter season the viscosity is almost exactly double that at the maximum temperature in the summer at 29 °C. The operating flux, on the other hand, differs by a factor of 4. This suggests that the correction of flux or permeability based on viscosity may be inappropriate.

5.3.2.2. Hamamatsu City, Shironishi WwTP

Client: Hamamatsu City
Primary contractor: Kobelco Eco-Solutions Co., Ltd
Primary consultant: Japan Sewage Works Agency

TABLE 5.14 Design Feed and treated Water Quality (mg/L), Wenyu River Plant

Stream	COD	BOD_5	NH_4^+-N	TN	TP
Feedwater	81.7	13.4	25.1	32.5	2.7
Design treated water	20	4	1.0	15	0.2

FIG. 5.40 The MBR plant at Hamamatsu City.

The 1.4 MLD PDF plant at Hamamatsu City has been operating since March 2008. MBR technology was selected primarily on the basis of the site footprint limitations (Fig. 5.40). The municipal wastewater flows to the MBR via a 5.5 h equalization tank and a 1 mm fine bar screen. The total bioreactor volume is 365 m^3 (HRT: 6.4 h at a design capacity). The bioreactor is evenly divided between the anoxic and aerobic tanks with internal recycle of nitrified mixed liquor. The designed average MLSS is 10 g/L, and the designed lowest temperature is 8 °C.

Two trains of five membrane racks, each consisting of 12 membrane modules, are submerged in the aerobic tank. The total membrane surface is 3000 m^2. The membrane operating cycle consists of 9 min filtration and 1 min backflush at 24 LMH. The designed net flux is 19 LMH at a mean TMP of 0.15 bar. Scouring air is supplied at a rate of 0.20 $Nm^3/(m^2\ h)$, corresponding to the SAD_p value of 10.5. Maintenance cleaning is carried out with 3 g/L NaOCl, incurring 1.5 h of downtime. The effluent quality has been satisfactory, BOD and T-N concentrations being <2 and <10 mg/L, respectively and <10 CFU enterococci per 100 mL.

5.3.2.3. Machida Food Factory, Fuji City

Primary contractor: Asahi Kasei Chemicals Co.
Primary consultant: Asahi Kasei Chemicals Co.

The 0.9 MLD ADF (1.0 PDF and 1.5 peak hourly flow) MBR at the Machida Food factory has been treating food processing effluent since July 2004. The wastewater flows to the MBR through a 2 mm vibrating screen. The plant is for nitrification and carbonaceous organic matter removal only, with a single 950 m^3 aerobic tank (hence ~1 day HRT) which also contains the membrane modules; the mean MLSS is 10 g/L and 300 kg DS of WAS is generated daily. The

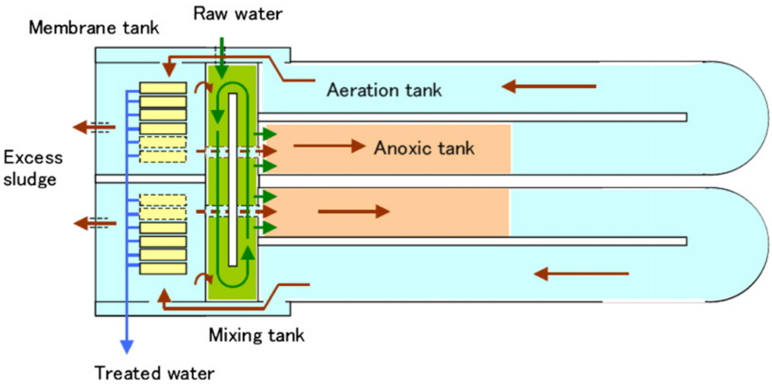

FIG. 5.41 Plant design for Hainan and Daya Bay.

membrane tank has two trains of 12 racks, each containing four membrane modules, to give 2400 m^2 of membrane area. The operational cycle comprises 9 min of filtration and 1 min of backflushing at 16 LMH, sustaining a net flux of 16 LMH and a mean TMP of 0.15 bar. Thus, at the standard SAD_m of 0.20 $Nm^3/(m^2\,h)$ the SAD_p value is 12.5. The membranes are maintenance cleaned every two weeks with 1000 mg/L NaOCl for 1.5 h, and recovery cleans using 3 g/L NaOCl or 1 wt% oxalic acid employed every two years. The treated effluent is <5 mg/L BOD_5, <30 mg/L COD and has non-detectable ammonia. The plant was still operating with the original membranes after five-and-a-half years.

5.3.2.4. Other Asahi Kasei Plants: Petrochemical Wastewater

There are a number of Asahi Kasei reference sites in China and Japan, and these include two significant petrochemical effluent treatment plants on the South East coast of China, both operating at the same design mean flux of around 12 LMH. The design (Fig. 5.41) is similar to that of the Wenyu River plant, with buffering and an anoxic zone upstream of the aerobic tank, but with additional anaerobic treatment. The 10.8 MLD Hainan plant was installed in August 2006, and the larger 25 MLD plant at Daya Bay five months later in January 2007. Daya Bay achieves 9% COD removal and 88% nitrification, but also removes almost 98% of the mineral oil and over 99% of the phenol at mean feed concentrations of 57 and 27 mg/L, respectively.

5.3.3. Beijing Origin Water

5.3.3.1. ShenDingHe (Shending River)

Client: Shiyan Wastewater Treatment Company
Primary contractor: Beijing Origin Water
Primary consultant: Central and Southern China Municipal Engineering Design & Research Institute

FIG. 5.42 The site at the Shending River.

The MBR plant at the Shending River at Shiyan, Hubei province was the largest in the world at the time it was commissioned in November 2009, with an ADF of 110 MLD and a peak hourly flow of 143 MLD. The MBR plant (Fig. 5.42) is an upgrade of the existing works, and the MBR technology was selected on the basis of the limited available space and the high quality (low nutrient) effluent demanded.

The wastewater flows through sequential 20 mm and 5 mm chain-rake screens prior to grit removal in an aerated channel, and then fine screening to 1 mm with a mesh drum screen. Biotreatment comprises anoxic, anaerobic and aerobic treatments in tanks of 10,600, 7090 and 21,700 m^3, respectively, with recirculation through the 8400 m^3 membrane tanks at a recycle ratio of 2–3 depending on the influent N load. The MLSS is maintained at 7.5 g/L in the biotank and 9 g/L in the membrane tank, with the WAS flow being 16.4 tonnes DS/day (and hence ~22 d SRT). 20–30 mg/L PACl is dosed between the biological and membrane tank to provide chemical P removal.

The membrane tanks are fitted with a total of 171,720 m^2 (from 18 streams of six stacks) of membrane area with each stack providing 1590 m^2 from 60 of the 26.5 m^2 *BSY* modules. The mean design flux of 25 LMH is maintained by air scouring at 0.26–0.31 $Nm^3/(m^2\,h)$, 0.28 on average (corresponding to a mean SAD_p of 11.2), and the TMP ranges from 0.1 to 0.25 bar under these operating conditions. Maintenance cleaning is with a weekly 2-h 500 mg/L NaOCl CIP.

The plant operates at energy demands of 0.24, 0.4, 0.04 and 0.04 kWh/m^3 for the membrane aeration, total aeration, sludge transfer and permeation, respectively; giving a total specific energy demand of 0.48 kWh/m^3: 83% of the energy demand is for aeration, more than half of this being for membrane scouring. The plant removes more than 96% of the influent BOD_5 and ammonia (down to <6 and <1 mg/L, respectively), and provides residual TN and TP levels of <15 and <0.3 mg/L, respectively.

5.3.4. Korea Membrane Separation (KMS)

5.3.4.1. Dalsung

Client: Environmental Management Corporation (EMC)
Primary contractor: SsangYong Engineering and Construction Co., Ltd
Primary consultant: Hanjo Engineering Co., Ltd
Membrane supplier: KMS (Korea Membrane Separation) Co., Ltd

The MBR plant at Dalsung (Figs 5.43 and 5.44) is a retrofit on an existing conventional works which prior to the retrofit comprised: coarse screening

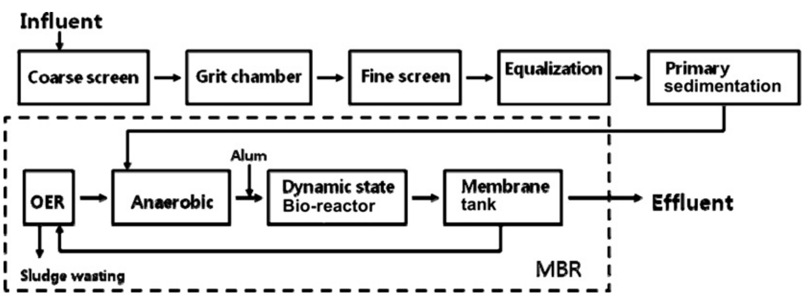

FIG. 5.43 The treatment scheme at Dalsung.

FIG. 5.44 The Dalsung WwTW site.

(6 mm), degritting, flow equalization (4.5 h HRT at peak daily flow), primary sedimentation (2.8 h HRT at peak daily flow), fine screening (1.5 mm), biotreatment, secondary sedimentation and then tertiary sedimentation with alum dosing. Secondary and tertiary sedimentation have since been obviated by the retrofit. The influent is around 70% industrial effluent from nearby industrial activities which include food processing, steel production, textile dyeing and paper manufacture, with the remaining 30% being of municipal origin. The 25 MLD plant (peak daily flow) provides 15 MLD of water for direct reuse by two of the paper production companies, with the remaining water being discharged into the nearby Nak-dong River. MBR technology was selected for improved treated water quality and footprint constraints. The plant is the largest example of the *KSMBR* (Section 4.3.8) and began operation in June 2009.

The plant is designed for biological nutrient removal, with a recycle ratio between 0.6 and 1.0 to the anaerobic zone. The process consists of an anaerobic tank (1728 m^3), two 'dynamic state' bioreactors with a combined volume of 2773 m^3, an aerobic membrane tank of 2830 m^3 and a 583 m^3 'oxygen exhausted reactor' (OER) for de-oxygenating. The total tank volume of 7914 m^3 provides an HRT of 7.6 h at the peak daily flow. The dynamic state bioreactors are aerated intermittently with, at any one time, one reactor operating continuously under anoxic conditions and the other operating in batch mode with anoxic—oxic—anoxic conditions — three shifts over a 1-h cycle. This achieves simultaneous nitrification and denitrification. The OER reduces the DO in the return activated sludge to below 0.2 mg/L in the OER tank before entering the anaerobic tank. Alum dosing at the dynamic state bioreactor inlet has been installed to provide supplementary chemical P removal in the event of unusually high P loadings. A WAS stream of 178 m^3/day is discharged from the membrane tank, which is maintained at 8 g/L MLSS.

Membrane separation is based on the *KMS 6007CF* stack and has displaced all the unit operations downstream of the primary sedimentation tank. The plant has two trains of 22 stacks, each containing 112 cartridges of 18 m^2 in an array which is 7 cartridges high, 8 long and 2 deep (Fig. 5.45), providing a total membrane area of 88,704 m^2. Filtration is on a 12 min on, 3 min relaxation basis, with no backflushing. Scouring air is provided at a rate of 0.12—0.16 $Nm^3/(m^2\,h)$, or 0.145 on average, which sustains a mean flux of 12.5 LMH within a TMP range of 0.065 and 0.53 bar. Maintenance cleans comprising a 2-h CIP using 2 g/L NaOCl are applied every two weeks, and recovery cleans with 4 g/L NaOCl anticipated twice yearly.

The plant operates at a total energy demand of around 0.7 kWh/m^3 with 0.343, 0.027 and 0.086 kWh/m^3 for aeration, sludge pumping and permeation, respectively, and thus far the plant has exceeded the discharge consents (Table 5.15).

FIG. 5.45 A single 112-cartridge frame being installed at Dalsung.

TABLE 5.15 Actual and Consented Permeate Water Quality, Dalsung

	Permeate*, mg/L	% Removal*	Consent
TSS, mg/L	0.4	98.9	10
BOD_5, mg/L	0.7	98.3	3.7
COD, mg/L	9.1	79.8	–
Ammonia, as N	2.8	55	–
Total nitrogen, as N	6.8	62.3	20
Total phosphorus, as P	0.97	70.7	2

**Nov–Dec 2009.*

5.3.4.2. Okchun

Client: Environmental Management Corporation (EMC)
Primary contractor: SsangYong Engineering and Construction Co., Ltd
Primary consultant: Hanjo Engineering Co., Ltd
Membrane supplier: KMS (Korea Membrane Separation) Co., Ltd

The background to the plant at Okchun is not dissimilar to that of Dalsung (Section 5.3.4.1). MBR technology was selected for improved discharged water quality and increased capacity given space limitations on site, and the process displaced secondary and tertiary treatment processes on site whilst primary sedimentation, providing 2.1 h HRT on average, was retained. Three 0.5 mm fine mesh drum screens have been fitted upstream of the MBR to provide protection of the membranes from clogging, and the effluent is discharged to the So-ok stream near the plant. The plant has an 18 MLD capacity (21.6 MLD peak hourly flow) and the first phase has been operational since September 2008 and the second since July 2009. A third phase has been operational since April 2010.

Biotreatment is based on the *KSMBR* process, which provides biological nutrient removal, whose design and operation are as described for Dalsung. The total tank volume of 4287 m^3 (and hence 5.7 h HRT) comprises anaerobic, dynamic bioreactor and oxygen exhaustion tank volumes of 552, 1737 and 283 m^3, respectively, with the membrane tank being 1714 m^3. The 1424 kg/day DS discharged sludge (equating to 24 d SRT) is dewatered on site by ferric dosing.

The MBR employs the *KMS 7206 CF* unit and provides 72,500 m^2 membrane area from six trains of five stacks, each containing 144 cartridges in a 6 high × 12 long × 2 deep array. Membrane operation is based on 12 min filtration and 3 min relaxation. Scouring air is provided at 0.2–0.232 $Nm^3/(m^2\,h)$, 0.212 on average, which sustains a mean flux of 10.3 LMH (and hence a SAD_p of 21) within a TMP range of 0.065 and 0.53 bar. Chemical cleans are as described for Dalsung. The plant operates at a total energy demand of around 0.52 kWh/m^3 with 0.39 and 0.44 kWh/m^3 for membrane and total aeration, respectively, and with a further 0.05 and 0.03 kWh/m^3, respectively, for sludge transfer and permeate extraction. The plant is subject to the same discharge water quality limits as Dalsung and produces a permeate water of similar characteristics.

5.3.4.3. Other KMS Plants

KMS has been a major player in the Korean MBR market since 2000, with over 1000 installations in total (albeit most of them less than 0.5 MLD in capacity). Although their product is based on one of the less favoured materials (high-density polyethylene), evidence suggests that it is reasonably robust. The oldest KMS plant was installed in 2003 and is still operating with its original membranes.

5.3.5. Koch Membrane Systems – Puron

5.3.5.1. *Sobelgra*

Client: Boortmalt
Primary contractor: Veolia Water Solutions & Technologies

The largest industrial MBR plant at Sobelgra in Belgium was commissioned in November 2004, and the site is located in Antwerp harbour. The MBR plant has a capacity of 2 MLD and treats effluent from a malting operation, where barley is converted to malt for beer by a natural enzymatic process. Malting effluent is particularly challenging, being very high in organic content and with a relatively low BOD/COD ratio (Table 5.16). The MBR plant was selected as a result of the plant capacity being extended from 110 to 250 ktonnes p.a., making it the largest independent malting company in Belgium. This meant that the capacity of the existing conventional WwTP had to be doubled and, due to lack of space on the factory site, conventional technology could not be used. The MBR plant (Fig. 5.46) was piloted in the Spring of 2003 and installed by the Belgian Engineering Company SEE:WATER (now Veolia Water Solutions & Technologies) supplying and assembling the MF (microfiltration) system.

The MBR was retrofitted into the existing aeration tank, which was divided into two parts to provide an aeration tank and a separate membrane tank. The tank of the former clarifier has been converted to an additional bioreactor such that, in the new process, 25% of the total treatment volume is occupied by the membranes. The membrane tank is equipped with $16 \times 500\ \mathrm{m}^2$ *PSH 500 PURON®* membrane modules, each containing sixteen 9-bundle module rows, providing a total membrane area of $8000\ \mathrm{m}^2$. The plant operation consists of a screening at 0.25 mm prior to biotreatment using two tanks in series. The sludge is then transferred to two membrane compartments, with a third chamber being available for future extensions of the plant. The membrane compartments are fed from the bottom so that the sludge flows upwards through the membrane

TABLE 5.16 Water Quality, Sobelgra Plant

Parameter, mg/L	Feed Concentration, mg/L	Load, kg/day	Effluent Concentration, mg/L
COD	1880–2100	4000	100–200
BOD	700–930	2000	2–5
TSS	330–460	800	0
Total N	35–50	100	1–2
Total P	13–15	30	<1

FIG. 5.46 The KMS PURON® modules at Sobelgra.

modules. Permeate is extracted from the membrane modules under a slight vacuum. The RAS is fed back into the biology tanks by means of spillways.

In 2009 the malting capacity was extended by another 80 ktonne p.a., making it the largest malting site in Europe. The capacity of the MBR has been extended accordingly with an additional four *PSH 500* modules installed in the third chamber (as described above). The extension of the MBR was again carried out by Veolia Water Solutions & Technologies.

5.3.5.2. Monheim

Client: City of Monheim
Primary consultant: ATM Abwassertechnik Detlef Wedi

The 0.6 MLD plant at Monheim (Fig. 5.47a) was installed in November 2008 as a replacement for an existing alternative commercial HF MBR module. It is a BNR plant with coagulant dosing applied for chemical phosphorus removal. An equalization tank of 2000 m^3 is installed upstream of the bioreactor, along with grit removal and bar screening to 1 mm. The bioreactor comprises two denitrification and nitrification tanks, each of 360 m^3, and four membrane tanks of 70 m^3 from which sludge is wasted at around 45 m^3 a day to maintain the MLSS concentration at 8–10 g/L in the bioreactor and ~12 g/L in the membrane tank.

The single membrane tank (Fig. 5.47b) has a train of six *PSH 500* units providing an overall membrane area of 3000 m^2. Operation is on a cycle of 6 min filtration and 30 s backflushing at 30 LMH. The maximum TMP is

FIG. 5.47 The MBR plant at Monheim: (a) the site and (b) plan view of the membrane modules.

600 mbar and the permeability ranges from 100 to 200 LMH/bar, with a mean flux of 10 LMH. Scouring air is delivered at a SAD_m of 0.3−0.6 $Nm^3/(m^2\ h)$, 0.3 on average. Fortnightly maintenance cleaning employs 2.5 g/L hydrogen peroxide and 2 g/L of citric acid/phosphoric acid. Recovery cleans are applied twice yearly using the same reagents but with the peroxide concentration raised to 10 g/L and that of the citric acid to 4 g/L.

The plant has performed well with no process upsets, producing a water quality of 1.4, 0.12, 4.2 and 0.4 mg/L of BOD_5, $N\text{-}NH_4$, $N\text{-}NO_3$ and TP, respectively, despite a relatively low wastewater temperature of 4.5−19 °C, 11.7 °C on average. Intestinal enterococci levels are at 10 and *Escherichia coli* at 25 CFU/100 mL.

5.3.5.3. Avranches

Client: City of Avranches
Primary contractor: Veolia Water

The 8 MLD plant at Avranches treats municipal water to meet the Bathing Water Directive for coastal discharge. As with Monheim, it is a BNR plant with biological and chemical nutrient removal, and began operation in May 2009. The plant employs grit removal upstream of a 2 mm perforated screen. Biotreatment comprises two streams each with an anoxic, anaerobic tank and aerobic tank of 377, 465 and 3048 m^3 in volume, respectively, with three membrane tanks of 190 m^3 each. The MLSS concentration is maintained at 8 g/L in the biotanks and 10 g/L in the membrane tanks.

The three membrane tanks each have four *PSH 1500* units, each containing 48 module rows, giving a total membrane area of 18,000 m^2. The membranes operate on a cycle of 6 min filtration and 30 s backflushing at 30 LMH. The target permeability is 180 LMH/bar, with a mean net flux of

18.5 LMH, and permeation is sustained by scouring air delivered at a SAD_m of 0.135−0.27 $Nm^3/(m^2\,h)$, 0.15 on average. A weekly 4.5 h maintenance clean is employed with 1000 mg/L NaOCl supplemented with a citric acid clean at a pH of 3. Annual 30 h recovery cleans may be used using the same reagents at double the hypochlorite concentration.

The plant produces effluent of 5, 35, 10 and 2 mg/L BOD, COD, TKN and TP, respectively, as well as <500 FC CFUs and <10 viruses PFUs per 100 mL.

5.3.5.4. Santa Paula

Client: City of Santa Paula
Primary contractor: PERC Water
Primary consultant: PERC Water (with Trussell Technologies Inc.)

The 12.9 MLD ADF (27 MLD PDF, 39 MLD peak 8-h flow) plant at Santa Paula treats municipal water, with significant infiltration, for discharge into spreading basins near the Santa Clara River with a product water quality target of 5 mg/L N total nitrate and nitrite. As of Spring 2010, this facility was the largest KMS MBR plant under commissioning in the USA. The plant is extremely compact, with all aeration basins underground and office, laboratory and operations buildings constructed above them. It is also the first design, build, operate and finance (DBOF) facility in the USA.

The plant has redundant 6 mm screens with vortex grit removal followed by 2 mm Huber fine screens, with each of the two trains rated at 64 MLD for the coarse screens and vortex grit chambers and 39 MLD for the fine screens. The three biotanks are operated at 8 g/L with 10 g/L in the six membrane tanks, with a design SRT and HRT of 15 d and 14 h, respectively, and a recycle ratio of 5 for sludge transfer. The membrane tanks (Fig. 5.48) each contain four 1500 m^2 units, giving a total membrane area of 36,000 m^2 and thus a design flux of 15 LMH at the ADF peaking to 31 LMH at the PDF. Membrane aeration is adjusted according to the flow with a design net SAD_p of 12, but the net SAD_p has been only 9 throughout commissioning. The maintenance clean frequency is fortnightly and the plant employs RO to provide softened water for cleaning.

5.3.6. Memstar

5.3.6.1. Shenghong Printing & Dyeing Co., Ltd

Primary contractor: Xiamen Visbe Co., Ltd as EPC

The 10 MLD ADF Memstar MBR at Shenghong treats effluent from dyeing operations upstream of a reverse osmosis plant, the treated water being reused for production processes. The plant was commissioned in October 2007.

The effluent flow is equalized for 8 h before passing on to the 5000 m^3 biotank (50:50 anoxic:aerobic) where the MLSS is held at around 3 g/L and recirculated to the 800 m^3 membrane tank, with sludge being wasted at 280 kg

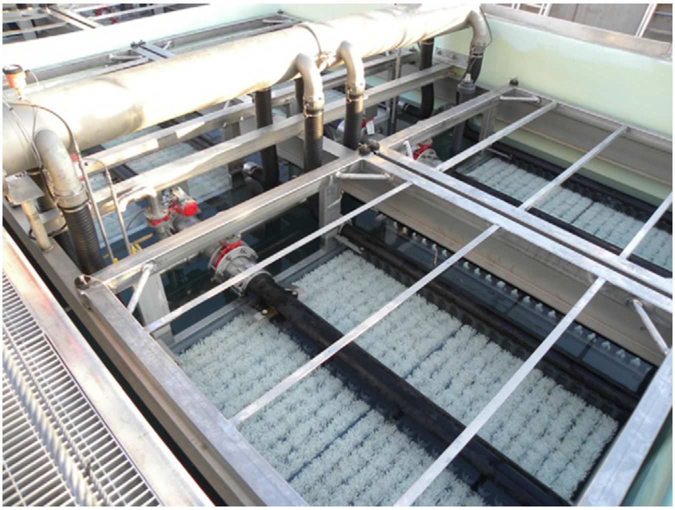

FIG. 5.48 The membrane tanks at Santa Paula.

DS/day from a total tank volume of 5800 m^3, i.e. 62 d SRT. The membrane tank holds 19,200 m^2 membrane area in six trains of four units, which each house 80 of the 10 m^2 *SMM-1010* HF modules. The net flux is maintained at 21 LMH and the TMP between 0.2 and 0.6 bar by backflushing at ~20 LMH for 1 min in every 10 min of operation, and air scouring at 0.2 $m^3/(m^2\ h)$. Maintenance cleaning comprises a monthly soak for 1.5 h in 200 mg/L NaOCl, with recovery cleans 1–2 times each year by soaking for 4 h each in 300–500 mg/L NaOCl and then 0.5 wt% citric acid. The plant achieves >90% COD removal down to effluent concentrations below 60 mg/L whilst incurring specific energy demand values in kWh/m^3 of 0.3 for both membrane and process aeration (and hence 0.6 kWh/m^3 for total aeration) and 0.006 kWh/m^3 for sludge transfer.

5.3.6.2. Sinopec Guangzhou Co., Ltd

Primary contractor: NOVO Envirotech (Guangzhou) Co., Ltd as EPC

The 4.8 MLD ADF plant at Guangzhou treats refinery wastewater to meet the GB50335-2002 4.2.2 water quality standard of China for reuse in cooling towers. The plant has been operating since November 2007.

The effluent flow passes through a DAF for removal of suspended oil and is then equalized for 8 h before passing on to the 3200 m^3 biotank, with a 1000 m^3 anoxic zone. The MLSS in the biotank is around 3 g/L, with sludge wasting at 220 kg DS/day (49 d SRT). The 400 m^3 membrane tank is fitted with 22,400 m^2 membrane area provided by 20 units fitted with 56 of the 20 m^2 *SMM-1520* HF modules, which are operated at a net flux of 12 LMH and a TMP of 0.03–0.3 bar. The membranes are relaxed for 1 min in every 10 min of operation, and maintenance cleaned by a soak once every 10 days for 1.5 h in 300 mg/L NaOCl. Recovery cleaning is through soaking (1−2 times a year) for 4 h each in 300−500 mg/L NaOCl followed by 0.5 wt% citric acid. The plant removes >90% of the COD down to <10 mg/L and >87% ammonia to <5 mg/L. The specific aeration energy demand for the plant is 0.55 kWh/m^3, evenly split between membrane and process aeration, with a further 0.006 kWh/m^3 for sludge transfer.

5.3.6.3. Sinopec Jinling Co., Ltd

Primary contractor: NOVO Envirotech (Guangzhou) Co., Ltd as EPC

The 6 MLD ADF plant at Jinling has been operating since December 2007. The plant treats wastewater, at a temperature of 28−42 °C, discharged from an oil refinery to achieve the standard set for recirculating cooling water by the petrochemical company: COD < 60 mg/L, NH_3-N < 5 mg/L.

The water is screened to 10 mm before entering a 5000 m^3 primary sedimentation tank and then a 10,000 m^3 equalization tank. The total tank volume is 4000 m^3 (hence 16 h HRT at the ADF), the anoxic and aerobic tanks each being 1600 m^3 (6.4 h HRT each), and recycling to the anoxic zone is at twice the forward flow. The MLSS concentration is kept at 3 g/L in the biotank and 50% higher in the membrane tank by wasting at 500 kg DS/day, and hence 26 d SRT. The 400 m^3 membrane tank contains 29,120 m^2 of membrane area from 28 units containing 52 *SMM-1520* modules, operated at a net flux of ∼10 LMH from an 8 min filtration cycle including 1 min relaxation and TMPs up to 0.6 bar. Maintenance cleaning is with 3 h soak (three times a month) in 500 mg/L NaOCl, with recovery cleans once or twice a year by soaking for 16−24 h in 0.8−1 g/L NaOCl. The plant removes 98% of the BOD, down to 3 mg/L, and 80% of the COD, down to 60 mg/L, with effluent ammonia and P levels of 2 and 0.2 mg/L, respectively. The total specific energy demand for the plant is 0.53 kWh/m^3, comprising 0.25, 0.38, 0.06 and 0.09 kWh/m^3 for membrane aeration, total aeration, sludge pumping and permeation, respectively.

5.3.7. Mitsubishi Rayon Engineering

5.3.7.1. Beijing Miyun

Client: Beijing Water Affairs Bureau of Miyun District
Primary contractor: Beijing Origin Water
Primary consultant: Beijing Guohuan Tsinghua Environmental Engineering Design and Research Institute

The 30 average MLD ADF plant in the Miyun Province of China came on-stream in April 2006, the year after the commercial launch of the MRE *SADF*™ membrane module and following successful demonstration of the technology at pilot scale. The hydraulic load to the plant increases from 27 to

36 MLD ADF from winter to summer. As at the end of 2009, it was the largest MRE MBR plant in the world. It is installed on the site of an existing modified sequencing batch reactor (MSBR) and, with a PDF of 45 MLD, is subject to significant infiltration. The plant is designed to produce effluent for landscape use.

The plant is fed from the MSBR effluent, which has a mean BOD_5 concentration below 50 mg/L and has been coarse-screened and degritted before passing through the 1 mm mesh drum screen upstream of the MBR. The bioreactor and membrane tank volumes are 4320 and 3240 m^3, respectively, providing a total HRT of ~6 h at ADF. At a mean WAS discharge rate of 950 kg DS/day (hence 55 d SRT) the MLSS is maintained at 6 and 8 g/L in the bioreactor and membrane tanks, respectively.

The membrane tank (Fig. 5.49) comprises six trains of six units which each contain $60 \times 25\ m^2$ *SADF* modules, giving an overall membrane area of 54,000 m^2. These operate on a cycle of 7 min filtration and 1 min relaxation and an SAD_m of 0.22−0.35 $Nm^3/(m^2\ h)$, 0.23 on average, which sustains a mean net flux of 28 LMH during the summer months and 21 LMH in the winter. The mean SAD_p is between 10 and 12, and the TMP ranges from 0.1 to 0.3 bar. Maintenance cleaning is with a weekly CIP using 500−3000 mg/L NaOCl supplemented by a quarterly clean with 1% citric acid, both incurring a 2-h downtime. Annual recovery cleans are based on a 24 h soak in 3000 mg/L NaOCl followed by a 4 h soak in 2% citric acid.

Ozonation is applied to the treated effluent for final disinfection and decolourization; TCs are not found on the treated effluent. Residual P levels are reduced to below 0.5 mg/L through dosing with 20−30 mg/L PACl. Other water

FIG. 5.49 The membrane tanks at Miyun.

quality determinants comprise <4, <30, <1 and <20, respectively, for BOD_5, COD, N-NH_4 and TN. The overall specific energy demand for the plant is 0.46 kWh/m^3, comprising 0.256 and 0.344 kWh/m^3 membrane and total aeration, with a further 0.02 and 0.096 kWh/m^3 for sludge transfer and permeation, respectively.

5.3.7.2. Kaetsu

Client: Wakasa Town (Fukui Prefecture)
Primary contractor: Mitsubishi Rayon Engineering Co., Ltd
Primary consultant: Japan Sewage Works Agency

The small 0.23 MLD Kaetsu WwTP in Japan treats municipal wastewater for coastal discharge. MBR technology was selected due to spatial constraints on the site combined with the requirement to meet stringent discharge water quality standards.

Pre-treatment at the plant comprises equalization for 4.5 h and fine screening to 0.5 mm by a bar screen. The anoxic and the combined aerobic/membrane tanks are both 31 m^3 in volume (and hence 6.5 h total HRT), and sludge is discharged at an average rate of 5 kg DS/day (equating to an SRT of 150 d) to maintain an MLSS of ~12 g/L. The aerobic tank contains three 210 m^2 *SUR*™ units operating at a mean design flux of 15 LMH. Scouring air is provided at a rate of 0.33–0.39 m^3/(m^2 h), 0.36 on average, and the mean TMP is around 0.45 bar. Biannual cleaning for 3 h with 3 g/L NaOCl is conducted at the plant. Aeration incurs an energy demand of 0.75 out of a total energy demand of 3.75 kWh/m^3, with sludge transfer and membrane permeation both contributing 1.5 kWh/m^3. The plant achieves BOD, TN and TP concentrations in the treated water of <3, <5 and <2 mg/L, respectively, with TC levels below 3 CFUs/100 mL.

5.3.8. Siemens

5.3.8.1. BeiXiao He (Beixaio River plant)

Client: Beijing Drainage Group
Primary contractor: Siemens China

The Beixiao River plant is the world's largest Siemens MBR plant, and was commissioned in June 2008. It is installed at the site of the existing works in north Beijing, expanding the plant capacity by 60–100 MLD as part of a number of water engineering projects relating to the Beijing Olympics with the water providing irrigation in the Olympic village and recharging the lake in the Olympic Park. MBR technology was selected largely on the basis of the limited space available on site. The total investment cost was 293m RMB (renminbi or Chinese yuan; around \$42m), including that of the 10 MLD RO plant downstream of the MBR which produces permeate meeting the most stringent standard for urban miscellaneous water reuse [GB/18920-2002]. The remaining

50 MLD permeate flow is disinfected by UV irradiation, and the target BOD, TSS and N-NH_4 standards are <6, <2 and <1.5 mg/L, respectively, the standard for Class III surface water [GB/3838-83] (MTC-IWA, 2009).

The 100 MLD flow is split 40:60 between the conventional plant and the MBR following coarse screening and grit removal in an aerated channel. There is then further pre-treatment of the 60 MLD flow by fine screening to 1 mm prior to anaerobic/anoxic/aerobic biotreatment for nutrient removal. The MLSS in the four biotank trains is maintained at 8−14 g/L by SRTs of 10−50 d, with PACl dosing applied for P removal. The wasted sludge is dewatered to 40−50% and then landfilled.

Recirculation through the eight membrane 150 m^3 tanks is at four times the feed flow; the HRT in these tanks is around 30 min. Each tank contains 38 racks, and each rack fitted with 16 of the *B30* modules, and thus a total membrane area of 145,920 m^2 is provided from the 304 racks (Walczenko, 2008). The membranes are operated at a mean flux of 16.5 LMH on a cycle of 12 min filtration, 1 min relaxation. The membranes are scoured at a minimum SAD_m of 0.33 $Nm^3/(m^2\,h)$, and hence a SAD_p of 20 at the average flux. Maintenance cleaning is weekly supplemented with recovery cleaning every three months (MTC-IWA, 2009).

5.3.8.2. Other Siemens Plants

At the end of 2009 there were over 80 Siemens MBR plants worldwide either installed, under construction or under contract − about 60% of these being in North America. About 30% of the installations are the *Xpress* package plants treating flows between 10 and 1500 m^3/day. Of the remaining plants, three are greater than 10 MLD in capacity: the Beixiao River plant, the Gippsland Water Factory in Australia and the plant serving Kuna in Idaho.

5.3.9. Sumitomo

5.3.9.1. Electronics Effluent, Japan

The 1.7 MLD ADF MBR plant was installed primarily for TN removal prior to discharge, and has been in service since May 2005. The plant operates without screening and at an MLSS of 8 g/L. The 420 m^3 membrane tank houses seven stacks (four in one tank and three in the other) containing 50 of the *SPMW-05B10* 10 m^2 modules in two trains, providing 3500 m^2 membrane area overall. The modules operate on a cycle of 9 min filtration and 1 min relaxation, with a SAD_m of 0.3 $Nm^3/(m^2\,h)$ which sustains a net flux of 25 LMH and a maximum TMP of 0.2 bar: the plant normally operates at >125 LMH/bar. Maintenance cleaning is through monthly soaking for 2 h in 2 wt% NaOH with 3000 mg/L NaOCl for a further 2 h. Recovery cleaning is with 2 wt% H_2SO_4 and 24 wt% NaOH for one and two days, respectively. The use of unusually aggressive chemicals reflects the robustness of the PTFE membranes to chemical attack. The total energy demand for the plant is 0.64 kWh/m^3, with 0.5, 0.08 and 0.06 kWh/m^3, respectively, used for aeration, sludge transfer and permeation. The BOD and TN are 96–97% removed to below 20 mg/L in both cases.

5.3.10. Tianjin Motimo

5.3.10.1. Tianjin City

Client: Kingway Brewery (Tianjin) Co., Ltd
Primary contractor: Tianjin Motimo Membrane Technology Ltd
Primary consultant: Guandong Provence Light Industry Design Company

The 3.5 MLD ADF (4.0 PDF) plant at Tianjin City has been operational since May 2008. It treats water from a brewery process to meet the standards required for scenic environmental use (GB/T 18921-2002). The water is post-treated with chlorination before being discharged to a lake on-site at the factory before going to a municipal plant.

The water undergoes 20 mm coarse screening, degritting, primary clarification, 7 h of equalization and fine mesh screening to 1 mm before entering the MBR. Biotreatment includes nutrient removal, with anoxic, anaerobic and aerobic tanks of 366, 215 and 648 m^3 volume, respectively, the membrane tank contributing a further 384 m^3 to give a total volume of 1613 m^3 (11 h HRT at ADF). The MLSS is 4.5–10 and 6–12 g/L in the process and membrane tanks, respectively, and the design WAS discharge rate 880 kg DS/day (and so 25–30 d SRT).

The plant is fitted with two different types of Motimo modules: the *FPAII* 20 m^2 'curtain' type and a 23 m^2 'column' module. There are four trains, each containing eight stacks with two trains each fitted with the curtain and column modules. There are 22 modules per stack of the curtain type, providing 7040 m^2 in total in two of the trains, with the other two trains providing 5888 m^2 in total from 16 stacks each containing 16 modules. The curtain and column modules thus, respectively, operate at mean fluxes of 11.8 and 14.1, with a maximum TMP of 0.6 bar. Membrane scouring is at 0.144–0.18 Nm^3/(m^2 h), 0.16 on average. The membranes are maintenance cleaned, for 30–60 min, twice a month with 300 mg/L hypochlorite. Recovery cleaning takes place once or twice a year and involves soaking for 6–8 h with 1.5–2 g/L sodium hypochlorite followed by 10 h of soaking in 5 g/L HCl.

The plant operates at an overall energy demand of 0.88 kWh/m^3, with 0.35 and 0.72 kWh/m^3 demanded for membrane and total aeration, respectively, and another 0.06 and 0.1 kWh/m^3 for sludge transfer and permeation. Permeate concentrations are <5, <40 and <3 mg/L of BOD_5, COD and ammonia, respectively, with 0.6–2.3 and 0–0.24 mg/L TN and TP. Log 4 removal of TCs is achieved to yield permeate concentrations below 50 CFUs per 100 mL.

5.4. SIDESTREAM MBRS

Multitube (MT) membrane modules for sidestream MBR (sMBR) technologies are standardized: the basic membrane element is of a standard size, and can be provided by a number of suppliers (Section 4.4). The standard size is generally 8″ (200 mm) diameter by 1 m long, and a large number of technologies are based on the most established commercial membrane product (Norit, Section 4.4.1) as well as other MT membrane product suppliers. Examples of installations based on these MT sMBR technologies are given in Section 5.4.1. The traditional sMBRs are based on liquid pumping through the externally placed membrane, linked to form a long, serpentine flow path for each train (or ‘loop’). This configuration is the most prevalent for industrial effluents containing recalcitrant organic matter, including landfill leachate (Figs 4.50a, 5.54). Some of the more recently developed MT sMBR applications, however, are based on the air-lift configuration, where the membrane is placed vertically and the flow path thus limited to the length of a single membrane element (Fig. 4.50b). This offers an alternative municipal wastewater treatment technology to the more established immersed configuration, and an example is provided in Section 5.4.1.2.

Whilst MT-based systems are by far the most prevalent of the sMBR installations, there are also FS and, less commonly, HF sMBRs. The FS sMBR actually predates the MT version, being the membrane configuration on which the original sMBRs were based. An example of this configuration is provided in Section 5.4.2.

FIG. 5.50 The Wehrle sMBR landfill leachate plant at Bilbao showing the tall bioreactor towers and sidestream membranes.

5.4.1. Multitube Systems

5.4.1.1. Eemshaven

Client: Holland Malt
Primary contractor: Norit PT

The 1.4 MLD ADF plant at Eemshaven in the Netherlands treats maltings effluents, and went on-line in June 2006. The selection of the technology was driven by constraints placed on the availability of the supplied freshwater for the industrial process. The recovered effluent is reused for the malting process following UV post-treatment.

The effluent is pre-treated with a 1 mm bow screen before passing into a 600 m^3 equalization tank (6 h HRT), which trims some of the more extreme, short-term hydraulic surges. There is further screening through a 0.8 mm mesh drum screen before the effluent passes into the 1749 m^3 biotank (17 h HRT), which includes a 300 m^3 anoxic zone. The biotank MLSS concentration is 15 g/L, compared to 17 g/L in the membrane modules themselves, and is pumped through the 8 mm internal diameter (ID) MT sidestream modules at a mean crossflow velocity (CFV) of 3.5 m/s and a TMP of 1−5 bar. This generates a mean flux of 110−125 LMH − somewhat higher than the design flux of 90 LMH and possibly aided by the high water temperature of 25−35 °C. The total membrane area is 648 m^2, provided by four lines of six modules, with each module comprising a pair of 27 m^2 MT Norit elements (Fig. 5.51).

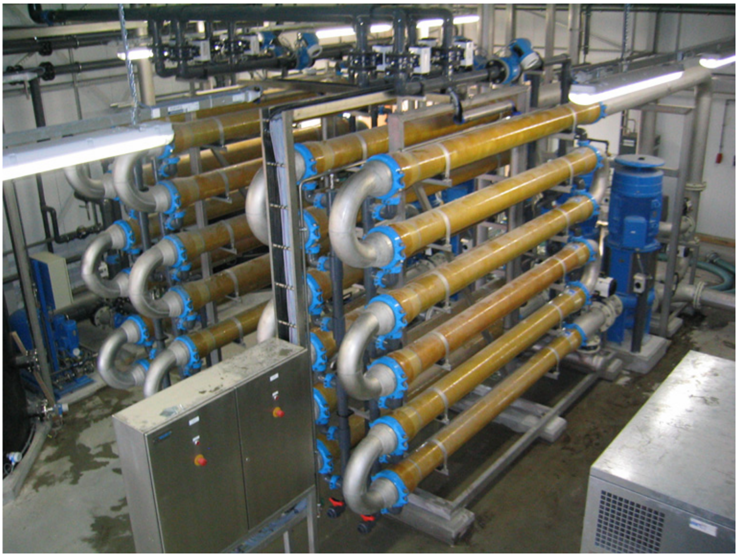

FIG. 5.51 The sidestream MT modules at Eemshaven.

Membrane chemical cleaning employs 400 mg/L NaOCl and 1 wt% citric acid, applied approximately every six to eight weeks.

The plant achieves >99.7% removal of BOD down to permeate levels below 5 mg/L, and >97% removal of the COD to <40 mg/L. TKN and TP effluent levels are <5 and <1 mg/L, respectively, the latter assisted by ferric dosing.

5.4.1.2. Ootmarsum

Client: Waterboard Regge en Dinkel
Primary contractor: Van Haarst (civil construction)
Primary consultant: Grontmij

The 3.6 MLD PDF (3.0 ADF) plant at Ootmarsum in the Netherlands is an example of the air-lift sidestream configuration of the Norit MT membrane, and has been in operation since October 2007 against municipal wastewater. An MBR was selected due to the water quality constraints of the receiving water body in 'the Pearl of Twente'.

Pre-treatment is through 6 mm bar screening, degritting and fine screening using a 2 mm drum screen, with 2 h of equalization. The BNR plant has anoxic, anaerobic and aerobic tank volumes of 150, 150 and 600 m^3, respectively, and thus 1000 m^3 and 6.7 h HRT in total. The MLSS is between 9 and 9.5 g/L. Whilst chemical dosing is available to enhance P removal no dosing has been required thus far. Six 14-module stacks, with 29 m^2 membrane area per module, provide a total area of 2436 m^2 from the 5.2 mm ID membrane tubes. The membranes operate at a flux of 50 LMH and 65 LMH at peak, and a SAD_m of 0.3 $Nm^3/(m^2\,h)$ with liquid pumping at a flow rate of 0.7 $m^3/(m^2\,h)$. The mean SAD_p is thus ~6, and the specific energy demand for membrane operation 0.35 kWh/m^3.

Physical cleaning comprises backflushing at 300 LMH for 7 s every 7 min. Dislodging of incipient matting (Section 3.6.2) at the module inlet is achieved through draining of the modules under gravity every 6 h for approximately 45 s. This is sufficient to completely suppress clogging at the inlet. Chemical cleaning comprises six weekly cleans for 4 h with 300 mg/L NaOCl, supplemented with overnight soaks in 1% citric acid every 18 weeks. The plant delivers effluent water quality of 0.8, <20, 1.3 and 2.3 mg/L BOD, COD, TKN and TP, respectively.

A further small (0.8 MLD) air-lift MBR plant in South India treats tannery effluent, with the treated effluent going to RO before the product water is reused in the process. The plant began operation in June 2009. The water is screened to 1 mm before equalization and then undergoes further fine screening to 0.5 mm, mainly to reduce the large amount of hair from this application. The biotank MLSS concentration is held at 10−10.5 g/L. At the feed temperature of 25−35 °C the flux is ~40 LMH. The filtration and physical cleaning cycle, as well as the liquid CFV and membrane aeration rate, is otherwise the same as for Ootmarsum. Maintenance cleaning is with monthly soaking for 4 h in 400 mg/L hypochlorite.

5.4.1.3. Kanes Foods

Primary contractor and consultant: Aquabio

The plants at Kanes Foods and Bourne Salads in the United Kingdom both represent examples of food effluent recycling with an sMBR, based on a pumped UF MT membrane, employed upstream of a reverse osmosis plant and, finally, a UV polishing unit. The Kanes Foods plant in Worcestershire was designed and installed by Aquabio, based on the company's *AMBR* pumped sidestream technology (Section 4.4.5), and commissioned in 2001. The plant has a 0.815 MLD capacity (Table 5.17), with 80% of this flow being recycled.

The process treatment scheme (Table 5.18) comprises upstream screening, flow balancing, DAF treatment (for fine vegetable solids removal), the MBR itself and downstream treatment by reverse osmosis followed by UV disinfection. The permeate water is blended with mains water for use within the factory. The MBR comprises two 250 m^3 bioreactors with four banks of crossflow membrane modules. The maximum MLSS concentration employed has been 20 g/L, but the bioreactor is generally operated at around 10 g/L resulting in food-to-microorganism (F:M) ratios of around 0.13 kg COD/kg MLSS day. Sludge production is calculated as being 0.14 kg DS/kg COD, removed from a sludge of age over 100 days. Each membrane bank (Fig. 5.52a) is fitted with four 200 mm diameter MT UF Norit membranes. The membranes operate at an average flux of 153 LMH normalized to 25 °C. The permeate water has average TSS, BOD and COD concentrations of 4, 7 and 16 mg/L, respectively. The UF permeate is passed to a two-stage reverse osmosis plant (Fig. 5.53) which achieves an overall recovery of 75−80%. The reject stream is discharged to sewer and the permeate, which typically has a conductivity of

TABLE 5.17 Design Basis for Plants at Kanes Foods

Parameter	Original Plant	New Plant
Volume to screening/balancing	1.200 MLD	2.250 MLD
COD concentration (average)	1000 mg/L	1000 mg/L
Volume to DAF system	0.815 MLD	1.435 MLD
Volume to MBR process	0.815 MLD	1.435 MLD
Volume to RO system	0.815 MLD	0.858 MLD
Volume of potable quality water for reuse	0.650 MLD	0.600 MLD

TABLE 5.18 Summary of Process Treatment Schemes, Kanes Foods

Process Stage	Old Plant	New Plant
Preliminary screening	*Old screen now redundant*	Rotary drum screen, 220 m^3/h, 0.7 mm
Balancing	900 m^3 balance tank used for outlet balancing	1640 m^3 balance tank with mixing/aeration
Clarification	DAF system, 50 m^3/h	DAF system, 60 m^3/h
Secondary screening	Backflushing filter/screen, 0.5 mm	Rotary drum screen, 80 m^3/h, 0.25 mm
Biological treatment	2 × 250 m^3 bioreactors c/w mixing/aeration	765 m^3 bioreactor with mixing/aeration
UF membrane separation	4 × membrane tanks, 324 m^2 total membrane area, 34 m^3/h	4 × membrane banks, 1282 m^2 total membrane area, 60 m^3/h
Reverse osmosis	RO system, 27 m^3/h	RO system, 25 m^3/h
UV disinfection	UV disinfection unit, 27 m^3/h	
Sludge handling	22 m^3 sludge tank	50 m^3 sludge tank

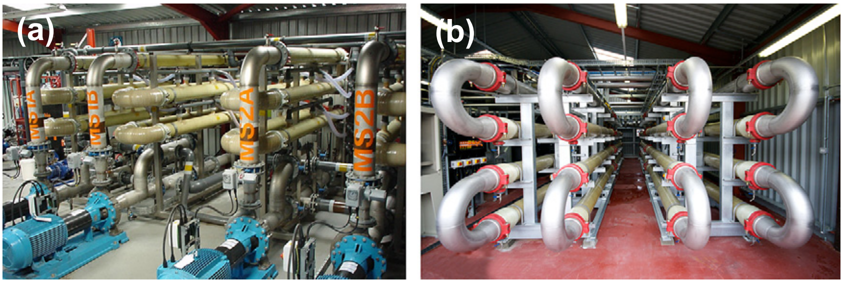

FIG. 5.52 The membrane banks at Kanes Foods: (a) the original four banks from 2001 (with one 'dummy' tube) and (b) the new *AMBR LE*™ banks.

40–100 μS/cm, is passed to the UV disinfection unit and then to the client's water supply tank.

The plant has performed consistently in terms of biological treatment, membrane performance and final reuse water quality. For the majority of the time membrane performance has been better than design, allowing one bank to be maintained as a standby and so offering greater process flexibility and lower energy use. Occasional reductions in membrane flux have been linked

FIG. 5.53 The RO skid at Kanes Foods.

to poor biomass health, which has been rectified by closer management of the process.

Due to continued production expansion and the consequent pressure on incoming water supply and discharge consent, a second wastewater treatment plant was constructed and commissioned in February 2010 to provide a further 1.44 MLD of biological/MBR treatment capacity and an additional 0.6 MLD of reuse water (Table 5.18). The plant replicates the successful process scheme of the existing plant but utilizes Aquabio's *AMBR LE*™ technology (Section 4.4.5) to provide significant energy savings compared to the original plant. New common inlet screening and flow balancing facilities have allowed the original balance tank to be used to balance outgoing flows to sewer. DAF pre-treatment is again employed to remove fine vegetable solids upstream of the MBR. A single bioreactor of 765 m^3, allowing for MLSS concentrations up to 12 g/L, is aerated by a blower assisted slot type aeration system.

The *AMBR LE*™ UF membrane system comprises four banks of membranes (Fig. 5.52b) providing a total membrane area of 1280 m^2. Low energy performance is achieved by the use of backflushing to control membrane fouling, allowing much reduced crossflow velocities and hence significantly lower energy use. The combination of backflushing frequency and variable crossflow velocity is optimized to give the lowest energy use for the required throughput. Significant process flexibility is offered by the inclusion of variable speed recirculation pumps and optional permeate pumping to control the membrane TMP and hence flux performance. As of May 2010, a filtration energy demand of 0.5 kWh/m^3 was being maintained at a net flux of 45 LMH.

5.4.1.4. Immingham Docks

Primary contractor and consultant: Aquabio

The 0.48 MLD ADF (0.6 MLD PDF) sMBR plant was installed in July 2006 at Simon Storage (the Immingham Storage Company, ISCo) at Immingham Docks in the Humberside region of the United Kingdom, following three months of successful piloting of the treatment scheme. MBR technology was chosen on the basis of footprint limitation combined with the requirement to remove hazardous 'red list' substances prior to discharge to the Humber estuary. A 0.24 MLD RO plant was subsequently installed downstream to recover some of the water for reuse as boiler feedwater in local industrial facilities. Although designed for 0.6 MLD the plant has received and successfully treated 1.7 MLD.

The run-off water passes through an existing 3000 m^3 buffer tank and a 1 mm basket strainer to a primary sedimentation tank fitted with lamella plates. The supernatant then passes to the biotank, where dosing with urea takes place to balance the C:N ratio to maintain effective biotreatment, and aeration and mixing is provided via a proprietary jet aeration technology (*JETOX*). The MLSS is held at around 12 g/L in the 600 m^3 biotank (hence 24 h HRT at PDF) by recirculation through the MT membrane skid, which comprises six modules each containing two 27 m^2 elements, and discharges sludge at an average rate of 3.6 m^3/day (hence >160 d SRT). The membrane feed is screened through a 0.8 mm basket strainer. The membranes operate at a mean CFV of 3.8 m/s and a flux of 60–200 LMH. MBR permeate flow of 0.4 MLD passes to an RO unit, which operates at a conversion of 60–70% with the reject being discharged to the Humber with the remaining 0.24–0.28 MLD of the MBR permeate. Chemical cleaning is conducted as required by flushing with UF permeate followed by a period of recirculation and soaking with sodium hypochlorite solution or an *Ultrasil* proprietary solution.

The plant meets stringent EA discharge consents for hazardous substances which include styrene, toluene, xylene, phenol, benzene, chlorinated volatile organic chemicals such as tetrachloroethylene and chloroform, and toxic metals such as mercury and chromium. The mean COD and ammonia levels are, respectively, reduced to 50 and 3 mg/L, and the biological aeration and membrane permeation specific energy demand values are 1.0 and 2.9 kWh/m^3, respectively.

5.4.1.5. Dairy Crest, Foston

Primary contractor and consultant: Wehrle

The 0.65 MLD plant at Foston (Fig. 5.54) in the United Kingdom treats dairy effluent for river discharge. The effluent is pre-treated by DAF assisted by 750 mg/L dosing with coagulant and 10 mg/L polymer, before fine screening to 1 mm with a basket strainer. The total volume provided by the pre-treatment, including equalization, is 600 m^3 (22 h HRT at the flow of 0.65 MLD),

FIG. 5.54 The Dairy Crest plant at Foston.

compared to the 700 m^3 biotank (which then correspondingly gives 26 h HRT). The MLSS is held at 15 g/L by wasting at ~100 kg DS sludge/day.

Filtration is through external sidestream MT UF modules, 3 m long and fitted with 8 mm membrane tubes in 200 mm (i.e. 8″) diameter modules, 12 elements being divided between two trains and providing a total membrane area of 318 m^2. When first operated it transpired that the sustainable flux was around 185 LMH, more than double the mean operating flux of 85 LMH and substantially higher than the design flux of 105 LMH. As a result, the number of elements in each train (or loop) was reduced from six down to four, with an effective total operating area of 212 m^2. The flux is sustained by a CFV of 3.5 m/s and a TMP of 2.8 bar, with maintenance cleaning through flushing for 2 h with 1% caustic soda every six to eight weeks combined with a further 2 h of flushing with 0.2 wt% hypochlorite. Filtration incurs an energy demand of 1.9 kWh/m^3.

The plant achieves 90, 99.4 and >99.9% removal of ammonia, COD and BOD_5, respectively, to give corresponding permeate concentrations of 3.5, 17 and <2 mg/L.

5.4.1.6. Baishi Ao, Foshan City

Primary contractors and consultants: Wehrle/Veolia

The landfill site at Baishi Ao, Foshan City, generates 1.05 MLD of leachate for treatment by MBR. The MBR permeate is post-treated by nanofiltration (NF) prior to discharge of 0.85 MLD final effluent to a surface water; the system operates at an overall conversion of 81%.

The leachate flow is equalized in a 500 m^3 tank (11.4 h HRT) and screened to 0.4 mm prior to entering the biotank, which is operated at an MLSS of 15 g/L. Biotreatment comprises 900 and 4200 m^3 of anoxic and aerobic

FIG. 5.55 The membrane skids at Baishi Ao, Foshan City: NF (left) and UF (right).

treatment respectively (with corresponding HRTs of 21 h and 4 d), the recycle ratio being ~10. The sidestream MT UF (Fig. 5.55) is configured as four trains of six 27 m^2 elements, providing a total membrane area of 648 m^2. The operating flux is 85 LMH, well in excess of the mean flux of 68 LMH, with the TMP and CFV at 3.2 bar and 4.8 m/s, respectively. The full flow capacity only arises during periods of sustained rainfall, such that during the dry season the plant can operate on two to three of the four loops. The membranes are cleaned every six weeks with alkaline and acid cleaners. Permeate BOD and COD levels are at <15 and 1500 mg/L, and those of ammonia, TKN and TN at <5, <100 and <300 mg/L, respectively.

5.4.1.7. Glen Meadows Retirement Home, Glen Arm, MD, USA

Primary contractor and consultant: Dynatec Systems, Inc.

The small (0.34) MLD MBR plant at the Glen Meadows Retirement Home provides water for reuse for utilities, and at some future point for irrigation. The plant system comprises equalization (300 m^3, 21 h HRT), rotary drum screening to 1 mm, anoxic/aerobic biotreatment with the option of supplementary caustic and organic carbon dosing to maintain denitrification, air-lift sidestream membrane filtration and UV post-treatment. The MLSS concentration is 10 g/L and the anoxic and aerobic tanks are, respectively, 48 and 80 m^3 in volume (hence 9 h HRT), with a recycle ratio of 5.

The membrane skid contains 12 vertical MT elements, each element having 29 m^2 membrane area provided from 5 mm diameter tubes, in two skids of six modules each: the peak flux is thus around 41 LMH. Membrane aeration is applied at a rate of 0.59 $Nm^3/(m^2\ h)$ (and so an SAD_p of ~14 at peak flow) and liquid pumping at 0.71 $m^3/(m^2\ h)$. The filtration cycle is 10 min long with 7 s of backflushing at up to 420 LMH, limited by pressure. Maintenance cleaning is monthly with a 1h soak in 300 mg/L NaOCl supplemented with a four-monthly 8 h soak in 2 wt% citric acid.

Permeate water quality of <5 and <20 mg/L BOD_5 and COD is attained, along with ammonia, TN and TP levels of <0.1, <5 and ~3 mg/L, respectively. Target nutrient levels are achieved without additional chemical dosing.

5.4.2. Flat Plate Systems

5.4.2.1. The Queen Mary II MBR WwTP

Primary contractor and consultant: Orelis Environment SAS

Legislation concerning water discharged to sea has been subject to increasing stringency. In the past, the practice had been to employ conventional wastewater treatment for foul waters and discharge greywater to the sea. The latter is no longer considered acceptable; a number of pieces of legislation applied to different geographical regions have been promulgated which effectively disallow greywater discharge, the most restrictive being the Alaskan regulations which limit both TSS and BOD to 30 mg/L, faecal coliforms to 20 per 100 mL and heavy metals to undetectable levels. Some commercial cruise ship companies have adopted targets which are even stricter than the legislation, since (a) the marine conservation lobby is likely to bring about ever more stringent standards; and (b) only slightly more rigorous treatment would permit water reuse for general purposes such as cooling and sluicing down.

Ship-board wastewater treatment presents a number of challenges. The deck height, the floor-ceiling distance in which the MBR has to be located, is normally around 2.5 m. This limits the aeration tank depth and prohibits inserting the membrane module into, or lifting it from, the tank in situ. The plant has to run with very little maintenance, since marine crew generally have little or no experience of wastewater treatment. Full plant automation is thus required, and in the case of the Queen Mary II (QM2) staff training has also been provided. Lastly, accurate effluent water data are at a premium in the ship industry.

The QM2 is one of the world's largest passenger ships, and is owned by the Carnival Group. A total of around 1.1 MLD of wastewater is generated on board the ship from four main sources, these being:

- Greywater from accommodation (0.65 MLD) and laundering operations (0.150 MLD);
- Galley water (0.2 MLD) from the kitchens and food waste drainage;
- Black (or foul) water (0.100 MLD); and
- Recreational pool and sauna wastewater (<0.01 MLD).

TABLE 5.19 Basis for Process Design: Wastewater Quality on Board the QM2

Parameter	Blackwater	Greywater
Flow, m^3/day	300	800
COD (average), mg/L	3500	600
TSS, mg/L	1000	1200
BOD_5, mg/L	1200	200
N-NTK, mg/L	Not defined	Not defined
P-tot, mg/L	Not defined	Not defined
pH	4.5–8	4.5–7.5
Temperature, °C	Not defined	Not defined
FOG, mg/L	200	20
TDS, mg/L	500	500
Phenol, mg/L	0.002	0.018
Arsenic, mg/L	Not defined	0.06
Copper, mg/L	Not defined	2.00
Lead, mg/L	Not defined	0.07
Zinc, mg/L	Not defined	2.90

These flows can be segregated into two streams of black- and greywater (Table 5.19). Carnival has set targets of 10 mg/L TSS and undetectable faecal coliforms for the treated water, thereby providing the option of reuse of the greywater stream.

The full treatment scheme (Fig. 5.56) comprises a pre-filtration unit using a 1 mm drum filter with hydrocyclones for sand removal. This is followed by biotreatment in a 150 m^3 tank fitted with two blowers each providing 400 Nm^3/h of air which maintains the dissolved oxygen at 0.5–2 mg/L. The tank height is 4.5 m installed within two decks: there is a 0.5 m clearance between the top of the tank and the floor of the deck above. The MLSS in the bioreactor is kept between ~8 and ~12 g/L and the reactor temperature between 20 and 35 °C. The sludge production rate is 0.15 kg MS/kg COD.

Filtration is achieved by two lines of five skid-mounted *Pleiade®* UF modules (Fig. 5.57) containing 200 membrane elements providing 70 m^2 total membrane area per module, and thus 700 m^2 of total membrane filtration area. The skid has a footprint of 9 m × 4 m and is 2.5 m high, and each line is fitted

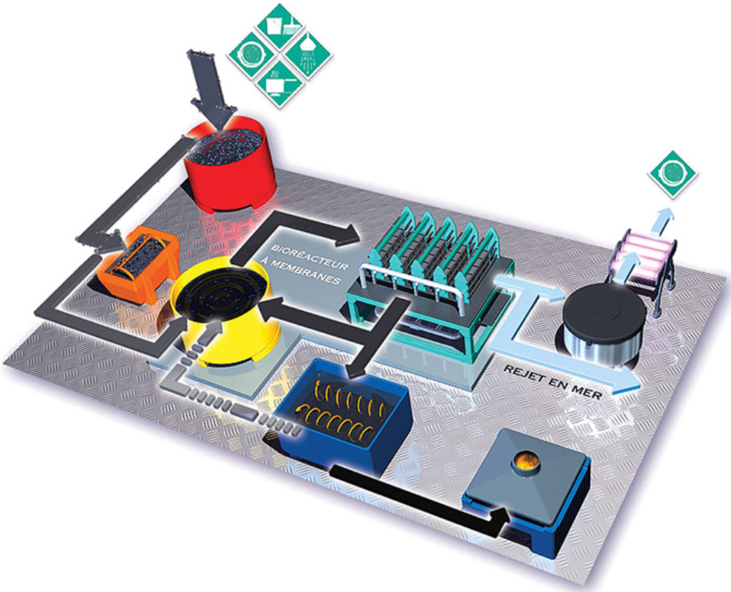

FIG. 5.56 The QM2 wastewater treatment plant.

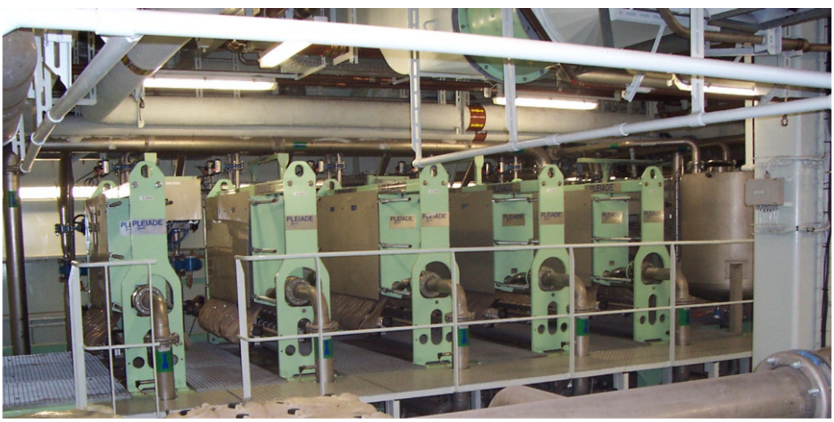

FIG. 5.57 *Pleiade*® skid on the Queen Mary II.

with a cleaning tank providing an automated CIP every three to five weeks to supplement the water flush every 2–4 h. The skid is fed with two feed pumps (duty and standby) and five circulation pumps, providing a mean TMP of 2 bar and a flux of 75 LMH. The greywater line is fitted with activated carbon and UV irradiation downstream of the membrane stack to enable reuse. The wasted sludge from both the grey- and blackwater lines is delivered to a centrifuge.

TABLE 5.20 Actual Wastewater Quality on Board QM2: Averaged Data from Three Samples of FWD and Two Samples of GGW

Parameter, mg/L	Food Waste Drainage (FWD)			Galley Greywater (GGW)	
COD	52,000	880	2000	25,000	2000
Total organic carbon	3480	270	580	3000	530
BOD_5		350	33,000	23,000	1300
TSS	21,100	110	39,000	3400	300
Ammonia nitrogen	63	1	12	5.8	ND
Total phosphorus		10	26	51	26
FOG		81	200		210
pH	3.44	5.59	3.42	3.79	5
Temperature				31	

**Not disclosed.*

A number of challenges were presented in the early stages of operation. The blackwater was out of specification, causing 'balling' of solids (formation of gross solid ~10 mm particles) from paper fibrous material which caused membrane channel blockage. This necessitated disassembling of the modules for manual cleaning of the membranes. After investigating a number of remedial measures, successful demonstration led to the installation of a supplementary pre-filter. The discharge from the kitchen waste was also out of specification (Table 5.20) downstream of the grease traps, causing foaming which demanded dosing with anti-foaming agent.

The plant has been in operation since January 2003 and is now operating well, thanks to better information regarding the quality of the various effluent streams and a constructive relationship with the crew.

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Appendix A

Conversion Factors

Common conversion factors between Imperial and metric membrane technology parameters are provided below.

Imperial Units	Metric Units	Imperial to Metric	Metric to Imperial
lbs	kg	0.454	2.20
inches	m	25.4	0.0394
SCFM	Nm^3/h	1.70	0.588
$SCFM/ft^2$	$Nm/(m^2\ h)$	18.3	0.0546
psig	bar	0.0681	14.8
psi	bar	0.0690	14.5
ft	m	0.305	3.28
sq ft	sq m	0.0929	10.8
US gal	litres	3.79	0.264
MGD	MLD	3.79	0.264
GFD	LMH	1.70	0.588
GFD/psi	LMH/bar	24.6	0.0406
GFD	m/d	0.0408	24.5
°F	°C	Subtract 32, multiply by 0.55	Multiply by 1.8, add 32

SCFM Standard cubic feet per minute.
psi(g) Pounds per square inch (gauge pressure).
MGD Mega-gallons per day (US).
MLD Megalitres per day.
GFD Gallons per square foot per day.
LMH Litres per square metre per hour.
1 billion = 1000 million.

Appendix B

MBR Biotreatment Base Parameter Values

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MWH

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The MBR Book.

Parameter	Value	Units	Reference	Process	Wastewater Tested and Additional Comments
$k_{e,n}$	0.21	1/d	Dinçer and Kargi (2000)	ASP with denitrification	Synthetic wastewater (100:100 COD:NH_4-N)
$k_{e,n}$	0.05–0.15	1/d	Tchobanoglous et al. (2003)		0.08 Typical
$k_{e,n}$	0.055	1/d	Jiang et al. (2008)	MBR	15 °C synthetic wastewater, 17 d SRT
$k_{e,n}$	0.12	1/d	Harremoës and Sinkjaer (1995)	ASP with denitrification	20 °C Municipal wastewater, 397–256 mg/L COD: 40–35 TKN, 18–21 d SRT
k_e	0.06–0.2	1/d	Tchobanoglous et al. (2003)		0.12 Typical
k_e	0.067	1/d	Yenkie, Gerssen, and Vogelphol (1992)	High-rate CAS	22 °C synthetic wastewater (1090:872 mg/L COD:BOD), SRT 0.3 d
k_e	0.050	1/d	Fan, Urbain, Qian, and Manem (1996)	MBR	Municipal wastewater, 30 °C, 411–72 mg/L COD: 26–53 NH_4-N, 20 d SRT
k_e	$0.85\theta_x^{-0.62}$	1/d	Huang, Gui, and Qian (2001)	MBR	Domestic wastewater (∼250:20:170 mg/L COD:NH_4-N:SS) SRT 5–40 d
k_e	0.023	1/d	Liu, Huang, Jinying, and Quan (2005)	MBR	Synthetic wastewater (220–512 mg/L COD, 36–72 NH_4-N) infinite SRT
k_e	0.08	1/d	Wen, Xing, and Qian (1999)	MBR	30 °C Urban wastewater (∼500 mg/L COD) SRT 5–30 d
k_e	0.025–0.075	1/d	Xing, Wu, Qian, and Tardieu (2003)	MBR	Variable (30–2234 mg/L COD) Municipal wastewater, SRT 5–30 d
k_e	0.048	1/d	Yilditz, Keskinler, Pekdemir, Akay, and Nuhoglu (2005)	MBR	26 °C Synthetic wastewater, 1090 mg/L COD, SRT 0.3 d
k_e	0.151–0.0261	1/d	Al-Malack (2006)	MBR	Lab-scale MBR, organic loading rate 0.4–3 kg COD/(kg MLSS d); 3000–15,000 mg/L MLSS
K_n	0.1–0.4	g/m^3	Harremoës and Sinkjaer (1995)	ASP with denitrification	20 °C Municipal wastewater, 397–256 mg/L COD: 40–35 TKN, 18–21 d SRT
K_n	0.5–1	g/m^3	Tchobanoglous et al. (2003)		0.74 Typical

K_n	0.1–0.15	g/m^3	Manser, Gujer, and Hansruedi (2005)	MBR & CAS in parallel	Domestic wastewater (quality not given), SRT 20 d
K_n	0.85	g/m^3	Wyffels et al. (2003)	MBR with low DO	30 °C Sludge digester supernatant 605:931 mg/L COD:TAN (total ammonium nitrogen), SRT > 650 d
K_n	0.2	g/m^3	Jiang et al. (2008)	MBR	15 °C Synthetic wastewater, SRT 17 d
K_n	0.01–0.34	g/m^3	Groeneweg, Sellner, and Tappe (1994)	Nitrifying chemostat	30 °C synthetic wastewater, 392 mg NH_4-N/L. Values were dependent on pH, temperature and bacterial species
K_s	5–40	g/m^3	Metcalf and Eddy (2003)		20 Typical
K_s	80	g/m^3	Yenkie, Gerssen, and Vogelphol (1992)	High-rate CAS	22 °C Synthetic wastewater (1090:872 mg/L COD:BOD), SRT 0.3 d
K_s	192	g/m^3	Yilditz et al. (2005)	MBR	26 °C Synthetic wastewater, 1090 mg/L COD, SRT 0.3 d
K_s	289–2933	g/m^3	Al-Malack (2006)	MBR	Lab-scale MBR, organic loading rates 0.4–3 kg COD/(kg MLSS d); MLSS: 3000–15,000 mg/L
Y	0.3–0.5	g VSS/g bCOD	Metcalf and Eddy (2003)		0.4 Typical
Y	0.44	per d	Yenkie et al. (1992)	High rate CAS	22 °C synthetic wastewater (1090:872 mg/L COD:BOD), SRT 0.3 d
Y	0.61	g VSS/g COD	Fan et al. (1996)	MBR	30 °C Municipal wastewater, 411–72 mg/L COD: 26–53 NH_4-N, 20 d SRT
Y	0.28–0.37	g VSS/g COD	Huang et al. (2001)	MBR	Domestic wastewater (~250:20:170 mg/L COD:NH_4-N:SS) SRT 5–40 d
Y	0.288	g VSS/g COD	Liu et al. (2005)	MBR	Synthetic wastewater (220–512 mg/L COD, 36–72 NH_4-N) infinite SRT
Y	0.40–0.45	g VSS/g BOD	Lübbecke, Vogelpohl, and Dewjanin (1995)	MBR	Synthetic wastewater (8500–17,600 mg/L COD, 36–72 NH_4-N), SRT 1.5–8 d
Y	0.56	g VSS/g COD	Wen et al. (1999)	MBR	30 °C urban wastewater (~500 mg/L COD) SRT 5–30 d
Y	0.25–0.40	g VSS/g COD	Xing et al. (2003)	MBR	Variable (30–2234 mg/L COD) Municipal wastewater, SRT 5–30 d

(*Continued*)

Parameter	Value	Units	Reference	Process	Wastewater Tested and Additional Comments
Y	0.58	g VSS/g COD	Yilditz et al. (2005)	MBR	26 °C Synthetic wastewater, 1090 mg/L COD, SRT 0.3 d
Y	0.487–0.583	g VSS/g COD	Al-Malack (2006)	MBR	Lab-scale MBR, organic loading rates 0.4–3 kg COD/(kg MLSS d); MLSS 3000–15,000 mg/L
Y_n	0.34	g TSS/g N	Dinçer and Kargi (2000)	ASP with denitrification	Synthetic wastewater (100:100 mg/L COD:NH_4-N)
Y_n	0.16	g VSS/g N	Harremoës and Sinkjaer (1995)	ASP with denitrification	7.5 °C Municipal wastewater, 397–256 mg/L COD: 40–35 TKN, 18–21 d SRT
Y_{obs}	0.31–0.36	g VSS/g COD	Tao et al. (2005)	3 MBR plants in parallel	SRT 14–28 d, settled sewage, 265 mg/L COD
Y_{obs}	0.11	g VSS/g COD	Liu et al. (2005)	MBR	Synthetic wastewater (220–512 mg/L COD, 36–72 NH_4-N) infinite SRT
Y_{obs}	0.16–0.38	g VSS/g COD	Wen et al. (1999)	MBR	30 °C Urban wastewater (∼500 mg/L COD) SRT 5–30 d
μ_{max}	3–13.2	per d	Metcalfe and Eddy (2003)		6 Typical
μ_{max}	0.125	per d	Yenkie et al. (1992)	High rate CAS	22 °C Synthetic wastewater (1090:872 mg/L COD:BOD), SRT 0.3 d
μ_{max}	3.24	per d	Yilditz et al. (2005)	MBR	26 °C Synthetic wastewater, 1090 mg/L COD, SRT 0.3 d
μ_{max}	1.28–6.46	per d	Al-Malack (2006)	MBR	Lab-scale MBR, organic loading rates 0.4–3 kg COD/(kg MLSS d); 3000–15,000 mg/L MLSS
μ_{max}/K_s	0.001–0.01	per d	Wen et al. (1999)	MBR	Lower than ASP. 30 °C urban wastewater (∼500 COD) SRT 5–30 d
$\mu_{n,max}$	0.1–0.2	per d	Fan et al. (1996)	MBR	30 °C Municipal wastewater, 411–72 mg/L COD: 26–53 NH_4-N, 20 d SRT
$\mu_{n,max}$	0.2–0.9	per d	Metcalf and Eddy (2003)		0.75 Typical
$\mu_{n,max}$	0.6	per d	Jiang et al. (2008)	MBR	15 °C Synthetic wastewater, SRT 17 d
$\mu_{n,max}$	2.02	per d	Wyffels et al. (2003)	MBR, low DO	∼30 °C sludge digester supernatant 605:931 mg/L COD:TAN (total ammonium nitrogen), SRT > 650 d

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Appendix C

Membrane Products

Contributors are as listed for Chapter 4

On the following pages are listed the membrane products that have either been used for membrane bioreactor (MBR) duties or are recommended for this purpose by the suppliers (Table A3.1, reproduced from Table 4.5). Providers of these data are as identified on the Chapter 4 title page. Specifications are provided, for all products where such information is available, in the subsequent sections. Some multitube (MT) products may be used by process suppliers interchangeably in their proprietary technologies, provided they are of standard module size (typically 8″, or 203 mm, diameter modules). There is also one supplier (Memos) who appears to supply bespoke HF and MT modules, such that their range of products is very broad.

The terminology used for describing the membranes is summarized in Table C.2. If the generic term element is applied to be the smallest replaceable component of membrane technology then the term 'module' applies to a single element for a rectangular hollow fibre technology, but it normally applies to a collection of elements for flat sheets. The term 'stack' can be considered to apply generally to a collection of elements, though the terms 'rack', 'unit', 'cassette' (for rectangular HF modules) and 'skid' are also used. A collection of immersed stacks is usually called a 'train', and the train may include the biological tank.

The packing density data provided refer to the membrane area provided per internal module volume, as either provided by the supplier or estimated from the available specification data. The overall packing density φ refers to the area based on the FS panel or HF/MT module. The area per module (FS) or rack (HF) is given in parentheses, where provided by the supplier. Abbreviations for membrane polymeric materials are the following:

PAN	polyacrylonitrile
(HD)PE	(high density) polyethylene
PES	polyethylsulphone
PS	polysulphone
PTFE	polytetrafluoroethylene
PVDF	polyvinylidine difluoride

Please note that the authors cannot be held liable for any errors in this appendix. Any specifications must be checked with the suppliers directly, the

TABLE C.1 Summary of Commercial MBR Membrane Module Products

Immersed (iMBR)		Sidestream (sMBR)
Flat sheet	Hollow fibre	Multitube/multichannel
A3 — *MaxFlow*	Asahi Kasei — *Microza*™	Berghof — HyPerm-AE; HyPerflux
Alfa Laval — *Hollow Sheet*	Beijing Origin Water Technology Co.	Norit X-Flow — F4385, F5385
Agfa-VITO	Canpure — *Canfil*	Orelis Environment —*PLEIADER*®, *KLEANSEPR*®
Brightwater — *MEMBRIGHT*®	Ecologix — *EcoFlon*™, *EcoFil*™	MEMOS — Membrance Modules Systems GmbH — *MEMCROSS*
Colloide — *SubSnake*	ENE Co., Ltd. — *SuperMAK*	**Hollow fibre**
Ecologix — *EcoPlate*™, *EcoSepro* ™	GE Zenon — *ZeeWeed*®	Ultra-flo®/Mann and Hummel
Huber — *VRM*®; *ClearBox*®, *Biomem*	Hangzhou H-Filtration Memb. Technol. & Engng. Co., Ltd. — *MR*	Polymem — *IMMEM*
Hyflux — *Petaflex*	Koch Membrane Systems — *PURON*®	**Flat disc ceramic**
Jiangsu Lantian Peier Memb. Co. Ltd	Korea Membrane Separations — *KSMBR*	Kerafol
LG Electronics — *Green Membrane*	(Hainan) Litree Purifying Technol. Co. Ltd. — *LH3*	Grundfos — *Biobooster*
Kubota — *ES/EK*	MEMOS Membranes Modules Systems — GmbH — *MEMSUB*	
MICRODYN-NADIR — *BioCel*®	Memstar Technol. Ltd — *SMM*	
Pure Envitech Co., Ltd. — *ENVIS*	Micronet Porous Fibers S.L. — *Micronet*®	
Shanghai Megavision Memb. Enging. and Technol. Co., Ltd.	Mitsubishi Rayon Engng. *Sterapore —SUR*™; *SADF*™	
Shanghai SINAP Membrane Science & Technology Co., Ltd.	Mohua Technology - *iMEM*-25	
Toray — *MEMBRAY*® *TMR*	(Tianjin) Motimo — *Flat Plat FPII*	
Suzhou Vina Filter Co. — *VINAP*	Philos Co. Ltd.	
Weise Water Systems GmbH — *MicroClear*®	SENUO Filtration Technol. Co., Ltd. — *SENUOFIL*	
Other developing technologies	Shanghai Dehong Biology Medicine Sci. & Technol. Dev. Co., Ltd.	
Inge — *FiSh*	Siemens Water Tech. — *MemPulse*™	
IWHR	Sumitomo Electric Industries — *POREFLON* ™	
	Superstring MBR Technol. Corp. — *SuperUF*	
	Suzhou Vina Filter Co. — *F08*	
	Zena SRO — *P5*	

Other products, mainly from China and all HF apart from that indicated (as FS):
Beijing Tri-High Membrane Technology Co., Dow - Omex, Moenda Group Fluid Equipment & Engineering Co. Ltd, Hangzhou Jeffel Membrane Technol. Ltd., Hangzhou Tianchuang Waterpure Equipment Co. Ltd., Jiangsu Dafu Membrane Tchnology Co. Ltd. (FS), Ningbo Jingyuan Membrane Technology Co. Ltd., Shandong Zhaojin Motian Co. Ltd. (Zhao Jin Motian) Shanghai De Yuan Science & Technology Development Co. Ltd., Wuxi Pegasus Membrane Engineering Co — *FMBR-A*.

TABLE C.2 Terminology for MBR Membrane Products and Technologies

Component	Flat Sheet (FS)	Hollow Fibre (HF) Rectangular Panel	Hollow Fibre (HF) Cylindrical Bundle; Multitube (MT)
Element	Panel, element	Module, panel, sub-unit*	Module, bundle
Collection of elements	Module, cassette, unit	Rack, cassette, cartridge*, skid	Skid, frame
	Stack, frame		

**Only used for the KMS (Korea Membrane Separation) technology.*

Web address for whom is provided in the 'Contributors' section at the front of this book.

Flat Sheet Membranes

Kubota

Membrane or Module Proprietary Name/model	510 Panel ES/EK Module	515 Panel RM/ RW Module
Membrane material	Chlorinated PE	
Pore size, μm	0.4 max., 0.2 ave.	
Panel dimensions, length × width × thickness, mm	1020 × 490 × 6, single nozzle	1560 × 575 × 6, dual nozzle
Panel effective membrane area, m^2	0.8	1.45
Module dimensions, length × width × height, mm	1300 × 510 × 180, *LF10*	
	1140–2920 × 600–620 × 2030, *ES*	2250–2930 × 575 × 2490, *RM*
	2200–2920 × 600–620 × 3500, *EK*	2250–2930 × 575 × 4290, *RW*
Number of panels per unit*	10, *LF10*	
	75–200, *ES*	150–200 *RM*
	300–400, *EK*	300–400, *RW*
Total membrane area per unit*, m^2	8, *LF10*	
	60–160, *ES*	217–290, *RM*
	240–320, *EK*	435–580, *RW*
Clean water permeability, LMH/bar	1386	
Maximum operating transmembrane pressure, bar	0.2	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.75 (*ES*); 0.53 (*EK*)	0.42 (*RM*); 0.29 (*RW*)

NB: Module employed by BUSSE GmbH: 600 × 380 mm (0.4 m^2), 12 panels per module (3 m^2), 600 × 440 × 200 mm.
**Referred to as 'submerged membrane unit, SMU' by the suppliers.*

A3 Water Solutions GmbH

Membrane or Module Proprietary Name/Model	M70	U70
Membrane material	PVDF	PES
Pore size, μm, or MWCO in kDa	0.14 μm	150 (0.038 μm)
Panel dimensions, length × width × thickness, mm	700 × 1040 × 6	
Panel effective membrane area, m^2	1.36	
Panel separation, mm	7	
Module dimensions, length × width × depth, mm	736 × 1070 × 716	
Number of panels per module	51	
Total membrane area per module, m^2	70	
Clean water permeability, LMH/bar	>1000	>300
Maximum operating transmembrane pressure, bar	0.25	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.685	

Backflushable at 50 mb.

Alfa Laval Environment Technology

Membrane or Module Proprietary Name/Model	Hollow Sheet, MFM 100, 200, 300
Membrane material	PVDF
Pore size, μm	0.2
Element dimensions, length × width × thickness, mm	–
Element effective membrane area, m^2	1.81
Element separation, mm	7
Module dimensions, length × width × depth, mm	1172 × 1194 × 1988, 3080, 4171
Number of elements per module	85, 170, 255
Total membrane area per module, m^2	154, 308, 462
Clean water permeability, LMH/bar (membrane only)	4000
Maximum operating transmembrane pressure, bar	0.08 (average: 0.01–0.04)
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.48–0.54, 0.3–0.36, 0.24–0.3

Brightwater FLI

Membrane or Module Proprietary Name/Model	MEMBRIGHT®
Membrane material	PES
Pore size, kDa (μm)	150 (0.038 μm)
Panel dimensions, length × width × thickness, mm	950 × 950 × 7
Panel effective membrane area, m^2	1.84
Panel separation, mm	~9
Module dimensions, length × width × depth, mm	1120 × 715, 1215, 1450
Number of panels per module	25, 50
Total membrane area per module, m^2	46, 92
Clean water permeability, LMH/bar	150
Maximum operating transmembrane pressure, bar	0.35
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.69

Colloide Engineering Systems

Membrane or Module Proprietary Name/Model	SubSnake
Membrane material	PES
Pore size, kDa (μm)	150 (0.038 μm)
Panel dimensions, width × thickness, mm	1000 × 5
Panel effective membrane area, m^2	10
Panel separation, mm	10
Module dimensions, length × width × depth, mm	–
Number of panels per module	5
Total membrane area per module, m^2	160
Clean water permeability, LMH/bar	–
Maximum operating transmembrane pressure, bar	NA
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	–

Ecologix

Membrane or Module Proprietary Name/Model	EcoPlate™	EcoSepro™
Membrane material	PVDF, PES	
Pore size, μm	0.08, 0.1 and 0.4	0.08, 0.1 and 0.4
Panel dimensions, height × width × thickness, mm	1000 × 490 × 6	1000 × 320 × 3
Panel effective membrane area, m^2	0.8	0.6
Panel separation, mm	6	
Module dimensions, length × width × height (single deck; double deck), mm	820–3150 × 610–660 × (2100;3200)	310–1810 × 700 × (1800;2820)
Number of panels per module: single deck; double deck	50–200; 100–400	60–720
Total membrane area per module, m^2: single deck; double deck	40–160; 80–320	36–432
Clean water permeability, LMH/bar	NA	
Maximum operating transmembrane pressure, bar	0.3	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.72–1.0	0.5–0.8, continuous

Membrane material supplied by Sepro Membrane, USA.

Huber Technology Inc.

Membrane or Module Proprietary Name/Model	Vacuum Rotation Membrane (VRM® 20, VRM® 30)	Membrane ClearBox® (MCB)
Membrane material	PES on PP backing	PES on PES (laminate)
Pore size, kDa (μm)	150 (~0.038 μm)	
Panel dimensions, length × width × thickness, mm	1000 × 1000 × 8*	800 × 400 × 4
Panel effective membrane area, m^2	0.75 (*VRM® 20*), 1.5 (*VRM® 30*)	0.38

(*Continued*)

Membrane or Module Proprietary Name/Model	Vacuum Rotation Membrane (VRM® 20, VRM® 30)	Membrane ClearBox® (MCB)
Panel separation, mm	6	8
Module dimensions, length × width × height, mm	1000 × 1000 × 52*	800 × 400 × 100**
Number of panels per module	4	9**
Total membrane area per module, m^2	1.5 (*VRM® 20*), 6 m^2 (*VRM® 30*)	3.5
Number of panels per plate	6 or 8 sub-modules	–
Plate area, m^2	9 (*VRM® 20*), 48 (*VRM® 30*)	–
Maximum number of plates per stack	80	–
Maximum stack area	3840	–
Clean water permeability, LMH/bar	>1000	
Maximum operating transmembrane pressure, bar	<0.5	0.7
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.15 – 0.25	0.6 –0.8

The VRM is a rotating membrane module with hexagonal or octagonal geometry.
**Trapezoidal.*
***Smallest unit, MCB 1.*

Hyflux

Membrane or Module Proprietary Name/Model	PetaFlex™ (100S)	
Model	***PTF316***	***PTF304***
Membrane material	PVDF	
Pore size, mm, or MWCO in kDa	0.1	
Panel dimensions, length × width × thickness, mm	–	–
Panel effective membrane area, m^2	–	–
Panel separation, mm	–	–
Module dimensions, length × width × height, mm	2310 × 790 × 4300	2500 × 1000 × 4300
Number of panels per skid	250*	
Total membrane area per module, m^2	690	
Clean water permeability, LMH/bar	–	
Maximum operating transmembrane pressure, bar	0.6	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.28	

**Five panels per cassette, 50 cassettes per skid.*

Jiangsu Lantian Peier Memb. Co., Ltd

Membrane or Module Proprietary Name/Model	PEIER (100, 150 & 175: three models)
Membrane material	PVDF
Pore size, µm	0.1–0.3
Panel dimensions, length × width × thickness, mm	1190, 1780, 2000 × 518 × 15*

Membrane or Module Proprietary Name/Model	PEIER (100, 150 & 175: three models)
Panel effective membrane area, m^2	1, 1.5, 1.75
Panel separation, mm	7*
Module dimensions, length × width × depth, mm	1650 (2350)** × 605 × 2000, 2660, 2900
Number of panels per module	100, 100–150, 100
Total membrane area per module, m^2	100, 150–225, 175
Clean water permeability, LMH/bar	–
Maximum operating transmembrane pressure, bar	0.1 (operating TMP)
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.72, 0.48, 0.41

*15 mm dimension assumed to contain both panel thickness and membrane separation.
**2350 mm is the length of the 150-plate module 0.25 m^2 panel also provided.

LG Electronics

Membrane or Module Proprietary Name/Model	LG Green Membrane
Membrane material	PES
Pore size, μm	0.01–0.2
Panel dimensions, length × width × depth, mm	1200 × 490 × 4
Panel effective membrane area, m^2	1
Panel separation, mm	7
Module dimensions, length × width × height, mm	2603 × 600 × 1730
Number of panels per module	100
Total membrane area per module, m^2	100
Clean water permeability, LMH/bar	>8000
Maximum operating transmembrane pressure, bar	0.6
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	0.6–0.9

MICRODYN-NADIR, GmbH – *BIO-CEL®*

Membrane or Module Proprietary Name/Model	BIO-CEL® (BC100)	BIO-CEL® (BC400)
Membrane material	PES	
Pore size, μm (MWCO, kDa)	0.04 (~150 kDa)	
Panel dimensions, length × width × thickness, mm	1000 × 500 × 2	2000 × 1000 × 2
Panel effective membrane area, m^2	1	4
Panel separation, mm	7	
Module dimensions, length × width × height, mm	1270 × 702 × 1563	1298 × 1152 × 2763
Number of panels per cassette	25	
Total membrane area per module, m^2	100	400
Clean water permeability, LMH/bar	350	
Maximum operating transmembrane pressure, bar	0.4	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.3–0.6	0.2–0.4

Pure Envitech

Membrane or Module Proprietary Name/Model	ENVIS
Membrane material	–
Pore size, μm	0.4
Panel dimensions, length × width × thickness, mm	1000 × 560 × 8
Panel effective membrane area, m^2	0.98
Panel separation, mm	–
Module dimensions, length × width × depth, mm	–
Number of panels per module	15*
Total membrane area per module, m^2	15
Clean water permeability, LMH/bar	–
Maximum operating transmembrane pressure, bar	0.79
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	–

**Assumed from available information.*

Shanghai MegaVision™ Membrane Engineering & Technology Co., Ltd

Membrane Module Proprietary Name/Model	FMBR-1.0-100
Membrane material	PVDF or PES
Pore size, μm	0.1 or 0.3
Panel dimensions, length × width × thickness, mm	930 × 610 × 16
Panel effective membrane area, m^2	1.0
Panel separation, mm	16
Module dimensions, length × width × height, mm	1800 × 715 × 1770
Number of panels per module	100
Total membrane area per module, m^2	100
Clean water permeability, LMH/bar	–
Maximum working operating pressure, bar	~0.3
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	0.75

Shanghai Sinap Membrane Science & Technology Co., Ltd

Membrane or Module Proprietary Name/Model	SINAP (25)	SINAP (80–100)	SINAP (150–150)
Membrane material	PVDF		
Pore size, μm	0.1		
Panel dimensions, length × width × thickness, mm	470 × 340 × 7	1000 × 490 × 7	1800 × 510 × 10
Panel effective membrane area, m^2	0.25	0.8	1.5
Panel separation, mm	7		
Module dimensions, length × width × height, mm	–	1600 × 650 × 2000	2350 × 650 × 3000
Number of panels per module	–	100	150
Total membrane area per module, m^2	–	80	225
Clean water permeability, LMH/bar	–	–	–
Maximum operating transmembrane pressure, bar	0.4		
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	–	>0.9	>0.48

Panel of 0.1 m^2 also provided.

Toray Industries Inc.

Membrane or Module Proprietary Name/Model	MEMBRAY® TMR140-100S	MEMBRAY® TMR090-100S
Membrane material	PVDF	PVDF
Pore size, µm	0.08	0.08
Panel dimensions, length × width × thickness, mm	1608 × 515 × 6.5	1059 × 515 × 6.5
Panel effective membrane area, m^2	1.4	0.9
Panel separation, mm	7	
Module dimensions, length × width × height, mm	1620 × 810 × 2100	1720 × 730 × 1470
Number of panels per module	100	
Total membrane area per module, m^2	140	90
Clean water permeability, LMH/bar	—	—
Maximum operating transmembrane pressure, bar	0.2	0.2
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.56	0.67
Other module types	50, 200 plate module	50 plate module

Suzhou Vina Filter Co., Ltd

Membrane or Module Proprietary Name/Model	VINAP-150
Membrane material	PVDF
Pore size, µm, or MWCO in kDa	0.05
Panel dimensions, length × width × thickness, mm	1780 × 510 × 6
Panel effective membrane area, m^2	1.5
Panel separation, mm	6–8
Module dimensions, length × width × depth, mm	—
Number of panels per module	—
Total membrane area per module, m^2	—
Clean water permeability, LMH/bar	550
Maximum operating transmembrane pressure, bar	0.2
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.4–0.6

Weise Water Systems GmbH

Membrane or Module Proprietary Name/Model	Microclear® (MC03)	Microclear® (MCXL)	MA04-150 (largest stack)
Membrane material	PES	PES	
Pore size, mm, or MWCO in kDa	0.05	0.05	
Panel dimensions, height × width × thickness, mm	492 × 165 × 2	490 × 375 × 3.5	
Panel (module) effective membrane area, m^2	0.146	0.333	(7)
Panel separation, mm	5.5	6	

(*Continued*)

Membrane or Module Proprietary Name/Model	Microclear® (MC03)	Microclear® (MCXL)	MA04-150 (largest stack)
Module dimensions, height × width × depth, mm	492 × 207 × 207	490 × 415 × 207	
Number of panels per module (modules per stack)	24	21	(75)
Total membrane area per module (stack), m^2	3.5	7	(525)
Clean water permeability, LMH/bar	>200	>200	
Maximum operating transmembrane pressure, bar	0.3	0.3	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	1.1*, 2.3**	0.76#, 1.1†	0.76

**MA03-2, double deck.*
***MA03-1, single deck.*
#MA04-12 to 150, triple deck.
†MA03-8 to 100, double deck.

Hollow Fibre Membranes

GE Water & Process Technologies (Formerly Zenon)

Membrane or Module Proprietary Name/Model	ZeeWeed® (ZW500C)	ZeeWeed® (ZW500D)
Membrane material	PVDF, braided	
Pore size, µm	0.04	
Filament outside diameter, mm	1.9	
Module dimensions, length × width × depth, mm	1923 × 720 × 60	(a) 1837 × 844 × 49 (b) 2198 × 844 × 49
Module effective membrane area, m^2	23.2	(a) 25.5 & 27.9 (b) 31.6 & 34.4
Cassette dimensions, length × width × depth, mm	(a) 992 × 743 × 2085 (b) 1828 × 743 × 2085 (c) 2668 × 743 × 2085	(a) 1744 × 738 × 2208 (b) 1744 × 738 × 2568 2122 × 1745 × 2590
Number of modules per cassette	(a) 10 (b) 22 (c) 32	(a) 16 (b) 16, 48
Total membrane area per module, m^2	(a) 232 (b) 510.4 (c) 742.4	(a) 408, 446.4 (b) 505.6, 550.4 1516.8, 1651.2
Packing density, m^2 membrane area/m^3 internal module volume	(a) 215 (b) 214 (c) 214	
Clean water permeability, LMH/bar	375	
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	(a) 0.36 (b) 0.34 (c) 0.26	(a) 0.42, 0.39 (b) 0.34, 0.31 0.31, 0.29

Asahi Kasei Chemicals Corporation

Membrane or Module Proprietary Name/Model	Microza® MUNC-620A
Membrane material	PVDF
Pore size, µm	0.1
Filament outside diameter, mm	1.3
Module dimensions, length × diameter, mm	2163 × 167
Module effective membrane area, m^2	25
Rack dimensions, length × width × depth, mm	1400 × 920 × 2900
Number of modules per rack	24
Total membrane area per rack, m^2	600
Packing density, m^2 membrane area/m^3 internal module volume	530
Clean water permeability, LMH/bar	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.2–0.28
Maximum operating transmembrane pressure, bar	0.8

Beijing Origin Water Technology Company

Membrane or Module Proprietary Name/Model	Membrane: BSY, RF; Module: MBRU
Membrane material	PVDF
Pore size, µm, or MWCO in kDa	MF: 0.1, 0.3; UF 100 kDa
Filament outside diameter, mm	*BSY*: 2.0; *RF*: 2.4
Module dimensions, length × width × thickness, mm	2000 × 1250 × 30
Module effective membrane area, m^2	*BSY-III* 26.5; *RF-III* 27.5
Cassette dimensions, length × width × height, mm	3334 × 1760 × 3076
Number of modules per cassette	60
Total membrane area per cassette, m^2	*BSY*:1602; *RF*: 1650
Packing density, m^2 membrane area/m^3 internal volume	450
Clean water permeability, LMH/bar	*BSY*: 2000–3000; *RF*: >3000
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	*BSY*: 130 (120–140); *RF*: 140 (130–150)

Canpure

Membrane or Module Proprietary Name/Model	Saveyor SVM 640, 660, 680
Membrane material	PVDF
Pore size, µm	0.075
Filament outside diameter, mm	1.3
Element dimensions, width × depth × height, mm	158 × 158 × 997, 1597, 2097
Element effective membrane area, m^2	14, 22, 30
Packing density, m^2 membrane area/m^3 internal element volume	562
Maximum backwash pressure, bar	0.05-0.05
pH range	Up to 12
Max free chlorine, mg/L	1000

Ecologix

Membrane or Module Proprietary Name/ Model	EcoFil™ (AK12-A)*	EcoFlon™ (EF12)**
Membrane material	PVDF	PTFE
Pore size, µm, or MWCO in kDa	0.1	
Filament outside diameter, mm	1.2 or 1.3	1.3
Module dimensions, length × diameter, mm	1500–3000 × 200	1500–3000 × 200
Module effective membrane area, m^2	14.4, 21.6, 28.8	
Stack dimensions, length × width × depth, mm	–	–
Number of modules per skid	–	–
Total membrane area per module, m^2	–	–
Packing density, m^2 membrane area/m^3 internal module volume	559:458	458
Clean water permeability, LMH/bar	–	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.4	
Maximum operating transmembrane pressure, bar	0.3	

**Membrane material supplied by Ecologix, USA.*
***Membrane material supplied by Markel, USA.*

ENE (Energy and Environment)

Membrane or Module Proprietary Name/Model	SuperMAK (SM-10)
Membrane material	PVDF
Pore size, µm	0.4
Filament outside diameter, mm	2
Module dimensions, length × width × height, mm	400 × 160 × 720
Module effective membrane area, m^2	10
Stack dimensions, length × width × height, mm	1400 × 1000 × 800*
Number of modules per stack	14
Total membrane area per module, m^2	140
Packing density, m^2 membrane area/m^3 internal element volume	217
Clean water permeability, LMH/bar	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.6
Maximum operating transmembrane pressure, bar	0.52

**Can be stacked to make a 1600-mm-high module.*

Hangzhou H-Filtration Membrane Technology & Engineering Co., Ltd

Membrane or Module Proprietary Name/Model	MR
Membrane material	PP
Pore size, µm	0.1
Filament outside diameter, mm	0.45
Module dimensions, length × width × depth, mm	810 × 525 × 55
Module effective membrane area, m^2	8
Stack dimensions, length × width × depth, mm	–
Number of modules per module/stack	–
Total membrane area per stack, m^2	–

Membrane or Module Proprietary Name/Model	MR
Packing density, m^2 membrane area/m^3 internal module volume	160
Clean water permeability, LMH/bar	120
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.1–0.16
Maximum filtration pressure, bar	0.3

Element packing density1073 m^2/m^3; ~12,000 fibres per module. Effective fibre width in module is 460 mm.
Horizontal orientation; triple deck possible.

KMS Co., Ltd (Korea Membrane Separation)

Membrane or Module Proprietary Name/Model	KMS-CF Series
Membrane material	HDPE
Pore size, μm	0.4
Filament outside diameter, mm	0.65
Cartridge* dimensions, length × width × height, mm	536 × 320 × 396
Cartridge effective membrane area, m^2	18
Stack dimensions, length × width × height, mm	784–2956 × 1186–1326 × 1520–3240
Number of cartridges per frame	12–112
Total membrane area per frame, m^2	216–2016
Packing density, m^2 membrane area/m^3 internal module volume	265 for cartridge; 159 for stack
Clean water permeability, LMH/bar	9000**
Recommended membrane aeration rate, Nm^3/h per m^2 stack projected area	75

**13 Sub-units of 1.385 m^2 per cartridge.*
***1500 mm Length in-out pressure test.*

Koch Membrane Systems

Module Designation	PURON® Submerged Membrane Module		
	PSH 250	PSH 500	PSH 1500
Membrane material	PES, braided		
Pore size, μm	0.05		
Filament outside diameter, mm	2.6		
Bundle dimensions, length × diameter, mm	1830 × 70		1990 × 70
Bundle effective membrane area, m^2	3.47		3.78
Module dimensions, length × width × height, mm	906 × 893 × 2385	1662 × 893 × 2422	2244 × 1755 × 2530
Number of bundles per module	72	144	396
Total membrane area per module, m^2	250	500	1500
Packing density, m^2/m^3	129.7	139.2	150.5
Clean water permeability, LMH/bar	≥500		
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area*	0.15–0.30		0.133 – 0.267

**Average values over a whole aeration cycle: aeration capacity can be modulated according to duty.*

Hainan Litree Purifying Tech. Co., Ltd

Membrane or Module Proprietary Name/Model	LJ1E-1500-F180
Membrane material	PVDF
Pore MWCO, kDa	150
Filament outside diameter, μm	1.8
Module dimensions, length × diameter, mm	–
Module effective membrane area, m^2	18–35
Skid dimensions, length × width × height, mm	721 × 70 × 1187–2087*
Number of modules per stack	–
Total membrane area per module, m^2	18–35
Packing density, m^2 membrane area/m^3 internal module volume	–
Clean water permeability, LMH/bar	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	–
Aeration cycle (if intermittent), on:off, min	–
Maximum filtration pressure, bar	0.6

**Assumed to be 1100–2000 membrane length.*

Memstar Technology Ltd

Membrane or Module Proprietary Name/Model	SMM-1525*
Membrane material	PVDF
Pore size, μm, or MWCO in kDa	<0.1 μm
Filament outside diameter, mm	1.2
Module dimensions, length × width × height*, mm	571 × 45 × 815–1535*
Module effective membrane area, m^2	10–25*
Skid dimensions, length × width × depth, mm	2410 × 1430 × 3500, 2070 × 1430 × 1850
Number of modules per skid	40, 96
Total membrane area per module, m^2	800, 1920
Packing density, m^2 membrane area/m^3 internal module volume	507 (159 for skid)
Clean water permeability, LMH/bar	350
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.05–0.15

**Models SMM-1010, SMM-1013 and SMM-1520 also supplied, providing a range of module lengths and thus areas.*

Micronet Porous Fibers, S.L.

Membrane or Module Proprietary Name/Model	Micronet® (R-MF)	Micronet® (R-UF)
Membrane material	PVDF	PVDF
Pore size, μm	0.2	0.05
Filament outside diameter, mm	2.45	2.1
Module dimensions, length × width × depth, mm	1935 × 106 × 106	

Membrane or Module Proprietary Name/Model	Micronet® (R-MF)	Micronet® (R-UF)
Module effective membrane area, m^2	6 – 7.5	
Stack dimensions, height × width × depth, mm	2590 × 2375 × 1020	
Number of modules per stack	91	
Total membrane area per module, m^2	540–750	
Packing density, m^2 membrane area/ m^3 internal module volume	–	
Clean water permeability, LMH/bar	1200	1500
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.4–0.8	0.3–0.8

Mitsubishi Rayon Engineering Co., Ltd

Membrane or Module Proprietary Name/Model	STERAPORESUR™*	STERAPORESADF™
Membrane material	PE	PVDF
Pore size, µm	0.4	
Filament outside diameter, mm	0.54	2.8
Module dimensions, length × width × thickness, mm	1035 × 524 × 13	2000 × 1250 × 30
Module effective membrane area, m^2	3	25
Cassette dimensions, length × width × height, mm	1442 × 1538 × 725	1610 × 1555 × 3124
Number of modules per cassette	70	20
Total membrane area per cassette, m^2	210	500
Packing density, m^2 membrane area/m^3 internal module volume	426	333
Clean water permeability, LMH/bar	–	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	–	–

**Stackable horizontally oriented membrane module.*

Tianjin Motimo Membrane Technology Co., Ltd

Membrane or Module Proprietary Name/Model	Flat Plat (FP II)
Membrane material	PVDF
Pore size, µm	0.2, 0.1
Filament outside diameter, mm	1.2
Module dimensions, length × width, mm	1523 × 534
Module effective membrane area, m^2	20
Stack dimensions, length × width × depth, mm	2000 × 1400 × 1700
Number of modules per stack	40
Total membrane area per module, m^2	800
Packing density, m^2 membrane area/m^3 internal element volume	190
Clean water permeability, LMH/bar	200
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	0.15
Maximum filtration pressure, bar	0.7

Philos Co., Ltd

Membrane or Module Proprietary Name/Model	Philosep (RCM)
Membrane material	PVDF, braided
Pore size, µm	0.1
Filament outside diameter, mm	2.35
Bundle diameter, mm	75*
Bundle effective membrane area, m^2	1.2
Skid dimensions, length × width × height, mm	700–1330 × 620 × 1980
Number of bundles per skid	42–105
Total membrane area per skid, m^2	51–126
Packing density, m^2 membrane area/m^3 internal element volume	170*
Clean water permeability, LMH/bar	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	–
Aeration cycle (if intermittent), on:off, min	–
Maximum filtration pressure, bar	0.52

**Estimated.*

Senuofil Filtration Technology (Tianjin) Co., Ltd

Membrane or Module Proprietary Name/Model	Senuofil (SN-MBR-0660)	Senuofil (SN-MBR-0680)
Membrane material	PVDF, PES	
Pore size, µm	0.1	
Filament outside diameter, mm	1.3	
Membrane module area, m^2	15	20
Membrane module dimensions, diameter × length, mm	160 × 1640	160 × 2150
Packing density, m^2 membrane area/m^3 internal element volume*	487	
Number of modules per unit	8	
Total membrane area per module, m^2	120	
Module dimensions, length × width × height, mm	1180 × 560 × 2060	
Recommended membrane aeration rate, Nm^3/h per m^2	0.267	
Aeration cycle (intermittent), on:off, min	–	
Maximum filtration pressure, bar	0.4**	

**Assumes end lengths of 110 mm.*
***Maximum backflush pressure of 2 bar.*

Shanghai Dehong Biology Medicine Science Technology Development Co., Ltd.

Membrane or Module Proprietary Name/Model	DH-MBR-LSE01 Series
Membrane material	PVDF
Pore size, µm	0.06–0.08
Filament outside diameter, mm	1.3
Bundle dimensions, diameter × length, mm	32–50 × 500–2000

Membrane or Module Proprietary Name/Model	DH-MBR-LSE01 Series
Element/module effective membrane area, m^2	1
Skid dimensions, length × width × height, mm	1060 (1000) × 1600 (1500) × 2120 (2000)*
Number of bundles per skid	100
Total membrane area per module, m^2	100
Packing density, m^2 membrane area/m^3 internal module volume	1450 at the pot, ~49 in bulk skid
Clean water permeability, LMH/bar	800–1600
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	15–20:1
Maximum filtration pressure, bar	0.5

**Effective membrane dimension provided in parentheses.*

Siemens Water Technologies Corp.

Membrane or Module Proprietary Name/Model	Memcor® (B40N*), Memjet®, Mempulse®
Membrane material	PTFE
Pore size, μm	0.04
Filament outside diameter, mm	1.3
Module dimensions, length × width × depth, mm	1600** × 203 × 203
Module effective membrane area, m^2	38
Rack dimensions, length × width × height, mm	3960 × 280 × 2220**
Number of elements/module per module/rack	16
Total membrane area per module, m^2	608
Packing density, m^2 membrane area/m^3 internal module volume	580
Clean water permeability, LMH/bar	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	–
Maximum filtration pressure, bar	0.2

Information taken from MC B40N DS 0709 and MGS400tps10 01 SpecSheet.doc (Siemens, 2009).
**B30R module also available.*
***Rack height includes headers and Mempulse™ unit.*

Sumitomo Electric Fine Polymer, Inc.

Membrane or Module Proprietary Name/Model	POREFLON™ (SPMW-05B10)	POREFLON™ (SPMW-06B10)
Membrane material	PTFE	PTFE
Pore size, μm	0.2	0.1
Filament outside diameter, mm	2.3	
Module dimensions, length × width × depth, mm	2410 × 164 × 154	
Module effective membrane area, m^2	10*	
Stack dimensions, height × width × depth, mm		
• 100 m^2 stack	3900 (2881) × 1050 (1030) × 840 (344)²	
• 200 m^2 stack	3900 (2881) × 2280 (1880) × 840 (344)**	

(*Continued*)

Membrane or Module Proprietary Name/Model	POREFLON™ (SPMW-05B10)	POREFLON™ (SPMW-06B10)
Number of modules per stack	10, 20	
Total membrane area per module, m^2	100, 200	
Packing density, m^2 membrane area/ m^3 internal module volume	98, 107	
Clean water permeability, LMH/bar	1500	500
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.3	
Maximum filtration pressure, bar	0.6	

**6 m^2 module also available.*
***Internal dimensions in parentheses.*

Superstring MBR Technology Corp.

Membrane or Module Proprietary Name/Model	SuperUF (10, 20 or 40)
Membrane material	PP
Pore MWCO, kDa	100*
Filament outside diameter, mm	1.25
Module dimensions, length × width × thickness, mm	1140 × 770 × 25**
Module effective membrane area, m^2	1
Stack dimensions, length × width × height, mm	330 × 770 × 1140
Number of modules per stack	10
Total membrane area per stack, m^2	10
Packing density, m^2 membrane area/m^3 total module volume	35#
Clean water permeability, LMH/bar	1500–2000
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	0.9–1.8
Maximum filtration pressure, bar	0.3

**0.2–0.35 μm pore size, according to bubble-point test.*
***Thickness estimated on basis of 8 mm panel separation.*
#Refers to external module volume.

Suzhou Vina Filter Co., Ltd

Membrane or Module Proprietary Name/Model	VINAFREE-15	PP-800
Membrane material	PVDF	PP
Pore size, mm, or MWCO in kDa	0.1	0.2
Filament outside diameter, mm	1.2 (0.8)	0.45
Bundle dimensions, length × diameter, mm	1640 (1540)* × 160	800 (750)* × 25
Bundle effective membrane area, m^2	15	1
Module dimensions, length × width × depth, mm	–	–
Number of bundles per module	–	–
Total membrane area per module, m^2	–	–
Packing density, m^2 membrane area/m^3 internal element volume	455	2500**
Clean water permeability, LMH/bar	600	240
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	0.3–0.5	0.2–0.6

Membrane or Module Proprietary Name/Model	VINAFREE-15	PP-800
Aeration cycle (if intermittent), on:off, min	8:2	60:1
Maximum filtration pressure, bar	0.5	0.3

**Effective membrane length provided in parentheses.*
***Per bundle: 1000 fibres in a 25 mm dia. × 800 mm long bundle of 0.45 mm dia. fibres.*

Zena SRO

Membrane or Module Proprietary Name/Model	P5S
Membrane material	PP
Pore size, μm	0.1
Filament outside diameter, mm	0.26
Bundle dimensions, diameter × length, mm	25 × 821
Bundle effective membrane area, m^2	0.8
Skid dimensions, length × width × depth, mm	784 × 590 × 1010
Number of bundles per module/stack	108
Total membrane area per module, m^2	86
Packing density, m^2 membrane area/m^3 internal module volume	1630 (184 for skid)
Clean water permeability, LMH/bar	–
Recommended membrane aeration rate, Nm^3/h per m^2 membrane	–
Maximum filtration pressure, bar	–

Multitube Membranes

Norit X-Flow BV

Membrane or Module Proprietary Name/Model	38GRH F5385*	38PRV F4385**
Membrane material	PVDF	
Pore size, μm	0.08	0.03
Membrane tube internal diameter, mm	8	5.2#
Module dimensions, length × diameter, mm	3000 × 203	
Module membrane area, m^2	27	33
Skid dimensions, length × width × depth, m	3800 × 1000 × 4000 (16 modules max)	3800 × 1700 × 4500 (303 modules max)
Number of modules per skid	16 max: 2 × 8 max in series	30 max in parallel
Total membrane area of skid, m^2	432 max	990 max
Clean water permeability, LMH/bar	>750	
Normal recommended range of crossflow velocity, m/s	2–4	0.4
Normal recommended transmembrane pressure, bar	0.7–5	0.1–0.4
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	NA	0.3

**For pumped sidestream system.*
***For air-lift sidestream system.*
#3 mm ID 'Megablock' system introduced in 2010: 21 modules max.

BERGHOF Membrane Technology GmbH & Co. KG

MBR Process Technology	**BioAir DS**
Membrane or Module Proprietary Name/Model	**HyperFlux I8LE**
Membrane material	PVDF or PES
Pore size, μm	0.03*
Membrane tube internal diameter, mm	8**
Module dimensions, length × diameter, mm	4000 × 254
Module membrane area, m^2	53.4
Rack dimensions, length × width × height, mm	3100 × 1500 × 4900
Number of modules per rack	10
Total membrane area of rack, m^2	534
Normal range of permeability, LMH/bar	75–250
Normal recommended range crossflow velocity employed, m/s	0.3–0.5
Normal recommended transmembrane pressure, bar	0.2–0.5
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	0.04–0.06

**Other sizes down to 5 kDa.*
***Other sizes: 5, 8, 10 , 11.5 , 12.5 , 12.7 mm.*

Other Sidestream Products: HF

Polymem

Membrane or Module Proprietary Name/Model	**Immem (WW120)**
Membrane material	PS
Pore size, μm (MWCO in kDa)	0.08 (300 kDa)
Filament diameter, mm	0.7–1.4
Module dimensions, length × diameter, mm	1000–1500 × 315
Module area, m^2	60–100 m^2
Clean water permeability, LMH/bar	500

Information taken from The MBR Book, First Edition.

Ultra-Flo Pte Ltd

Membrane or Module Proprietary Name/Model	**Ultra-Flo (U860)** **MANN + HUMMEL**
Membrane material	PAN
Pore size, μm	0.1
Filament outside diameter, mm	2.0
Module dimensions, length × diameter, mm	1524 × 203
Module effective membrane area, m^2	48
Module/stack dimensions, length × width × depth	–
Number of elements/module per module/stack/skid	1–50
Total membrane area per module, m^2	(1–50) × 48; 2400 max
Packing density, m^2 membrane area/m^3 internal element volume	840 m^2/m^3
Clean water permeability, LMH/bar	72 LMH
Recommended membrane aeration rate, Nm^3/h per m^2 membrane area	Intermittent on/off
Aeration cycle (if intermittent), on:off, min	Variable

Membrane or Module Proprietary Name/Model	Ultra-Flo (U860) MANN + HUMMEL
Maximum filtration pressure, bar	< 5 psi (<0.34 bars)
Recommended flux, LMH	8.5 LMH (5 gfd)
pH range	2–12
Max operating temperature	50 °C

Other Sidestream Products: FS

Orelis Environment SAS

Membrane or Module Proprietary Name/model	Pleiade®
Membrane material	PAN
Pore MWCO, kDa	40 kDa
Element dimensions, length × width, mm	2610 × 438
Element area, m^2	1.15
Membrane separation, mm	3
Module dimensions, length × width × depth, mm	2610 × 438 × 1710
Number of elements per module	–
Total membrane area per external module volume, m^2/m^3	36

Information taken from Judd (2006).

Kerafol GmbH

Membrane or Module Proprietary Name/model	Kerafol
Membrane material	Al_2O_3
Pore size, μm	2.0, 0.5, 0.2, 0.06, 0.03, 0.007
Disc outer and inner diameter, and thickness, mm	374 or 312, 91, 6
Panel effective membrane area, m^2	0.2, 0.14
Panel separation, mm	–
Module dimensions, height × width × depth, mm	–
Number of panels per module	–
Total membrane area per module, m^2	–
Clean water permeability, LMH/bar	–
Maximum operating transmembrane pressure, bar	5

Abbreviations

Abbreviations of membrane, membrane bioreactor and wastewater terms are provided on the following pages. Definitions of symbols assigned to all applicable parameters are listed in the Nomenclature section. A list of definitions of some of the terms used is given in the Glossary.

ABR Anaerobic baffled reactor
AD Anaerobic digestion
ADUF Anaerobic digester ultrafiltration
AF Anaerobic filter
a-lsMBR Air-lift sidestream membrane bioreactor (aerobic)
Alum Aluminium sulphate
aniMBR Anaerobic immersed membrane bioreactor
anMBR Anaerobic membrane bioreactor
ansMBR Anaerobic sidestream membrane bioreactor
AOC Assimilable organic carbon
AOTE Actual oxygen transfer efficiency
AOTR Actual oxygen transfer rate
ASCE American Society of Civil Engineers
ASP Activated sludge process
AX Anoxic
BAC Biologically activated carbon
BAF Biological aerated filter
BER Biofilm-electrode reactor
BFM Berlin filtration method
BNR Biological nutrient removal
BOD Biochemical oxygen demand
BOD_5 Five-day biochemical oxygen demand
BOO(T) Build, own, operate (transfer)
BPA Biological potential activity
bpCOD Biodegradable particulate COD
CA Cellulose acetate
CAPEX Capital expenditure
CAS(P) Conventional activated sludge (process)
CEB Chemically enhanced backwash
CF(V) Crossflow (velocity)
CFU Colony-forming units
CIA Clean(ing) in air
CIL Clean(ing) in line
CIP Clean(ing) in place
COD Chemical oxygen demand
COP Clean(ing) out of place
CP Concentration polarization

CPR Chemical phosphorus removal
CST Capillary suction time
CSTR Continuous stirred tank reactor
CT Capillary tube
Da Dalton
DE Dead-end (or full flow)
DFCm Delft filtration characterization method
dMBR Diffusion membrane bioreactor
DO Dissolved oxygen
DOC Dissolved organic carbon
DS Dry solids
EBPR Enhanced biological phosphate removal
EC Enterococci
ED Electrodialysis
EDCs Endocrine disrupting compounds
EGSB Expanded granular sludge bed
eMBR Extraction membrane bioreactor
(e)EPS (Extracted) extracellular polymeric substances
EPSc Extracellular polymeric substances (carbohydrate)
EPSp Extracellular polymeric substances (protein)
EQ Equalization
EQI Effluent quality index
FBDA Fine bubble diffused aeration/aerator
FC Filter cartridge
Flocs Flocculated particles
FO Forward osmosis
FOG Fats, oils and grease
FS Flat sheet (or plate and frame)
GAC Granular activated carbon
GLD Gigalitres per day
g-lsMBR Gas-lift sidestream membrane bioreactor (anaerobic)
GRP Glass-reinforced plastic
GT Gas transfer
HF Hollow fibre
HFRB Hair and fibre reinforced biomass
HPSEC High-performance size-exclusion chromatography
HRT Hydraulic retention time
HVAC Heat, ventilation and air conditioning
ID Internal diameter
IEMBR Extractive ion exchange MBR
IEX Ion exchange
iMBR Immersed membrane bioreactor (aerobic)
kDa kiloDalton
LMH Litres per m^2 per hour
LRV Log rejection value
MABR Membrane aeration bioreactor
MBBR Moving bed bioreactor
MBfR Membrane biofilm reactor
MCE Mixed cellulose esters

MD Membrane distillation
ME Membrane extraction
MF Microfiltration
MHBR Membrane hydrogenation bioreactor
MLD Megalitres per day
MLSS Mixed liquor suspended solids
MLVSS Mixed liquor volatile suspended solids
MPE Membrane performance enhancer
MST Membrane Sewage Treatment (Dorr Oliver)
MT Multitube
MW Molecular weight
MWCO Molecular weight cut-off
NADH Nicotinamide adenine dinucleotide hydrogenase
nbVSS Non-biodegradable volatile suspended solids
NF Nanofiltration
NIPS Non-solvent-induced phase separation
NOM Natural organic matter
NPV Net present value
O&M Operation and maintenance
OC Organic carbon
OD Outer diameter
ODE Ordinary differential equation
OEM Original equipment manufacturer
OLR Organic loading rate
ON Organic nitrogen
OPEX Operating expenditure
OTE Oxygen transfer efficiency
OTR Oxygen transfer rate
OUE Oxygen utilisation efficiency
P&F Plate and frame
P&ID Piping and instrumentation diagram
p.e. Population equivalent
PAC Powdered activated carbon
PAN Polyacrylonitrile
pCOD Total particulate COD
PDMS Polydimethylsiloxane
PDT Pressure decay testing
PE Polyethylene
PES Polyethylsulphone
PFI Private finance initiative
POEM Polyoxyethylene methacrylate
PP Polypropylene
PPCPs Pharmaceuticals and personal care products
PS Polysulphone
PTFE Polytetrafluoroethylene
PV Pervaporation
PVDF Polyvinylidene difluoride
RBC Rotating biological contactor
Redox Reduction−oxidation

rMBR (Biomass) rejection membrane bioreactor
RO Reverse osmosis
SAD Specific aeration demand
SAE Standard aeration efficiency (kgO_2/kWh)
SAF Submerged aerated filter
SBR Sequencing batch reactor
SCADA Supervisory control and data acquisitions
SDI Silt density index
SED Specific energy demand
SEDA Specific energy demand for aeration
SEM Scanning electron micrograph
SGD Specific gas demand
sMBR Sidestream membrane bioreactor (aerobic)
SME small to medium-sized enterprise
SMP Soluble microbial product
SMP_c Soluble microbial product, carbohydrate fraction
SMP_p Soluble microbial product, protein fraction
SNdN Simultaneous nitrification and denitrification
SOTE Standard oxygen transfer efficiency
SRF Specific resistance to filtration
SRT Solids retention time
SUVA Specific UV absorbance per unit organic carbon concentration
SVI Sludge volume index
SW Spiral-wound
TDS Total dissolved solids
TEP Transparent exopolymer particle
TF Trickling filter
TFC Thin film composite
THMFP Trihalomethane formation potential
TIN Total inorganic nitrogen
TIPS Thermal-induced phase separation
TKN Total Kjeldahl nitrogen
TMDL Total maximum daily load
TMP Transmembrane pressure
TN Total nitrogen
TOC Total organic carbon
TON Total organic nitrogen
TSS Total suspended solids
UASB Upflow anaerobic sludge blanket
UF Ultrafiltration
VFD Variable frequency drive (for feed pumps)
VFM VITO fouling measurement
VRM Vacuum rotating membrane
VSS Volatile suspended solids
WRP Water recycling (or reclamation) plant
WwTP/W Wastewater treatment plant/works

Nomenclature

SYMBOLS USED IN THIS BOOK ARE DEFINED BELOW

a Gas–liquid or water–air interface surface area per unit volume (1/m)
A_m Membrane area, m^2
A_t Tube cross-sectional area, m^2
C Dissolved oxygen concentration value, kg/m^3
C* Saturated oxygen concentration value, kg/m^3
c_c Cleaning reagent concentration, mg/L
d Diameter
F Module footprint, m^2
f_{anox} Anoxic tank volume as a proportion of aerobic tank, %
f_{bp} Slowly biodegradable COD fraction
f_{bs} Readily biodegradable COD fraction
f_d Fraction of the biomass that remains as cell debris, gVSS/g substrate
f_{up} Particulate non-biodegradable COD fraction
f_{us} Soluble non-biodegradable COD fraction
F_x Fouling factor
F/M Food to micro-organism ratio
F/M_b Ratio of food to active biomass in the anoxic zone
g Acceleration due to gravity, m/s^2
H Hydrostatic head of pressure, m
H Hydraulic retention time
HRT_{aer} Aerobic hydraulic retention time, h
HRT_{anox} Anoxic hydraulic retention time, h
$HRT_{process}$ Total process HRT, h
i Discount rate, %
iTSS Inert total suspended solids, mg/L
J Flux, LMH
J' Temperature-corrected flux, LMH
J_b Backflush flux, LMH
J_c Critical flux, LMH
J_{net} Net flux, LMH
$J_{net,peak}$ Maximum allowed flux during a limited time period
K Permeability, LMH/bar
K' Temperature-corrected permeability, LMH/bar
k_e Death coefficient, gVSS/(gVSS.d)
$k_{e,n}$ Death coefficient for nitrifying bacteria, gVSS/(gVSS.d)
k_L Volumetric mass transfer coefficient, m/s
K_n Half saturation coefficient for nitrification
K_o Half saturation coefficient for oxygen
K_s Saturation coefficient, g/m^3
L Module length

$\boldsymbol{M}_{\mathbf{b}}$ Oxygen required by biology aeration, kg/d
$\boldsymbol{M}_{\mathbf{m}}$ Oxygen transferred by membrane aeration, kg/d
$\boldsymbol{M}_{\mathbf{o}}$ Total oxygen required, kg/d
$\boldsymbol{M}_{\mathbf{total\ sludge}}$ Total sludge production, kg/d
$\boldsymbol{M}_{\mathbf{x}}$ Mass of component x in the system
$\boldsymbol{M}_{\mathbf{x,aut}}$ Sludge production from nitrification, in gVSS/day
$\boldsymbol{M}_{\mathbf{x,bio}}$ Biomass yield, gVSS/d
$\boldsymbol{M}_{\mathbf{x,het}}$ Heterotrophic biomass produced by a biological system, gVSS/d
$\boldsymbol{M}_{\mathbf{x,TSS}}$ Total sludge yield, g mixed liquor suspended solids, gMLSS/day
$\boldsymbol{n}$ Number of physical cleaning cycles per chemical clean
$\boldsymbol{N}$ Total ammonia, nitrogen or TKN in the influent, mg/L
$\boldsymbol{N}_{\mathbf{e}}$ Total ammonia, nitrogen or TKN in the effluent, mg/L
$\boldsymbol{NO}_{\mathbf{e}}$ Effluent nitrate concentration, mg/L
NO-loading Nitrate load to the anoxic zone, g/d
$\boldsymbol{NO}_{\mathbf{r}}$ Denitrification capacity, g/d
$\boldsymbol{NO}_{\mathbf{x}}$ Concentration of TKN oxidizable to nitrate, mg/L
$\boldsymbol{O}_{\mathbf{A,m}}$ Mass percentage of oxygen in air, %
$\boldsymbol{O}_{\mathbf{out}}$ O_2 in air leaving the surface of the aeration tank, %
$\boldsymbol{P}_{\mathbf{A,1}}$ Blower inlet pressure, Pa
$\boldsymbol{P}_{\mathbf{A,2}}$ Blower outlet pressure, Pa
pCOD Total particulate COD, g/m^3
$\boldsymbol{P}_{\mathbf{d}}$ Pressure at base of aeration tank, Pa
$\boldsymbol{pK_a}$ Acid dissociation constant
$\boldsymbol{P}_{\mathbf{max}}$ Threshold pressure beyond which operation cannot be sustained, bar
$\boldsymbol{PU}_{\mathbf{x}}$ Pollution units
$\boldsymbol{Q}$ Feed flow rate, m^3/day
$\boldsymbol{Q}_{\mathbf{A,b}}$ Net air flow for biological requirements, Nm^3/h
$\boldsymbol{Q}_{\mathbf{A,m}}$ Membrane aeration rate, Nm^3/h
$\boldsymbol{Q}_{\mathbf{int}}$ Recirculation flow rate, m^3/h
$\boldsymbol{Q}_{\mathbf{P}}$ Permeate flow rate, m^3/h
$\boldsymbol{Q}_{\mathbf{peak}}$ Peak influent flow rate, m^3/h
$\boldsymbol{Q}_{\mathbf{pump}}$ Pumping flow rate, m^3/h
$\boldsymbol{Q}_{\mathbf{R}}$ Retentate flow rate, m^3/h
$\boldsymbol{Q}_{\mathbf{w}}$ Sludge wastage rate, m^3/day
$Q_{\mathbf{w'}}$ Sludge waste per unit permeate, m^3/m^3
$\boldsymbol{R}$ Resistance, 1/m
$\boldsymbol{R}_{\mathbf{c}}$ Cake resistance
$\boldsymbol{R}_{\mathbf{col}}$ Resistance offered by colloidal matter, 1/m
$\boldsymbol{r}_{\mathbf{int}}$ Recirculation rate
$\boldsymbol{r}_{\boldsymbol{mr}}$ Membrane recirculation ratio
$\boldsymbol{R}_{\mathbf{sol}}$ Resistance offered by soluble matter, 1/m
$\boldsymbol{R}_{\mathbf{ss}}$ VSS/MLSS ratio in the biomass
$\boldsymbol{R}_{\mathbf{sup}}$ Resistance offered by supernatant materials (colloidal and soluble matter), 1/m
$\boldsymbol{R}_{\mathbf{tot}}$ Total resistance, 1/m
$\boldsymbol{S}$ Substrate (BOD or COD) concentration, mg/L
S/X Substrate to biomass concentration ratio
$\mathbf{SAD_m}$ Specific aeration demand with respect to membrane area, Nm^3/(m^2 h)
$\mathbf{SAD_p}$ Specific aeration demand with respect to permeate volume, Nm^3 air/m^3 permeate

SDNR Specific denitrification rate, gNO_3-N/gVSS
$\boldsymbol{S_e}$ Effluent dissolved substrate, g/m^3
$\mathbf{SEDA_m}$ Specific energy demand for membrane aeration, kWh/Nm^3
$\mathbf{SGD_m}$ Specific gas demand with respect to membrane area, $Nm^3/(m^2\,h)$
$\mathbf{SGD_p}$ Specific gas demand with respect to permeate volume, Nm^3 air/m^3 permeate
$\mathbf{SOTE_{coarse}}$ Standard oxygen transfer efficiency, coarse bubble aeration, %/m
$\mathbf{SOTE_{fine}}$ Standard oxygen transfer efficiency, fine bubble aeration, %/m
$\boldsymbol{T}$ Temperature, °C
$\boldsymbol{t}$ Time
$\boldsymbol{T_a}$ Air temperature, °C
$\boldsymbol{t_c}$ Chemical cleaning interval
$\boldsymbol{t_{crit}}$ Critical filtration time
$\boldsymbol{T_{K,1}}$ Blower inlet temperature, K
$\boldsymbol{t_p}$ Physical cleaning (backflush or relaxation) interval
$\boldsymbol{T_p}$ Physical cleaning (backflush or relaxation) duration
$\boldsymbol{U_G}$ Gas velocity, m/s
$\boldsymbol{U_L}$ Liquid crossflow velocity, m/s
$\boldsymbol{U_R}$ Retentate velocity, m/s
$\boldsymbol{V}$ Volume
$\boldsymbol{V_{aer}}$ Aeration tank volume, m^3
$\boldsymbol{V_{an}}$ Anaerobic tank volume, m^3
$\boldsymbol{V_{anox}}$ Anoxic tank volume, m^3
$\boldsymbol{V_{buffer}}$ Buffer tank volume, m^3
$\boldsymbol{V_m}$ Membrane tank volume, m^3
$\boldsymbol{V_{m,min}}$ Minimum membrane tank volume, m^3
$\boldsymbol{V_{process}}$ Total process volume, m^3
$\boldsymbol{W}$ Power demand, kW
$\boldsymbol{W_{Aeration}}$ Blower power, kW
$\boldsymbol{W_{bw}}$ Pumping power required for backwashing, kWh/d
$\boldsymbol{W_{perm}}$ Pumping power required for permeate pumping, kWh/d
$\boldsymbol{W_{sludge}}$, $\boldsymbol{W'_{sludge}}$ Power required for sludge pumping, kWh/d, kW
$\boldsymbol{W_x}$ Power requirement for permeate pumping and backwashing, kWh/d
$\boldsymbol{X}$ MLSS concentration, mg/L
$\boldsymbol{X_{aer}}$ MLSS concentration in aerobic zone, mg/L
$\boldsymbol{X_{anox}}$ MLSS concentration in anoxic zone, mg/L
$\boldsymbol{X_{b,anox}}$ Active biomass in the anoxic zone, mg/L
$\boldsymbol{X_m}$ Mixed liquor suspended solids level in membrane tank, g/m^3
$\boldsymbol{Y}$ Biomass yield, mass of cells formed per mass of substrate consumed, usually gVSS/BgBOD
$\boldsymbol{y}$ ***or*** $\boldsymbol{y_x}$ Aerator depth
$\boldsymbol{y_{coarse}}$ Coarse bubble aerator depth, m
$\boldsymbol{y_{fine}}$ Fine bubble aerator depth, m
$\boldsymbol{Y_n}$ Nitrification sludge yield, $gVSS/gNH_4$-N
$\boldsymbol{Y_{obs}}$ Observed yield, g.day
$\boldsymbol{\alpha}$ Ratio of mass transfer of oxygen in suspension to that in pure water
$\boldsymbol{\beta}$ Ratio of mass transfer of oxygen in saline water to that in pure water
$\boldsymbol{\gamma}$ Shear rate, 1/s
$\boldsymbol{\delta}$ Separation, m

ψ Specific cake resistance, 1/m^2
Δh, ΔH Total head loss, m
ΔK Permeability change, LMH/bar
ΔP Pressure difference, bar
ΔP_m Transmembrane pressure, bar
α Ratio of mass transfer of oxygen in the sludge to that for pure water
β Ratio of mass transfer of oxygen at the operating dissolved solids concentration to that for pure water
ε Process efficiency, %
η Viscosity, kg/(ms)
θ_x Solids retention time, or sludge age, d
$\theta_{x,aer}$ Aerobic SRT or sludge age, d
$\theta_{x,process}$ Total process SRT
κ Membrane geometry-dependent constant
λ Specific heat capacity of air
μ Growth rate per day, gVSS/(gVSS.d)
μ_m Maximum specific growth rate, gVSS/(gVSS.d)
μ_n Specific growth rate of nitrifying bacteria, g VSS/(gVSS.d)
$\mu_{n,m}$ Maximum specific growth rate of nitrifying bacteria, gVSS/(gVSS.d)
ξ Blower efficiency
ξ_p Permeate pump efficiency
$\xi_{p,sludge}$ Sludge pumping efficiency
ρ Density, kg/m^3
ρ_A Air density, kg/m^3
τ_c Chemical cleaning duration
τ_p Physical cleaning duration (backflush, relaxation)
ϕ Ratio of mass transfer of oxygen at operating temperature to that at standard temperature (20 °C)
$\varphi_{external}$ Membrane packing density: area per external module volume, m^2/m^3
$\varphi_{internal}$ Membrane packing density: area per internal module volume, m^2/m^3
φ_{tank} Membrane packing density: area per membrane tank, m^2/m^3
ω Correction factor exponent for oxygen mass transfer equation
ω_{coarse} ω-Factor for coarse bubble aeration
ω_{fine} ω-Factor for fine bubble aeration

Greek alphabet: aide memoire

Α	α	alpha	Ν	ν	nu
Β	β	beta	Ξ	ξ	ksi
Γ	γ	gamma	Ο	ο	omicron
Δ	δ	delta	Π	π	pi
Ε	ε	epsilon	Ρ	ρ	rho
Ζ	ζ	zeta	Σ	σς	sigma
Η	η	eta	Τ	τ	tau
Θ	θ	theta	Υ	υ	upsilon
Ι	ι	iota	Φ	φ	phi
Κ	κ	kappa	Χ	χ	chi
Λ	λ	lambda	Ψ	ψ	psi
Μ	μ	mu	Ω	ω	omega

Glossary of Terms

A number of key terms used in the book are defined below. Proprietary names and processes are not included.

Aerobic Conditions where oxygen acts as electron donor for biochemical reactions
Air-lift The use of air to lift liquid up a channel
Allochthonous Of terrestrial origin
Anaerobic Conditions where biochemical reactions take place in the absence of oxygen
Anisotropic Having symmetry only in one plane
Annular flow Flow through an annulus (or gap created by concentric cylinders)
Anoxic Conditions where an oxyanion, rather than oxygen, acts as the electron donor for biochemical reactions
Anthropogenic Of human origin or derived from human activity
Autochthonous Of microbial origin
Autotrophic Using carbon dioxide as sole carbon source for growth and development
Backflushing Reversing flow through a membrane to remove foulants (also called *backwashing*)
Biofilm Film or layer containing biological material
Biological treatment Process whereby dissolved organic chemical constituents are removed through biodegradation (also called *biotreatment*)
Biomass Viable (living) micro-organisms used to achieve removal of organics through biotreatment
Blocking/blinding Occlusion of the membrane pores at the surface by depositing solids
Bubble flow Air/liquid two-phase flow where the liquid is the continuum
Cake Solid material formed on the membrane during operation
Cassette See Appendix C
Churn flow Air/liquid two-phase flow at high air/liquid ratio
Clogging Accumulation of solids within the membrane channels
Concentration polarization Tendency of solute to accumulate at membrane:solution interface during crossflow operation
Conditioning First stage of membrane fouling through adsorption of material fouling
Critical flux Flux below which permeability decline is considered negligible
Critical suction Threshold pressure arising during sub-critical flux fouling pressure
Crossflow Retentate flow parallel to the membrane surface
Cyclic aeration Aeration on an 'n s on/n s off' basis, where n is normally between 5 and 30 s
Dalton (Da) Molecular mass relative to that of a hydrogen atom
Dead-end or full-flow Flow where all of the feed is converted to permeate
Death coefficient A biokinetic parameter defining the rate at which micro-organisms become inactive
Denitrification Biochemical reduction of nitrate to nitrogen gas
Dense membrane Membrane of high selectivity attained by specific physicochemical interactions between solute and membrane

Diffusive MBR MBR configured so that the membrane acts to pass gas into the bioreactor in molecular (bubble-less) form

Electrodialysis Membrane separation process by which ions are removed via ion exchange membranes under the influence of an electromotive force (voltage)

Electron donor Species capable of donating an electron to a suitable acceptor, thus providing oxidation

Element See Appendix C

Endogenous Developing or originating within, or part of, a micro-organism or cell

Exogenous Originating outside the micro-organism or cell

Extractive MBR MBR configured so that priority pollutants are selectively extracted into or out of the bioreactor via the membrane

Facultative Conditions where oxyanions, such as nitrate, act as electron donor for biochemical reactions

Filament Single hollow fibre or capillary tube

Filamentous index Parameter indicating relative presence of filamentous bacteria in sludge

Fixed film process Process configured with the biofilm attached to a solid medium (which may be a membrane)

Floc Aggregated solid (biomass) particle

Flux Quantity of material passing through a unit area of membrane per unit time

Flux-step Critical flux identification method whereby flux is incrementally increased and the TMP or permeability response recorded

***F*/*M* ratio** Rate at which substrate is fed to the biomass compared to the mass of biomass solids

Forward osmosis Extractive membrane separation process by which water is extracted into a draw solution under an osmotic pressure difference

Fouling Processes leading to deterioration of flux due to surface or internal blockage of the membrane

Gas/air-lift Lifting of liquid using gas/air

Gas/air sparging Introduction of gas/air bubbles

Gel layer Precipitation of sparingly soluble macromolecular species at membrane surface

Heterotrophic Requiring an organic substrate to provide carbon for growth and development

Humic matter Organic matter of terrestrial origin

Hydraulic loading rate Rate at which water enters the reactor

Hydrogenotrophic Feeding on hydrogen

Hydrophilicity Water-absorbent, or extent of 'wetting by water'

Hydrophobic Water repellent

Immersed (membrane) (Membrane) placed inside the bioreactor

Inoculum Medium containing micro-organisms initially introduced into a reactor to establish new populations and start the biotreatment process

Interfacial region Region at the membrane:solution interface

Interstitial Inter-membrane space

Irrecoverable fouling Fouling which is not removed by physical or chemical cleaning

Irreversible fouling Fouling which is removed by chemical cleaning, also referred to as *permanent fouling*

Isoporosity Property reflecting narrowness of pore size distribution

Lamella plate Angled plate in a sedimentation tank designed to enhance settlement

Lumen-side Inside the fibre/filament/lumen

Macropore Pore with diameter above 50 μm
Maintenance cleaning Cleaning with less aggressive chemicals to maintain membrane permeability
Membrane distillation Membrane separation process by which water vapour is extracted through the membrane under a pressure difference
Mesophilic Thriving at intermediate temperatures (20–45 °C)
Mesopore Pore with diameter between 2 and 50 μm
Methanogens Micro-organisms producing methane as a metabolic by product
Microfiltration Membrane separation process by which particles are rejected by the membrane and water and dissolved matter is passed through it
Micropore Pore with diameter below 2 μm
Mist flow Air/liquid two-phase flow where the air is the continuum
Mixed liquor The biomass-containing slurry formed in the bioreactor during biological processing, also referred to as *sludge*
Modularization Based on modules: using more modules at higher flows, rather than increasing the unit process size
Module See Appendix C
Monod kinetics Kinetics defining biomass growth and decay during biotreatment
Multitube A multiple of tubular membranes in a module
Nanofiltration Pressure-driven membrane separation process by which divalent ions and medium to high molecular weight organic matter is rejected by the membrane
Nitrification Biochemical oxidation of ammonia to nitrate
(Organic) loading rate Rate at which (organic) matter is introduced into the reactor
Packing density Membrane area per unit module volume
Panel See Appendix C
Percolation theory Theory defining probability of water flowing through a medium containing a three-dimensional network of interconnected pores
Permeability Ease of flow through membrane, represented by flux:pressure ratio
Permeate Water or fluid which has passed through the membrane
Perm-selectivity Permeation of some components in preference to others
Pleated filter Type of flat sheet module cartridge
Plug flow Flow in which no back-mixing or dispersion occurs along the length of the pipe or reactor
Pore plugging The complete blocking of pores by suspended/colloidal matter – also called pore occlusion
Porous membrane Membrane of low selectivity operating by physical straining alone
Primary treatment The first stage of conventional sewage treatment, normally considered to be sedimentation
Psychrophilic Thriving at relatively low temperatures (0–20 °C)
Rack See Appendix C
Recovery/conversion Fraction of feedwater converted to permeate product
Recovery clean Cleaning with aggressive chemicals to recover membrane permeability
Redox Conditions defined by the presence of either dissolved oxygen or some other species capable of providing oxygen for biochemical conversion
Relaxation Ceasing permeation whilst continuing to scour the membrane with air bubbles
Resistance Resistance to flow, proportional to flux:pressure ratio
Retentate Water or fluid which is rejected by the membrane

Reverse osmosis Pressure-driven membrane separation process by which most charged dissolved materials are rejected by the membrane

Reversible fouling Gross solids attached to the membrane surface and which can be removed by physical cleaning (backflushing and/or relaxation), also called temporary fouling

Secondary treatment The biochemical treatment stage of conventional sewage treatment

Septum Coarse membrane filter

Shear (stress) Force applied to a body which tends to produce a change in its shape, but not its volume

Shear-induced diffusion Diffusion of matter away from the membrane under the influence of the shear imparted just beyond the hydrodynamic boundary layer

Shell-side Outside the membrane fibre/filament/lumen

Sidestream (membrane) Stream (containing the membrane) outside the bioreactor

Sludge See *mixed liquor*

Slug flow Air/liquid two-phase flow at moderate air/liquid ratios

Stack See Appendix C

Struvite Magnesium ammonium phosphate salt

Substrate Surface or medium on which an organism grows or is attached

Supernatant Liquid clarified by sedimentation

Surface porosity Percentage of the surface area occupied by the pores

Sustainable flux Flux for which the TMP increases gradually at an acceptable rate, such that chemical cleaning is not necessary

Tertiary treatment Final 'polishing' stage of conventional sewage treatment, normally considered to be supplementary clarification and/or disinfection

Thermophilic Thriving at relatively high temperatures (49−57 °C)

TMP jump Sudden TMP increase when operating under sub-critical flux conditions

TMP-step Critical flux identification method where TMP is incrementally increased and the flux or permeability response recorded

Tortuosity Ratio of pore length to membrane thickness

Ultrafiltration Pressure-driven membrane separation process by which particles, colloids and macromolecules are rejected by the membrane and water and dissolved matter passed through it

Upflow clarification Dynamic clarification by sedimentation

Zeta potential Potential (in mV) at the shear plane of a solid:liquid interface

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