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Membrane Bioreactor for Wastewater Treatment

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Preface

There is a trend that stricter and stricter wastewater discharge standards are going to be implemented. Therefore, available techniques for municipal wastewater treatment need to be upgraded in order to meet the requirements of those standards. Meanwhile, new techniques are also under developing. While conventional secondary settlers are replaced by membranes for sludge and waster separation, the combination of reactor and membrane is called membrane bioreactor (MBR). The main benefits of the introduction of MBR are that much smaller footprints are required since secondary settlers are not required; and much better effluent qualities can be achieved due to the application of membranes. The main drawback of the application of MBR is that the cost of construction and operation of MBR is higher than conventional bioreactors. Nevertheless, MBR has been applied from 1990s and now is more and more popular.

This edited book tries to include the most related and practical knowledge about MBR. Detailed discussion about deep scientific knowledge in all fields of MBR is not addressed. The book starts from the introduction of biological wastewater treatment. In addition, the book focuses on knowledge of membrane, performance of MBR and its operation. This book is suitable for person who does not have any background in MBR. After reading this book, readers should be acquainted with MBR.

Please be noted that, up to now, most knowledge is about operation of MBRs for municipal wastewater treatment in which aerobic MBRs are applied. Regarding the application of anaerobic MBR, the operation experience is relatively limited. However, recent development of anaerobic MBR is also included in this book.

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1 Overview of biological wastewater treatment

In 2007, the development of sanitation was voted to be the greatest medical advance in the last 166 years in a contest run by the British Medical Journal. This confirms the utterly important role of proper sanitation in achieving and maintaining good public health. In many industrialized countries, wastewater is transported safely away from the house-holds. Proper sewage treatment is however not always in place, in particular in many developing countries where sanitation coverage is, by far, less in comparison with water supply. The need for proper sanitation was made explicit in the United Nations Millennium development goals. Goal number 7 urges for the reduction by half of the population living without proper sanitation. Despite significant efforts, progress on sanitation targets is very slow and still lacking behind. Acknowledging the impact of sanitation on public health, poverty reduction, economic and social development and the environment, the General Assembly of the United Nations declared 2008 to be the International Year of Sanitation. The goal was to focus the world's attention on the need to start implementing proper sanitation solutions for all (Mogens et al. 2008).

Important in this is to not only connect people to sanitation solutions, but to make this connection last in an environmentally sustainable way. Sewer systems and wastewater treatment plants have proven to be very efficient in conveying and removing pathogens, organic pollutants and nutrients. However, they require proper operation and maintenance, and a good understanding of the processes involved (Mogens et al. 2008).

The aim of this chapter is to introduce basic knowledge about wastewater treatment.

1.1 Wastewater constituents

The constituents of wastewaters, which origin from different sources of wastewaters, vary significantly. Traditionally, wastewater treatment mainly includes the removal of suspended matters and organic matters. However, the removal of nutrient (ammonia, nitrite and nitrate, phosphate) is necessary in many countries.

1.2 Organic material removal

1.2.1 Principle of biodegradable organic material removal

A wide variety of heterotrophic bacteria can consume organic materials, in which a lot of bioprocesses occur. During the bioprocess, energy is released and used to sustain activities of the bacteria. In addition, biopolymers are synthesized, which connects bacteria together and forms biological flocs, known as suspended sludge, or biofilms. The flocs can be separated from liquid by gravity settling, which promotes the retention of sludge. The retention of sludge is essential for biological wastewater treatment. Meanwhile, bacteria also undergo an endogenous respiration process, which reduces the concentration of suspended sludge.

The above processes can be simplified and represented by the following equations (Tchobanoglous et al. 2004). In the equations, $C_5H_7NO_2$ represents bacteria.

Oxidation and synthesis

$$COHNS + O_2 + nutrients \rightarrow CO_2 + NH_3 + C_5H_7NO_2 + other\,end\;products \tag{1-1}$$

Endogenous respiration

$$C_5H_7NO_2 + 5O_2 \rightarrow 5CO_2 + 2H_2O + NH_3 + energy \tag{1-2}$$

Using acetate as an example, the two equations can be combined together:

$$0.125CH_3COO^- + 0.0295NH_4 + 0.103O_2 \rightarrow 0.0295C_5H_7NO_2 + 0.0955H_2O + 0.0955HCO_3^- + 0.007CO_2 \tag{1-3}$$

1.2.2 Classification of organic matters

Apparently, not all organic matters are biodegradable. In order to design a wastewater treatment plant, it is suggested to carefully characterize the constituents of organic matters. The organic matters can be classified into four categories: soluble and non-biodegradable (S_I), soluble and biodegradable (Ss), non-soluble and non-biodegradable (X_I), non-soluble and biodegradable (X_S). In order to know the characteristics of a kind of wastewater, the four kinds of constituents should be determined by measurements. The following determination example is based on experiences on Dutch municipal wastewater (Roeleveld and van Loosdrecht 2002).

The consequence for the characterisation of wastewater is summarized below:

1. Determination of S_I based on the inert soluble chemical oxygen demand (COD) in the effluent of the wastewater treatment plant;
2. Determination of S_s by subtracting the fraction S_I from the soluble COD in the effluent;
3. Determination of X_s by subtracting the fraction S_s from the biodegradable COD;
4. Determination of X_I with the equation $X=COD_{influent}-S_I-S_s-X_{s.}$

1.2.3 Calculation for organic matter removal

Based on the characterization of organic materials in wastewater, following calculation for organic matter removal is based on these preconditions:

1. The amount of volatile suspended solids (VSS_s) in the plant for organic matter removal
   = the amount of microorganisms + respiration residual products + non-biodegradable organic particles

$$= \frac{(X_S + S_S)\cdot Q \cdot f'_{s'up} \cdot SRT}{f_{cv}} + f_H \cdot b_H \cdot SRT \cdot \frac{(X_S + S_S)\cdot Q \cdot f'_{s'up} \cdot SRT}{f_{cv}} + \frac{X_I \cdot Q \cdot f'_{s'up} \cdot SRT}{f_{cv}}$$

2. The amount of inert suspended solid in the plant for organic matter removal
   = contribution from influent + inorganic part in sludge

   $$= Q \times SRT + \frac{(X_S + S_S)\cdot Q \cdot f'_{s'up} \cdot SRT}{f_{cv}} \cdot f_{i,OHO}$$

3. The effluent COD $=S_I$ + VSS ×coefficient
   The VSS at here can be obtained based on a relationship between VSS and turbidity of the effluent.

4. Required oxygen supply= COD removal + sludge respiration

   $$= (X_S + S_S)\cdot Q \cdot (1 - f_{cv} \cdot Y_{HV}) + \frac{(X_S + S_S)\cdot Q \cdot f'_{s'up} \cdot SRT}{f_{cv}} \cdot b_H \cdot (1 - f_H)$$

5. COD balance
   Influent COD= COD input by oxygen supply + effluent COD+ COD of discharged sludge.

Where

Q : flow rate
$f'_{s'up}$: particulate un-biodegradable fraction of total influent COD
f_{cv} : COD to VSS ratio of the sludge
f_h : un-biodegradable fraction of the ordinary heterotrophic organisms (OHOs)
b_H : specific rate of endogenous mass loss of OHOs
$f_{i,OHO}$: inorganic contents of OHO
Y_{HV} : yield of OHO in terms of VSS

1.3 Nitrogen removal

1.3.1 Nitrification

The ammonium ions are produced in wastewater from the hydrolysis of urea and degradation of organic-nitrogen compounds. Hydrolysis and degradation of organic-nitrogen compounds results in the release of amino groups and the production of ammonium ions (Gerardi 2002). Biological nitrification is the conversion or oxidation of ammonium ions to nitrite ions and then to nitrate ions. During the oxidation of ammonium ions and nitrite ions, oxygen is added to the ions by a unique group of organisms. Nitrification occurs in nature and in activated sludge processes. Nitrification in wastewater treatment is required because of (Tchobanoglous et al. 2004):

1. The effect of ammonia on receiving water on dissolved oxygen (DO) concentration and fish toxicity;
2. The need to provide nitrogen removal to control eutrophication;
3. The need to provide nitrogen control for water-reuse applications.

Although ammonium ions and ammonia are reduced forms of nitrogen, it is the ammonium ion, not ammonia, which is oxidized during nitrification. The quantities of ammonium ions and ammonia in an aeration tank are dependent on the pH and temperature of the activated sludge. In the temperature range of 10 to 20 °C and pH range of 7 to 8.5, which are typical of most activated sludge processes, about 95% of the reduced form of nitrogen is present as ammonium ions (Gerardi 2002).

The oxidation of ammonium ions and nitrite ions is achieved through the addition of dissolved oxygen within bacterial cells. Because nitrification or the biochemical reactions of oxygen addition occur inside biological cells, nitrification occurs through biochemical reactions (Gerardi 2002). Nitrification includes converting ammonia into nitrite which is further oxidized into nitrate. The processes are depicted by the two following reactions:

$$2NH_4^+ + 3O_2 \longrightarrow 2NO_2^- + 4H^+ + 2H_2O \tag{1-4}$$

$$2NO_2^- + O_2 \longrightarrow 2NO_3^- \tag{1-5}$$

The two reactions make

$$NH_4^+ + 2O_2 \longrightarrow NO_3^- + 2H^+ + H_2O \tag{1-6}$$

Considering biomass synthesis, the biochemical reaction is

$$NH_4^+ + 1.863O_2 + 0.098CO_2 \rightarrow 0.019C_5H_7NO_2 + 0.98NO_3^- + 0.094H_2O + 1.98H^+ \tag{1-7}$$

The above reaction shows that, while each gram of ammonia nitrogen (as N) is consumed, 4.25 g of O_2 are depleted and 7.07 g of alkalinity as $CaCO_3$ are removed.

Factors influencing nitrification

In order to achieve a good nitrification effect, the following factors should be taken into account while operating a wastewater treatment plant.

1. Maximum ammonia oxidation bacteria growth rate (μ_{max})
   It is between 0.3–0.75. The value should be determined by experiments, or else a low value should be applied. The determination procedure can be referred to published literature (Mogens et al. 2008);

2. DO
   Should be higher than 1 mg/L;
3. Temperature
   Should be between 20–30 $^{\circ}$C;
4. un-aerated zone
   Maximum 0.6 of the overall plant;
5. Shortcut flow
   Should be avoided;
6. pH and alkalinity
   Optimum values are between 7.5 and 8.0. Usually control it between 7.0 and 7.2;
7. SRT
   Should be twice than the minimum doubling time;
8. Organic matter
   Biodegradable oxygen demand (BOD) should be between 15–20 mg/L.

1.3.2 Denitrification

Wastewater denitrification describes the use of nitrite ions or nitrate ions by facultative anaerobes (denitrifying bacteria) to degrade BOD. Although denitrification often is combined with aerobic nitrification to remove various forms of nitrogenous compounds from wastewater, denitrification occurs when an anoxic condition exists (Gerardi 2002).

Facultative anaerobes make up approximately 80% of the bacteria within an activated sludge process. These organisms have an enzymatic ability to use free molecular oxygen, nitrite ions, or nitrate ions to degrade BOD. Facultative anaerobes prefer and use free molecular oxygen when it is available. The use of free molecular oxygen provides the bacteria with more energy for cellular activity, growth, and reproduction than does the use of nitrite ions or nitrate ions (Gerardi 2002).

Bacterial degradation of BOD is 'respiration'. Respiration may be aerobic (oxic) or anaerobic. Aerobic respiration occurs when free molecular oxygen is available and is used to degrade BOD (Gerardi 2002). Anaerobic respiration occurs when free molecular oxygen is not available and another molecule is used to degrade BOD. Molecules other than free molecular oxygen that can be used to degrade BOD include nitrite ions and nitrate ions. The molecule used for the degradation of BOD is dependent on its availability, the presence of other molecules, and the enzymatic ability of the bacterial population. If nitrite ions or nitrate ions are used to degrade BOD, such as a five carbon sugar, this form of respiration is termed 'anoxic' (Gerardi 2002).

During anoxic respiration, nitrite ions and nitrate ions are reduced (oxygen removed from the ions) through several biochemical steps or reactions. The principle gaseous end product of the biochemical reactions is molecular nitrogen (Gerardi 2002).

Anoxic respiration or denitrification is termed 'dissimilatory' nitrite or nitrate reduction, because nitrite ions and nitrate ions, respectively, are reduced to from molecular nitrogen. The nitrogen in the nitrite ions or nitrate ions is not incorporated into cellular material, the nitrogen in the ions is loss to the atmosphere as a gas (Gerardi 2002).

There are three kinds of organic materials can be used for denitrification, i.e. organic materials from wastewater; products from endogenous decay; external added organic materials. Different organic donors would produce different denitrification rates. The following equations show how denitrification occur while different organic materials are used as electro donors (Tchobanoglous et al. 2004).

Wastewater

$$C_{10}H_{19}O_3N + 10NO_3^- \rightarrow 5N_2 + 10CO_2 + 3H_2O + NH_3 + 10OH^- \tag{1-8}$$

Methanol

$$5CH_3OH + 6NO_3^- \rightarrow 3N_2 + 5CO_2 + 7H_2O + 6OH^- \tag{1-9}$$

Acetate

$$5CH_3COOH + 8NO_3^- \rightarrow 4N_2 + 10CO_2 + 6H_2O + 8OH^- \tag{1-10}$$

Factors influencing denitrification

In order to achieve a good denitrification effect, the following factors should be taken into account while operating a wastewater treatment plant.

1. Carbon source
   If BOD_5 /TN>3-carbon source is considered sufficient and external carbon source addition is required.
2. pH
   It is better to control pH between 6 and 7.
3. DO
   In general, the proteins required for denitrification are only produced under (close to) anaerobic conditions, and if anaerobically grown cells are exposed to O_2 then the activities of the proteins are inhibited.
4. Temperature
   Should be between 20–40 °C.

1.4 Phosphate removal

1.4.1 Principle for phosphate removal

Some bacteria can survive while anaerobic and aerobic conditions both exist. These bacteria are able to accumulate a large amount of phosphate in their inner bodies. The bacteria are therefore called phosphate accumulating organisms (PAOs). By the discharge of the bacteria from reactors, biological phosphate removal can be achieved.

Unlike most other microorganisms, PAOs prefer take up carbon sources such as volatile fatty acids (VFAs) under anaerobic conditions, and store them intracellularly as carbon polymers, namely poly-b-hydroxyalkanoates (PHAs). The energy for this biotransformation is mainly generated by the cleavage of polyphosphate and the release of phosphate. Reducing power is also required for PHA formation, which is produced largely through the glycolysis of internally stored glycogen. Aerobically, PAOs can use their stored PHA as energy source for biomass growth, glycogen replenishment, P uptake and polyphosphate storage. Net P removal from the wastewater is achieved through the removal of waste activated sludge containing a high polyphosphate content (Mino et al. 1998). The schematic description of biological phosphate removal is shown in figure 1.1.

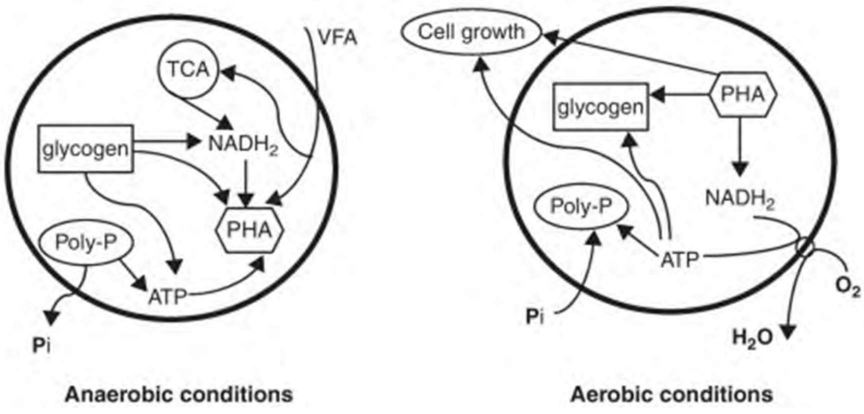

Figure 1.1 Metabolism of PAO under anaerobic and oxic conditions (van Haandel and van der Lubbe 2007)

1.4.2 Factors influencing phosphate removal

When operated successfully, the enhanced biological phosphate removal (EBPR) process is a relatively inexpensive and environmentally sustainable option for P removal; however, the stability and reliability of EBPR can be a problem. It is widely known that EBPR plants may experience process upsets, deterioration in performance and even failures, causing violations to discharge regulations. In some cases, external disturbances include:

1. high rainfall;
2. excessive nitrate loading to the anaerobic reactor;
3. nutrient limitation;
4. Competition between PAO and GAO.

Microbial competition between PAOs and another group of organisms, known as the glycogen (non-polyphosphate) accumulating organisms (GAOs), has been hypothesized to be the cause of the degradation in P removal. Like PAOs, GAOs are able to proliferate under alternating anaerobic and aerobic conditions without performing anaerobic P release or aerobic P uptake, thus they do not contribute to P removal from EBPR systems.

Complex substrates do not favour the growth of PAOs. However, VFA promote proliferation of PAOs. In addition, Low COD/P (i.e. 10–20 mg COD/P) is preferred.

It is reported that only the readily biodegradable fraction of the influent COD was used by the PAOs (vanLoosdrecht et al. 1997). Recent studies have suggested that propionate may be a more favourable substrate than acetate for successful EBPR performance. These studies suggest that a propionate feed source may provide an advantage to PAOs over GAOs.

Numerous studies have found that a high COD/P ratio (e.g. 450 mg COD/ mg P) in the wastewater feed tends to favour the growth of GAOs instead of PAOs. Thus, a low COD/P ratio (e.g. 10–20 mg COD/mg P) should be more favourable to the growth of PAOs.

Some environmental conditions also can have an impact on phosphate removal. These environmental conditions include: pH, temperature and Do.

1. pH
   High pH favours the growth of PAOs and negatively influences GAOs ability to take up VFA. 7.25 is suggested as critical pH. Many studies have shown that a higher ambient pH in enriched PAO sludges has resulted in a higher anaerobic P release. Aerobically, a series of batch tests has shown that P uptake, PHA utilization and biomass growth were all inhibited by a low pH (6.5), and suggested that a higher aerobic pH (7–7.5) would be more beneficial for PAOs. This suggests that a higher pH not only results in a higher energy demand for acetate uptake, but also negatively affects the ability of GAOs to take up acetate.

2. Temperature
   An increase in the fraction of GAOs and decrease in the fraction of PAOs was concluded with an increase in temperature from 20 to 35 ^{o}C, which correlated well with a decrease in the P content in the sludge. The experimental evidence obtained thus far suggests that GAOs tend to become stronger competitors with PAOs at higher temperatures. This implies that competition by GAOs with PAOs in EBPR plants may be more problematic in warm climates, and during the summer months. Successful EBPR operation has been observed at very low temperatures, even 5 ^{0}C, though a higher sludge age was necessary at low temperatures due to the decrease in the kinetics of the process at low temperatures. This improved performance has been hypothesized to be due to a shift in the microbial community from GAOs to PAOs.

3. DO
   Poor P removal performance was more frequently observed at very high DO concentrations of 4.5–5.0 mg/L, while DO concentrations of approximately 2.5–3.0 mg/L seemed to correlate with a greater abundance of PAO. It has been observed that aerobic and anoxic P uptake is inhibited by the presence of nitrite. Thus, it seems that the presence and accumulation of nitrite inhibits PAOs, thereby favouring the growth of GAOs.

2 Introduction of membrane bioreactor

The aim of this chapter is to give a brief introduction about membrane bioreactor, which will be helpful for readers to go further to other chapters.

2.1 Introduction of membrane

2.1.1 Membrane types

Membranes are usually made of polymer or inorganic materials. There are lots of small pores in membranes. Figure 2.1 shows the surface of a membrane. The pores are so small that they are only can be seen with the help of microscopes. Due to their small sizes, the membrane pores only allow small particles and water to penetrate from one side of a membrane to the other side. As a result, water can penetrate the membranes while most particles are retained. Hence, clean water can be produced through the application of membranes. The clean water is then called permeate. Please be noted that there are also many tiny particles in the permeate of membranes. However, the amount of residual particles and their sizes depend on the type of applied membranes. In addition, permeate of membrane is characterized by much better water qualities than the effluent of conventional secondary clarifiers.

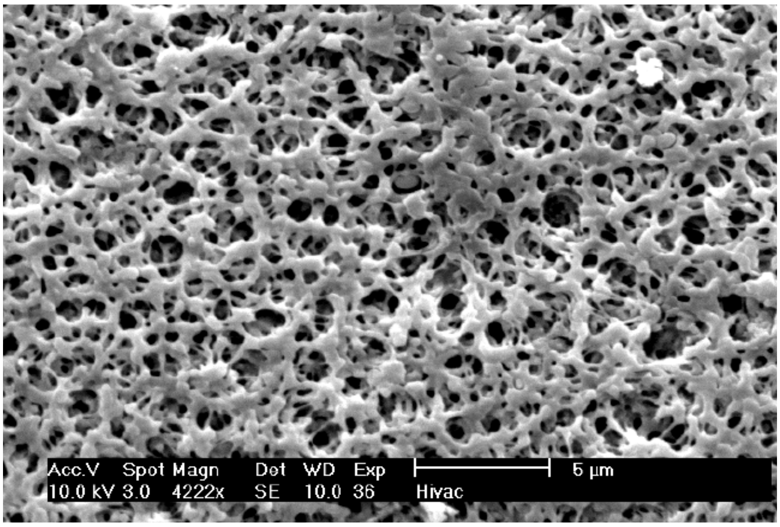

Figure 2.1 microscopy of morphology of membrane (Churchouse 2005).

Filtration class	Particle capture size (µm)	Typical contaminants removed	Typical operating pressure ranges
Microfiltration (MF)	0.1–10	Suspended solids, bacteria, protozoa	0.1–2 bar
Ultrafiltration (UF)	0.003–0.1	Colloids, proteins, polysaccharides, most bacteria, viruses (partially)	1–5 bar (cross flow) 0.2–0.3 bar (dead-end and submerged)
Nanofiltration (NF)	0.001	Viruses, natural organic matter, multivalent ions (including hardness in water)	5–20 bar
Reverse osmosis (RO)	0.0001	Almost all impurities, including monovalent ions	10–100 bar

Table 2.1 classification of membrane types (European-Commission 2010)

Based on the sizes of the pores, the membranes are called microfiltration (MF) membrane, ultrafiltration (UF) membrane, nanofiltration membrane or reverse osmosis (RO) membrane, respectively. The classification of membranes and their functions are shown in table 2.1. Usually, microfiltration and ultrafiltration membranes are applied together with bioreactors in wastewater treatment, which is termed membrane bioreactors (MBRs).

2.1.2 Configurations of membranes

There are three principal membrane configurations that are currently applied in MBR e, i.e. hollow fiber, flat sheet and tubular membrane. Hollow fiber and flat sheet membranes are usually submerged in liquid while tubular membranes are usually placed outside of bioreactors.

Figure 2.2 hollow fiber membranes

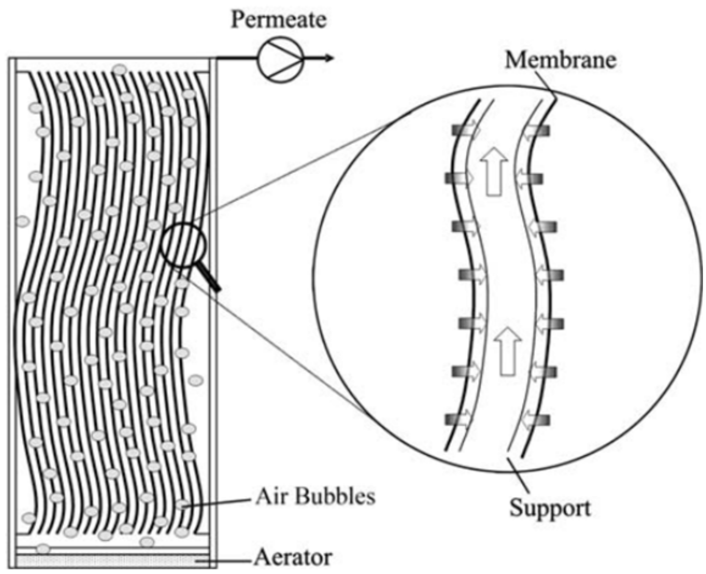

Figure 2.3 schematic view of a hollow fiber membrane module (Cornel and Krause 2008)

In the hollow fiber module, large amounts of hollow fiber membranes make a bundle (figure. 2.2). In hollow-fiber modules, the flow is usually from outside to inside (figure 2.3). The diameters of the hollow fibers generally are a few hundred micrometres. The modules are installed either horizontally or vertically. These modules feature a high packing density and are submerged in biomass (Cornel and Krause 2008).

The rate of water passing through membrane surface is called membrane flux. Higher flux means that more water can be produced by a unit membrane surface. However, during operation, under certain operation conditions, the membrane flux inclines to decrease, which is called membrane fouling. The cause of membrane fouling is that particles accumulate on membrane surface, which increases membrane filtration resistance. Membrane fouling is a major problem in membrane operation because it increases operation cost and requires membrane cleaning. Hence, certain approaches should be employed to prevent and remove the fouling in order to recover the membrane flux, which is called fouling control. Detailed contents about membrane fouling and its control are introduced in chapter 3.

Fouling control of hollow fiber membranes is usually achieved by aerators installed underneath the membrane module. Often coarse bubble aeration acts as a source of scour at membrane surface, but fine bubble aerators are also employed. Membrane operation may include periodic relaxation (pressure release) and/or back-flushing for removing the fouling layer from the membrane surface. In hollow-fiber modules, in particular, sludging and braiding of the membranes can be a major problem. Braiding is caused by hair and/or long fibers (e.g., additive cellulose fibers) that may loop around the membrane bundle, typically in the upper part of the module. Braiding is to be avoided by carefully screening the influent or the recirculated mixed liquor. In the lower part of the module, sludge may accumulate due to insufficient water and/or airflow. Especially when reducing the air scour to save energy, it should be kept in mind that a sufficient flow is crucial to preventing sludging of hollow-fiber bundles. Both effects are likely to reduce the active membrane surface area and mechanical cleaning is necessary (Cornel and Krause 2008).

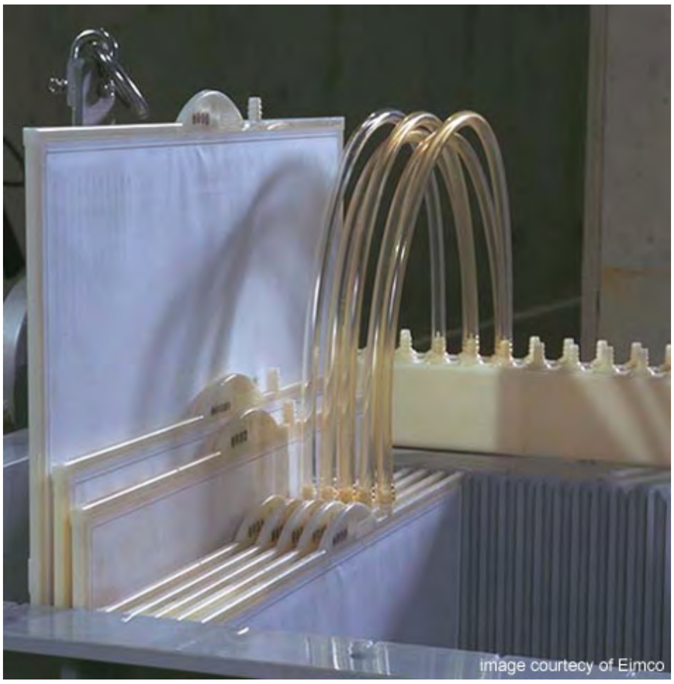

Figure 2.4 flat sheet membranes

Flat sheet membrane modules comprise of flat sheet membranes with separators. The pieces of these sheets are clamped onto a plate. The water flows across the membrane and permeate is collected through pipes emerging from the interior of the membrane module which is operated under vacuum (figure 2.4) (Radjenović et al. 2008). A schematic view of the operation of flat sheet membranes is shown in figure 2.5.

Plate-and-frame membrane modules are less prone to braiding. Long fibers and hair cannot loop around the membrane. In plate-and-frame modules, fouling may occur in the peripheral area. Also, blocking of the gaps occurs. At present, plate-and-frame modules are operated at high airflow rates in order to minimize these phenomena. Most of the commercially available plate-and-frame modules cannot be back-flushed, which restricts the cleaning (Cornel and Krause 2008).

Plate and fibre membranes are usually applied in municipal wastewater treatment, while the use of tubular membrane is relatively limited. Table 2.2 shows the advantage and disadvantages of plate and fibre membranes (van Haandel and van der Lubbe 2007).

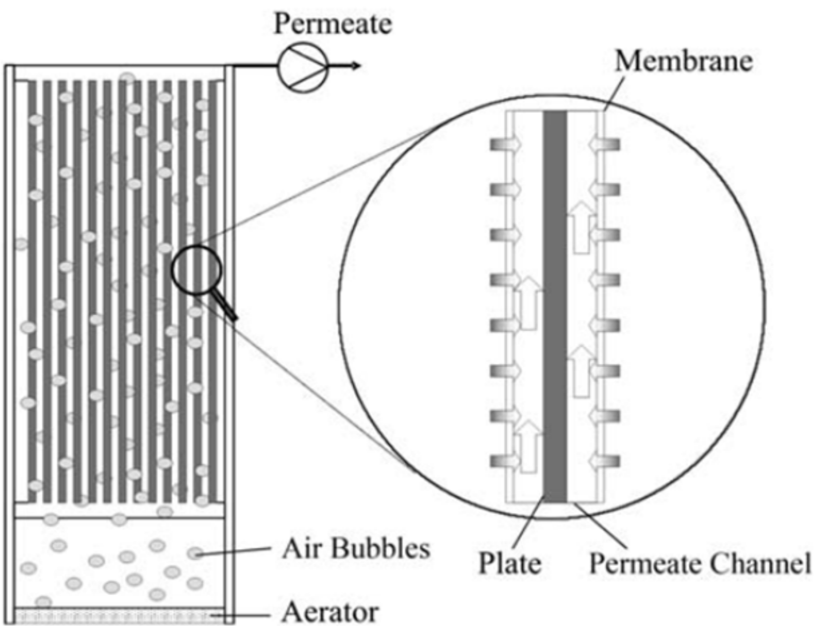

Figure 2.5 schematic view of a vertically arranged submerged plate membrane module (Cornel and Krause 2008).

Type	Advantages	disadvantages
Plate	✓ Robust ✓ Less susceptible to clogging compared to the fibre membrane with top header or two header configuration ✓ Simple system & process control configuration ✓ Manual cleaning possible ✓ Low frequency of leaning	✓ Less specific surface area per m3 module ✓ Back flushing not possible ✓ Higher aerating requirement ✓ More susceptible to channelling: the air speed between the two plates is high but at the plate surface itself it is low. This leads to solids build-up on the membrane surface ✓ Automated cleaning is expensive
fibre	✓ Back flushing possible ✓ High specific surface area ✓ Lower aeration requirement ✓ Completely automated cleaning possible	✓ Susceptible to clogging, depending on module configuration ✓ Manual cleaning non-practical ✓ More complex system

Table 2.2 advantages and disadvantages of plate and fibre membranes

There are also membrane configurations such as tubular module, which is not widely used as the plate and fibre membranes. Typically, tubular membranes are encased in pressure vessels, and mixed liquor is pumped to tubular membranes. Tubular membranes are predominantly used for side-stream configurations. Tubular modules are mainly operated in the 'inside-out' mode, whereas hollow fiber and flat sheet modules are mostly immersed directly in mixed liquor with permeate drawn through the membranes using vacuum pumps (Radjenović et al. 2008).

Figure 2.6 tubular membrane

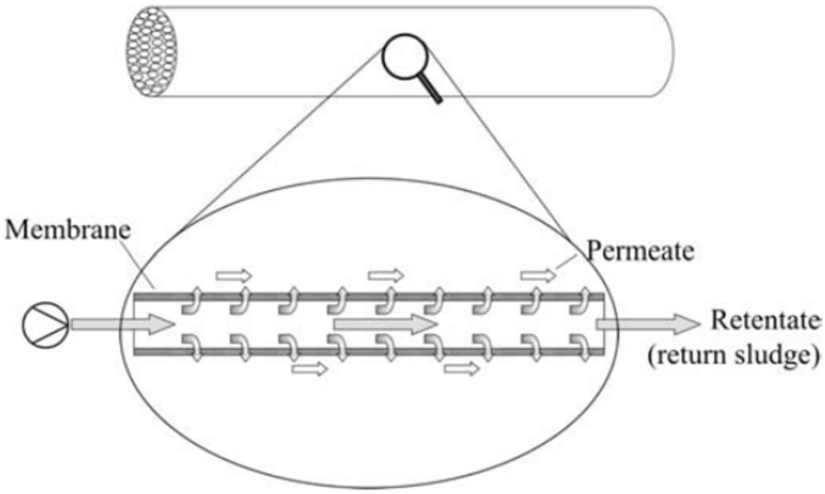

Figure 2.7. schematic view of a tubular membrane module (side-stream module) (Cornel and Krause 2008)

Over the years, the diameter of tubular membrane has been progressively reduced from more than 20 mm to 5 mm in order to increase the packing density. Tubular modules can have installation lengths of up to 6 m. These side-stream modules are operated at flow velocities of 1–4 m/s. Arrangement may be vertical or horizontal. Some systems are operated with additional aeration for fouling control. Tubular modules provide more direct hydrodynamic control at the membrane surface. Compared to submerged modules, the flux per surface area is higher, however, at the cost of a higher specific energy demand (kWh/m^3). The modules can be back pulsed or flushed to clean the membrane surface. Some tubular membranes and a schematic view of a tubular module are shown in figure 2.6 and figure 2.7, respectively (Cornel and Krause 2008).

In tubular modules, hair and/or long fibers may accumulate in or even block the inflow region where the mixed liquor enters the tubes. This results in a higher energy demand and/or a flux decline (Cornel and Krause 2008).

The difference of the three membrane configurations can be seen in table 2.3.

Configuration	Turbulence promotion	Back flushable	Packing density (m^2/m^3)	price (per m^2)
Hollow fiber	Poor	YES	600–1200	LOW
Plate	Fair	Generally No	100–250	Low-high
tubular	Good	YES	20–90	High

Table 2.3 comparison of different membrane configurations (Cornel and Krause 2008, Simmon and Clair 2011)

2.1.3 Membrane materials

There are two types of membrane material, i.e. polymeric and ceramic. Metallic membrane filters also exist but 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, but generally nearly always comprise a thin surface layer which provides the required perm-selectivity on top of a more open, thicker porous support layer which provides mechanical stability (figure 2.8). Classic membrane is thus anisotropic in structure, having symmetry only in the plane orthogonal to the membrane surface (Simon et al. 2008).

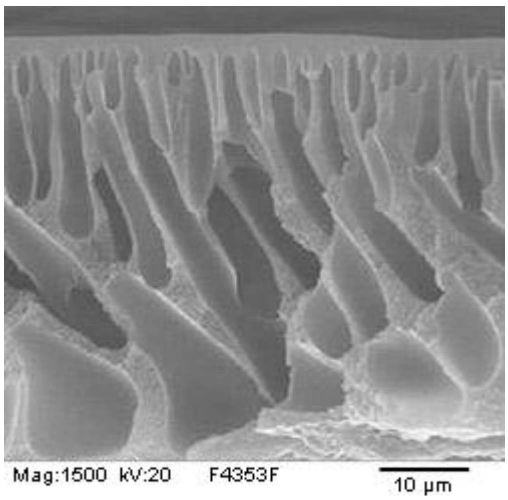

Figure 2.8 cross-section of a membrane

Polymeric membranes are also usually fabricated both to have a high surface porosity, and narrow pore size distribution to provide a high throughput and selectivity. The membrane must also be mechanically strong, i.e. to have structural integrity. Lastly, the material normally have some resistance to chemical attack, i.e. extremes of temperature, pH and/or oxidant concentrations that normally arise when the membrane is chemically cleaned, and should ideally offer some resistance to fouling (Simon et al. 2008).

Whilst, in principal, and polymer can be used to form a membrane ,only a limited number of materials are suitable for the duty of membrane separation, the most common materials are (1) polyvinylidene difluoride (PVDF), (2) polyethylsulphone (PES), (3) polyethylene (PE), and (4) polypropylene. All the above polymers can be formed, through specific manufacturing techniques, into membrane materials having desirable physical properties, and they each have reasonable chemical resistance. However, they are also hydrophobic, which makes them susceptible to fouling by hydrophobic matter in the bioreactor liquors they are filtering. This normally necessitates surface modification of the base material to produce a hydrophilic surface using such techniques as chemical oxidation, organic chemical reaction, and plasma treatment or grafting. It is this element that, if at all, most distinguishes one membrane product from another formed from the same base polymer. This modification process, the manufacturing method used to form the membrane from the polymer, most often PVDF for many MBR membranes, and the method for fabricating the membrane module from the membrane are all regarded as proprietary information by most suppliers (Simon et al. 2008).

2.1.4 Basis of membrane Operation

During MBR wastewater treatment, solid–liquid separation is achieved by MF or UF membranes. Feed water passes over the membrane surface and the product is called permeate, whereas the rejected constituents form concentrate or retentate (figure 2.9).

Mass balance of the solute in the process can be described by equation (2-1):

$$Q_f C_f = Q_p C_p + Q_c C_c \tag{2-1}$$

Where

Q_f : feed flow rate;
C_f : solute concentration in feed flow;
Q_p : permeate flow rate;
C_p : solute concentration in permeate;
Q_c : solute concentration in concentrate;
C_c : solute concentration in concentrate.

According to equation (2-2), membrane rejection of solutes can be calculated:

$$R = (C_f - C_p) / C_f \tag{2-2}$$

Where C_f represents concentration of solute in feed flow and C_p represents its concentration in permeate.

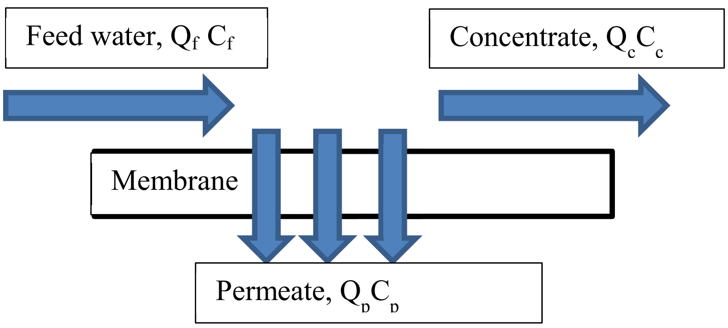

Figure 2.9. Basic principle of membrane filtration (Radjenović et al. 2008).

The fraction of feed flow converted to permeate is called yield, recovery or water recovery (S). Water recovery of the membrane process is given by equation (2-3):

$$Y = Q_p / Q_f \tag{2-3}$$

Recovery is normally close to 100% for dead-end filtration; however, it varies significantly for cross-flow filtration, which depends on the nature and design of membrane process. Permeate flux (usually denoted as *J*) is the volume of water passed through a unit area of membrane per unit of time and it is often normalized to a standard time. The common unit for *J* is usually L m^{-2} d^{-1}, and most available data for MBR is given in that manner rather than in SI units. MBR membranes generally operate at fluxes between 10 and 100 L m^{-2} h^{-1}. The flux is related to its driving force which is transmembrane pressure (TMP or ΔP) while the membrane performance can be estimated from the membrane permeability (K), which is calculated as permeate flux per unit of TMP and is usually given as L m^{-2} h^{-1} (Radjenović et al. 2008).

2.2 Membrane bioreactor

2.2.1 Introduction of membrane bioreactor

The application of biological wastewater treatment dates back to the late nineteenth century. It became a standard method of wastewater treatment by the 1930s. Both aerobic and anaerobic biological treatment technologies have been extensively applied to domestic and industrial wastewater treatment. In these technologies, biomass needs to be separated from liquid stream by the application of secondary clarifiers, in order to obtain clean effluent. However, secondary clarifiers have limited capacity of solid/liquid separation. Therefore, secondary settling clarifiers are normally large to guarantee good liquid/sludge separation effect. Nevertheless, the limited separation capacity cannot guarantee high sludge concentrations in reactors, which restricts the biological capacities of bioreactors.

To eradicate the disadvantage of conventional technologies, the biological technologies can be integrated with membrane technology. Membrane filtration and biological technologies can be efficiently applied together. The biological technologies converts dissolved organic matter into suspended biomass, which reduces membrane fouling. On the other hand, membranes not only replace secondary clarifiers for solid–liquid separation but also prevent visible particle from getting into effluent, which promotes high sludge concentrations in reactors (Hai and Yamamoto 2011). The application of MBR in municipal wastewater treatment has grown widely. This is due primarily to more stringent effluent water quality requirements, space constraints, lower operator involvement, modular expansion characteristics and consistent effluent water quality capabilities (AMTA 2007).

There are two kinds of configures of MBRs. While membrane is located in reactors, MBR is called submerged or immersed MBR. Alternatively, while membrane is located outside of a reactor, MBR is called sidestream MBR. Figure 2.10 shows a secondary settler, which is usually large, and a submerged MBR and a sidestream MBR (only membrane is shown).

Figure 2.10 MBR. Top: secondary settler; middle: submerged MBR; bottom: cross flow MBR (only membrane module is shown).

Side stream MBR

Side-stream configurations typically use tubular membranes. In the side-stream configuration, mixed liquid suspended sludge (MLSS) is pumped into membrane modules. Membrane fouling is controlled by a well-defined flow velocity, i.e. 4 m/s (Cornel and Krause 2008).

More recently, a Dutch company has developed an external membrane configuration consisting of cross-flow membrane modules that are mounted vertically, and relying on the combination of air and liquid flow to create high shear inside tubular membranes which are 5.2 mm in diameter. In this case, the liquid velocity is approximately 0.5 m/sec, the air flow translates to an airflow velocity in the range from 0.3 to 0.5 m/sec and the operating TMP is typically less than 15 percent of those in the conventional external membrane systems.

Recently, a German company introduced a lower velocity, lower TMP (versus conventional external membrane systems) configuration. The configuration consists of cross-flow, tubular membrane modules, which are mounted horizontally. Velocity of sludge in the tubular membranes is approximately 1.2 m/sec, and frequent membrane back pulsing is applied to maintain flux performance. No airflow is used inside the tubes (Sutton 2006).

The capital and operating costs associated with the membrane component of an MBR system are significantly affected by permeate flux. Conventional external membrane systems typically are operated at fluxes between 85 to 135 L/m^2 .h and at TMPs that are higher than 210 kPa. However, the non-conventional, back washable, lower velocity, lower pressure, air-lift systems and liquid only based systems (i.e., non-conventional external membrane systems), typically are operated at fluxes between 40 to 80 L/m^2 .h and at TMPs less than 30 kPa (Sutton 2006).

Submerged MBR

In the immersed or internal membrane MBR system, membranes are directly submerged in bioreactor's mixed-liquor, preferably located in compartments or a separate tank coupled to the bioreactor to minimize membrane cleaning efforts. This configuration typically involves the use of polymeric membranes. The membranes are either vertically or horizontally oriented hollow fibers contained in a rectangular or tubular support structure, or vertically oriented flat sheets contained within a support structure. The mixed-liquor is located on the shell side of the membranes and the effluent is extracted into the lumen of the membrane. The driving force across the membrane is typically achieved by creating negative pressure on the lumen or permeates side of the membrane. The membrane component of this configuration involves substantially more membrane area per unit volume, compared to the membrane component of the external MBR configuration. Although the shear across the membrane fibers is increased by continuous or intermittent aeration and other methods are used in certain designs to minimize the build-up of solids on the membrane surface (e.g., frequent membrane back pulsing, intermittent permeation), the typical mean permeate fluxes in these systems are less than 35 L/m^2 .h (Sutton 2006).

2.2.2 Advantages and disadvantages of MBR technology

Compared to conventional biological wastewater treatment technologies, MBR is characterized by its advantages and disadvantages, which are summarized in table 2.4 and will be further selectively discussed below.

advantages	disadvantages
Small footprint	Relatively expensive to install and operate
High effluent quality	Frequent membrane monitoring and maintenance
No need of disinfection	Membrane fouling
High organic load	Higher operation cost
Low sludge production rate	Limitations imposed by pressure, temperature, and pH requirements to meet membrane tolerances
Fast start up	Membranes may be sensitive to some chemicals
Less influence from bulk sludge	Less efficient oxygen transfer caused by high MLSS concentrations
Easy Modulation	Treatability of surplus sludge is questionable
Easy operation	Require pre-treatment
High SRT	
Upgrade available plants	

Table 2.4 summary of advantages and disadvantages of MBR (Melin et al. 2006, Tom et al. 2000)

1. High effluent quality

MBR systems provide high effluent quality in a greatly simplified process. This requires only head works, biological processes, membrane filtration and disinfection to meet the most stringent water quality standards. More importantly, the effluent quality is highly consistent (AMTA 2007). Membrane permeate quality can be seen in table 2.5.

Because membrane retains most particulate matter, permeate is very low in total suspended solids, turbidity, BOD, and most pathogens (Daigger et al. 2005). Particulate, colloidal and high molecular weight organics are retained, providing a maximum opportunity for biological degradation of these compounds. Non-biodegradable compounds tend to be discharged together with residual sludge rather than with treated water. The application of MBR also eliminates concern of varying biomass settling characteristics (e.g., filamentous growth) and associated cost implications (e.g., polymer addition, chlorine addition to control filaments) (Sutton 2006).

Parameter	Removal efficiency (%)	Permeate quality	Wwtp effluent
TSS, mg/L	>99	n.d.	5–8
Turbidity, NTU	98.8–100	<0.5	
COD, mg/L	89–98	10–30	30–70
BOD, mg/L	>97	<5	4–15
NH_4^+-N, mg/L	80–90	<5–6	5–12
N_{tot}, mg/L	36–80	<10	
P_{tot}, mg/L	62–97	0.3–3	1–3
Total coliforms, CFU/100 mL	5–8 log	<100	
Faecal coliforms, CFU/100 mL	-	<20	
Bacteriophages, PFU/100 mL	>3.8 log	-	

Table 2.5 MBR permeate quality (Melin et al. 2006)

2. Small footprint

One of the advantages of MBR is its compactness, because large sedimentation tanks are not required. An interesting parameter in this respect is the surface-overflow rates for the two systems. The overflow rate of a secondary clarifier is defined as the volume of water that can be treated per square meter of tank. In practice, values around 22 m/d are used. For an MBR filtration tank, an overflow rate can also be estimated from the permeate flux and the membrane-packing density within the tank. Following this method, with an average permeate flux of 15 L/m^2 .h, the overflow rates of the membrane tanks are in the range 25–62 m/d which is up to 3 times higher than the overflow rate of a conventional secondary clarifier. Compared to an average overflow rate of 22 m/d with a secondary clarifier, the space consumption for sludge-water separation in an MBR is 10–60% lower when flux is 15 L/m^2 .h and 50–80% lower when flux is 25 L/m^2 .h (Hai and Yamamoto 2011).

Higher MLSS concentration in MBRs than that in conventional technologies results in a further reduction in footprint. MBR systems are highly space efficient because it combines space efficient membrane systems and is operated at increased mixed liquor concentrations (commonly 8,000–18,000 mg/l). This estimation does not take into account back flushing or relaxation periods, which reduces the overflow rate. Nevertheless, full-scale MBR plants also manifest these space-saving characteristics. (Hai and Yamamoto 2011).

3. Easy operation

MBR systems do not require any more significant operational attention, in each case much less than conventional activated sludge (CAS) process. With modern process-control equipment, such as programmable logic controllers, most of the operations can be automated (Daigger et al. 2005). MBR systems can operate largely unattended except for occasional routine performance checks and maintenance of mechanical components. A process control of an MBR system is reduced to monitoring the MLSS concentration, occasional adjustments of the chemical feed rates, and scheduling membrane recovery cleaning. Therefore, MBR is much better solution for the small plants where CAS is non-feasible due to its requirement for constant attention and monitoring (Radjenović et al. 2008).

4. High SRT

Largely unencumbered control of the SRT provides optimum control of the microbial population and flexibility in operation. Provides opportunity to consider design/operation of bioreactor at very short or very long SRT (e.g., 1 day or less, or greater than 30 days) as process requirements dictate versus concerns for achieving a flocculant biomass. A short SRT maximizes biomass production and its organic content which if the biomass is anaerobically processed, maximizes digester gas production and therefore its energy value. A long SRT favours aerobic digestion of bio-solids, which may be attractive under certain circumstances (Sutton 2006). High mixed liquor concentrations in the reactor allow wastewaters to be treated efficiently at long SRTs, minimizing biomass yield (Sutton 2006). In addition, rapid initial process start-up due to retention of all microbial seed material can be achieved, and slower growing organisms, such as nitrifying bacteria and those capable of degrading complex organics, can be readily maintained in MBRs (Sutton 2006).

5. Low sludge production

A low sludge organic load (F/M) ratio means that less substrate is available per unit of biomass. Part of the energy contained in the supplied substrate is used for maintenance functions that are independent of growth rate. When the energy supplied to the bioreactor is lowered, the biomass ceases to grow and to utilize the substrate for maintenance. In this manner, the sludge production in the process is much lower. Nevertheless, due to the low F/M ratio, there is a significant decrease of sludge production in MBR in comparison to CAS, which then decreases the cost of excess sludge handling (Radjenović et al. 2008).

Table 2.6 provides a general comparison of the sludge-production rates from different treatment processes. It should be noted that the primary sludge production in the case of the MBR is lower. The suited pre-treatment for the MBR is grids and/or sieves, and in an average, screened water was observed to contain 30% more solids than settled water. MBR sludge treatment is almost the same compared to CAS systems. The dewater-ability of waste-activated sludge from the MBR seems to pose no additional problem, compared to aerobic stabilized waste sludge from CAS (Hai and Yamamoto 2011).

Treatment process	Sludge production kg (kg BOD)$^{-1}$
Submerged MBR	0.0–0.3
Structure media biological aerated filter	0.15–0.25
Trickling filter	0.3–0.5
Conventional activated sludge	0.6
Granular media BAF	0.63–1.06

Table 2.6 sludge production in case of different treatment processes (Hai and Yamamoto 2011)

6. Upgrade available plants

MBRs allow for exceptional versatility in the design of new plants or the retrofitting of existing wastewater-treatment facilities, because membranes can be added in modules (Daigger et al. 2005). Therefore, MBR represents an attractive technology for upgrading and/or expanding an existing activated sludge system plagued by clarifier performance problems or excessive operational needs, or where site constraints dictate against addition of new structures (Sutton 2006).

7. operation problems

Although many results from research activities have been applied in these full-scale plants, lower membrane permeability than anticipated was observed in many actual operations. This indicates that results from bench or pilot scale experiment are not always correlated to the application in full-scale plants. Additional research on full-scale plants in long-term operation could provide valuable insight on this issue. Some problems detected in practical MBRs are listed below:

1. Bioreactor temperature impacting performance in cross flow MBR;
2. Entrained air impacting suction pump operation in cross flow MBR;
3. Membrane fouling due to build-up of oil grease in the bioreactor;
4. Membrane fouling during permeate back pulsing;
5. entrained air impacting suction-pump operation;
6. Bioreactor foaming;

7. Inefficient aeration due to partial clogging of aerator holes;
8. No significant decrease of bio solid production;
9. scale build up on membrane and piping;
10. Corrosion of concrete, hand rails, and metallic components due to corrosive vapour produced during high temperature;
11. membrane delamination and breakage during cleanings;
12. odour from screening, compaction, drying beds, and storage areas (although normally less than in CAS);
13. Failure of control system.

2.2.3 Evaluation of MBR potentials

At present MBR treatment can be considered to be a proven technology for wastewater treatment. In the last 15 years significantly progress has been made in the design and operation of MBR's. One of the reasons is the availability of government funding, which has made it possible to conduct extensive research on a practical scale. An example is the research project at the Beverwijk WWTP in the Netherlands where membrane suppliers, engineering firms and water boards have worked together to test several pilot MBR systems during a two year period (2001–2002), with additional pilot plants tested afterwards. The main results of this project were increased reliability of operation, increased membrane life time, reduced energy cost and a better understanding of the nature, prevention and removal of membrane fouling (van Haandel and van der Lubbe 2007).

Several hundreds of small (5–50 $m^3.h^{-1}$) industrial installations have been built from 1990 onwards in Europe and North America. Also in Japan a large number (>1000) of small industrial and municipal MBR's have been constructed. However, large scale application of MBR to municipal waste water lagged behind due to the diluted nature of this waste water. From 1998 onwards, several municipal MBR's have been constructed, sometimes as a demonstration project with practical governmental funding (van Haandel and van der Lubbe 2007).

With membrane prices decreasing, MBR is becoming more competitive compared to conventional treatment, but still the number of municipal installation in use is very limited. The main reasons to implement a MBR instead of conventional activated sludge system are (van Haandel and van der Lubbe 2007):

1. Limited availability of space;
2. Strict effluent limits (i.e. discharge of effluent in a vulnerable area or possible reuse of effluent)
3. Difficulties for solid/liquid separation in the final settler.

Disadvantages of the MBR process compared to a conventional system are (van Haandel and van der Lubbe 2007):

1. Increased investment and operation cost;
2. Increased complexity of the system, requiring skilled operation and a considerable amount of automation/ instrumentation;
3. Ecological considerations such as the use of chemicals for cleaning and the increased consumption of energy (greenhouse gas emission).

As long as the effluent from a municipal waste water treatment plant has to comply with effluent limits similar or less stringent to those currently applicable in the Netherlands (i.e. N<10 mg N L^{-1}), P<1mg P L^{-1}, and X_{te}=20 mg TSSL^{-1}), treatment with conventional activated sludge systems will be sufficient and in many cases more competitive. For low strength municipal wastewater, annualised investment cost of an MBR system are stil about fifty percent higher than those of a conventional installation. This is not only due to the cost of installing (and replacing) the membrane. Other factors are the need for more extensive pretreament, a much higher degree of automation and higher energy requirments (van Haandel and van der Lubbe 2007).

The cost difference might be reduced in the future, as the continuing competition between membrane suppliers might cause membraen prices to decrease further, and if land prices go up. For difficult and/ or high strength industrial wastwaters,MBR reactors are often already an attractive alternative, as in this case the cost fraction of the membrane part will be limited compared to that of the biological treatment volume (van Haandel and van der Lubbe 2007).

Should stricter effluent limits be applied in the future, MBR will certainly become more attactive. This may be the case in Europe as a result of the EU water framework derective. In the period from 2000 to 2015, the membrane states will have to increase the water quality of surface and ground water. This amy result in stricter effluent limits. However, MBR will then have to compete with other systems capable of delivering the required effluent quality such as conventional treatment followed by polishing steps such as sand filtration (van Haandel and van der Lubbe 2007).

If the effluent of the wastewater treatmnet plant is to be resued as a high quality water source (e.g.as process water, boiler feed make-up water,cooling make up water or even potable water), once again the MBR is an attractive alternative,as an effluent free of solids is produced, which can be directly processed in downstream processing operation such as nanofiltration or reverse osmosis (van Haandel and van der Lubbe 2007).

Finally, if available space is limited, MBR might be very interesting as well. This may be the case for many industrial locations, but certainly also for minicipal wastewater treatment. In a large number of cities ini developing countris, the rapid expansion of the population has two main effects:

1. the existing conventional wastewater treatment plants are overloaded and
2. there will often be no space available for expansion of the wastewater treatment plant. Retrofitting the old conventional treatment plant into an MBR system increases treatment capacity by a factor of 3–4 without any additional space requriements (van Haandel and van der Lubbe 2007).

2.2.4 Side stream MBR versus submerged MBR

As mentioned above, there are two configures of MBR applied in practise, i.e. submerged MBR and side stream MBR. Each has its advantages and disadvantages, which results in different application potentials. Relevant discussion is provided below.

1. Advantages of cross flow MBR compared to submerged MBR are (van Haandel and van der Lubbe 2007):

The investment costs of cross flow membranes are lower than those of submerged membranes, but the energy consumption of cross flow membranes is much higher. This is partly compensated by the energy required for the aeration of the submerged membranes.

1. Superior operational reliability and significantly reduced vulnerability to membrane fouling, which makes application to difficult waste waters easier;
2. The performance of cross flow membranes is less dependent on sludge characteristics than that of submerged membranes;
3. Cross flow membranes are also much more robust than submerged membranes; this allows application at higher temperature (up to 60 °C) and also allows more intense cleaning;
4. Cross flow membranes are easily accessible, which facilitates maintenance;
5. External MBRs have maintained interest from the research community for specific application areas. This is apparent from Fig. 6, as a steady number of research papers (around 12 per year) were published on this configuration in the last 7 years. Many of these papers deal with particular industrial wastewater applications or study fundamental aspects related to cross-flow membrane filtration (Yang et al. 2006). External MBRs were considered to be more suitable for wastewater streams characterized by high temperature, high organic strength, extreme pH, high toxicity and low filterability. Studies on the treatment of municipal wastewater with MBRs mostly utilized the submerged configuration (Yang et al. 2006);
6. Cross flow membranes are operated at a much higher differential pressure over the membrane (4–7 bar) than submerged membranes (0.1–0.4 bar);

7. Cross flow membranes may be operated at higher suspended solids concentration: 15–50 kg $TSS.m^{-3}$, compared to 12–20 kg $TSS.m^{-3}$ for submerged membranes. However, operating at such a high biomass concentration will cause other problems: the oxygen transfer rate can be a limiting factor and excessive foaming may become an issue. Sometimes pure oxygen is used for aeration when an MBR is operated at a very high biomass concentration.

2. Disadvantages of side stream MBR compared to submerged MBR

A disadvantage of cross flow MBR is that this configuration is less suitable to deal with large fluctuation in feed flow rate. Submerged membranes can be operated (temporarily) at a higher flux than their normal operating flux by simply increasing the flow rate of the permeate pump, although this will result in an increase in the differential pressure over the membranes. The operational flexibility of subered membrane systems is very convenient for MBR reactors treating highly variable flows such as municipal sewage with a high ratio between rainwater flow and dry weather flow (van Haandel and van der Lubbe 2007).

Submerged MBRs have proven to be more cost- and energy-effective than tubular side-stream modules (Cornel and Krause 2008). Due to the absence of a high-flow recirculation pump, submerged MBRs consume much lower power than external MBRs. This was the primary driver for propelling submerged MBRs into the purview of large-scale wastewater treatment plants in a few dozens of countries around the world. In the last 3 years, many more studies were performed on submerged MBRs than external MBRs (Yang et al. 2006). The cost of oxygen demand is superior in MBR. Energy consumption of MBR comes from power requirements for pumping feed water, recycling retentate; permeate suction (occasionally) and aeration. The two MBR configurations have substantial differences in terms of aeration. In the side-stream configuration, aeration is supplied by fine bubble aerators that are highly efficient for supplying oxygen to the biomass. In submerged MBRs, the aeration mode is turbulent and cross-flow is generated, which scours the membrane surface and provides oxygen to the biomass. Aeration cost in the latter-mentioned configuration represents around 90% of the total costs, whereas in side-stream MBR, only~20% derives from it. However, energy consumption of the side-stream system is usually two orders of magnitude higher than that of submerged systems. These low costs of submerged MBRs are associated with low fluxes, which in turn increase capital costs and footprints. Also, packing density influences the final cost of MBR: low packing densities of membrane modules mean that higher specific area of membrane is required to produce the same flux, which increases the energy requirements (Radjenović et al. 2008).

Submerged membranes are operated in a constant flux variable pressure mode: i.e. the differential pressure over the membrane increases in time due to fouling of the membrane surface, while the membrane flux remains constant (as it is set by the permeate pump capacity). For cross flow membrane this is exactly the opposite: the applied pressure remains the same but the membrane flux decreases in time (due to fouling).

Table 2.7 contains a summary of information comparing external and internal membrane MBR system configurations.

Comparative factor	Side stream MBR	submerged MBR
Membrane area requirement	Characterized by higher flux and therefore lower membrane area requirement	Lower flux but higher membrane packing density (i.e. membrane area per unit volume)
Space requirements	Higher flux membrane with bioreactor operating at high VSS concentration and skidded assembly construction, results in compact system	Higher membrane packing density and operation at bioreactor VSS concentration of 10 g/L or greater translates to compact system
Bioreactor and membrane component design and operation dependency	Bioreactor can be designed and operated under optimal conditions including those to achieve biological N and P removal, if required	Design and operation of bioreactor and membrane compartment or tank are not independent. High membrane tank recycle required (e.g. recycle ratio 4) to limit tank VSS concentration build up.
Membrane performance consistency	Less susceptible to changing wastewater and biomass characteristics	More susceptible to changing wastewater and biomass characteristics requiring alteration in membrane cleaning strategy and /or cleaning frequency

Comparative factor	Side stream MBR	submerged MBR
Recovery of membrane performance	Off-line cleaning required every 1 to 2 months. simple, automated procedure normally requiring less than 4 hours	Off line "recovery" cleaning required every 2 to 6 months. A more complex procedure requiring significantly more time and manual activity, at least on occasion may be required (i.e. physical membrane cleaning)
Membrane life or replacement requirements	Results imply an operating life of 7 years or more can be achieved with polymeric prior to irreversible fouling. Operating life of ceramics much longer.	Results imply an operating life of 5 years may be possible prior to irreversible fouling and/or excessive membrane physical damage.
Full scale application status	Conventional membrane based systems have a very long track record. Few nonconventional systems in operation in the U.S.	Full scale application widespread in the U.S.
economics	Non-conventional designs translate to comparable power costs. Comparable capital cost at least at lower wastewater feed rates (e.g., approaching 1893 m^3/day).	Power and capital cost advantage at higher wastewater feed rates.

Table 2.7 comparison of side stream and submerged (Sutton 2006)

It can be concluded that in general, submerged MBR is preferred over cross flow MBR. As shown in figure 2.11, submerged MBR share much more market than cross flow MBR. However, cross flow MBR can certainly be considered an attractive alternative to submerged MBR for small scale application (up to an influent flow of 10–20 m^3/h) treating difficult wastewater waters (van Haandel and van der Lubbe 2007).

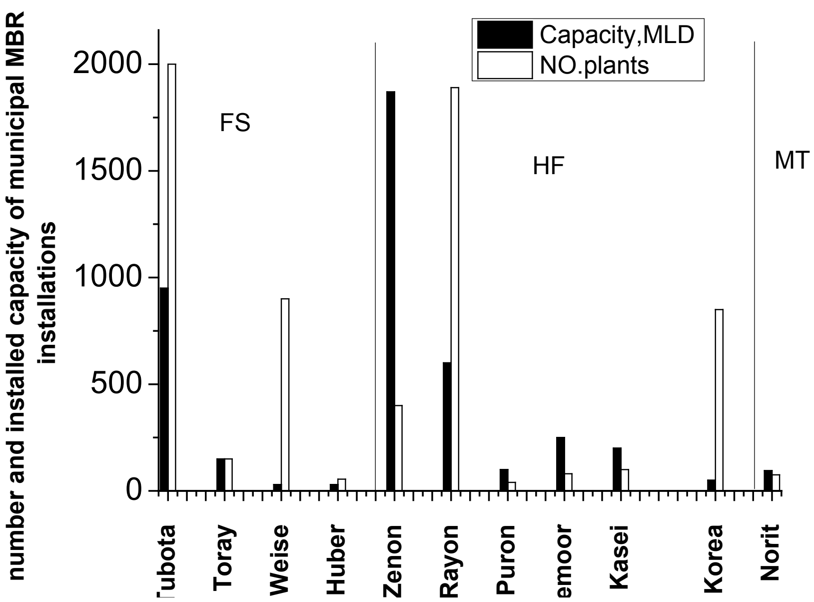

Figure 2.11 MBR municipal market (Simmon and Clair 2011)

2.2.5 Commercialized MBR Formats

The sidestream approaches are also divided into two formats – the long-established traditional method of cross flow, now used only for the most difficult feeds, and the newer concept of airlift, which uses air to recirculate the feed and thereby significantly reduces energy demand. Both sidestream formats use tubular membranes. Cross flow is more energy intensive – very high cross-flow velocities (up to 5–6 m^3 /h) may be necessary to control the fouling; but for the more difficult feeds, it may be the only option that works reliably. Airlift is a more cost-effective way of improving mass transfer through the creation of slug-flow conditions in the lumen of the membrane tubes, but there is a limit to how much air flow can be used while retaining slug-flow conditions. Airlift technology has a power cost similar to that of the submerged technology. As mentioned earlier, the energy cost of cross flow prohibits it as a treatment option for any application other than small scale or where there is no other treatment option. However, the airlift has very low energy use, and may even undercut the energy requirements of the submerged options, due to the advantage of containment of the feed inside the tubular membrane. Since airlift eliminates operator contact and has good operational characteristics, it may as well make a major impact on the MBR market in the long run. It was argued that the airlift format may find applications throughout a broader range than the submerged formats (Hai and Yamamoto 2011).

The submerged format is available with modules either in a flat-sheet configuration or as hollow fibers or capillary membranes. Originally, the favoured concept was to submerge the modules directly into the bioreactor for simplicity. However, in order to gain better control of the balance between the biological and filtration-treatment capacity, it is now more common to use the membrane in an external membrane tank. The external arrangement allows the size and design of the membrane tank to be optimized independently, with practical advantages for operation and maintenance (Hai and Yamamoto 2011).

In general, submerged MBR formats based on hollow fibers have been found to provide the most cost-effective solution for large-scale, easy-to-treat applications. Technology has been developed with optimized packing density and aeration bubble size to achieve stable performance at minimum energy use. However, this format can experience operational difficulties due to fibers becoming matted close to the potted ends and therefore pre-treatment and removal of hairs and fibers is essential. Hollow-fiber technology hence requires more instrumentation and control (Hai and Yamamoto 2011).

The submerged MBR formats based on flat sheets have been found to be cost effective for similar types of wastewater, but due to higher air use and lower compactness, tend to be selected for small- to medium-scale duties. The flat-sheet format has operational advantages in terms of plugging and cleaning, and has been used in somewhat more difficult feeds. Flat-sheet systems have the advantage of relatively low manufacturing cost compared to hollow-fiber systems. However, packing density tends to be significantly lower than a hollow-fiber system (e.g., by a factor of 2.5–3 times). Therefore, flat-sheet systems tend to have a cost advantage for small to medium-scale systems, whereas hollow fiber becomes more attractive at large scale due to the footprint advantage. The comparison is made more complicated, however, since aeration costs for hollow-fiber systems are often lower (Hai and Yamamoto 2011).

This means that the most cost-effective solution for total treatment costs at medium scale is closely contested, and both approaches are found across the size range due to site-specific circumstances, which could favour either solution. It is estimated that a future market share could be as follows:

1. For municipal applications, it is expected that the hollow-fiber submerged configuration would be competitive for medium to large-size plants.
2. For small to medium sizes, flat-sheet technologies would have an advantage. However, in case of larger plants, or a plant refurbishment, the alternative membrane scheme (secondary/tertiary treatment followed by an MF/UF membrane filtration) is very likely to be cost competitive, unless high-cost land has to be purchased for the construction. This multi-barrier scheme will also be easier to control and to optimize because of the disconnection of the treatment steps. It will also be associated with the lowest risk in relation to the membrane operation, as the membranes will be operated under smooth hydrodynamic conditions in terms of particle matter, turbulence, and backwash regime. The airlift format has been developed as a low-energy alternative to the energy-intensive cross-flow sidestream format, which has been used historically for the most difficult feeds (Hai and Yamamoto 2011).

3 Membrane fouling and its reduction

The aim of this chapter is introduce the main problem regarding the operation of membrane bioreactor, i.e. membrane fouling and its control method

3.1 Membrane Fouling

3.1.1 Membrane fouling phenomena

Although MBR has become a reliable alternative to CAS processes and an option of choice for many domestic and industrial applications, membrane fouling and its consequences in terms of plant maintenance and operating costs limit the widespread application of MBRs. Membrane fouling can be defined as the undesirable deposition and accumulation of microorganisms, colloids, solutes, and cell debris within pores or on membrane surface (Meng et al. 2009). It results from the interaction between the membrane material and the components of the activated sludge liquor, which include biological flocs formed by a large range of living microorganisms along with soluble and colloidal compounds. Thus, it is not surprising that the fouling behaviour in MBRs is more complicated than that in most membrane applications. The suspended biomass has no fixed composition and varies with both feed water composition and MBR operating conditions employed. Accordingly, although many investigations of membrane fouling have been published, the diverse range of operating conditions and feed water matrices employed, and the limited information reported in most studies on the biomass composition in suspension or on the membrane, have made it difficult to establish any generic behaviour pertaining to membrane fouling in MBRs (Hai and Yamamoto 2011).

3.1.2 Classifications of membrane fouling

1. Pore blockage and cake layer formation

During initial filtration, colloids, solutes, and microbial cells pass through and deposit inside the membrane pores. However, during the long-term operation of MBRs, the deposited cells multiply and yield extracellular polymeric substance (EPS), which clog the pores and form a strongly attached fouling layer. Chemical cleaning is usually required to remove such fouling (Hai and Yamamoto 2011).

The formation of a cake layer which can be described as a porous media with a complex system of interconnected inter-particle voids has been reported as the major contributor to membrane fouling in MBRs. Such fouling is usually physically removable (Hai and Yamamoto 2011).

The formation of pore blockage and cake layer can increase filtration resistance.

2. Removable fouling and irremovable fouling

According to whether membrane fouling can be removed, the membrane fouling can also be classified into removable fouling and irremovable fouling.

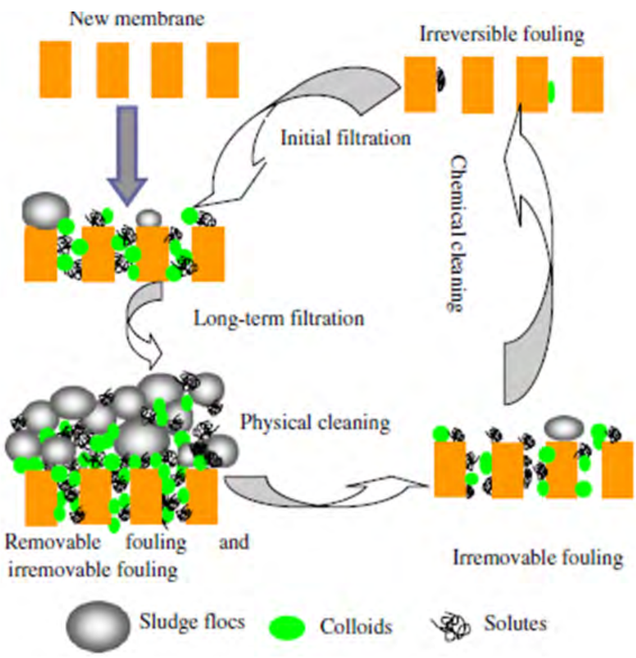

Figure 3.1 schematic illustration of the formation and removal of removal and irremovable fouling in MBRs (Meng et al. 2009)

As shown in figure. 3.1, the removable fouling can be easily eliminated by implementation of physical cleaning (e.g., backwashing) while the irremovable fouling needs chemical cleaning. The removable fouling is caused by loosely attached foulants; however, irremovable fouling is caused by pore blocking and strongly attached foulants during filtration. In general, removable fouling is attributed to the formation of cake layer, and the irremovable fouling is attributed to pore blocking (Meng et al. 2009).

MBR are normally operated under a constant flux. Since the fouling rate increases roughly and exponentially with the flux, MBR plants are operated at modest fluxes and preferably below the so-called critical flux. The critical flux concept assumes that there is a flux below which a decline of permeability with time does not occur, and above which fouling occurs. In MBR operations, critical flux is normally defined as the highest flux under which a prolonged filtration with constant permeability is possible. Critical flux is often determined by the flux step method, in which the flux is incrementally increased in number of steps with fixed duration, and the increase in TMP is recorded. It is then possible to observe the apparent flux where fouling occurs, which is observed as a significant TMP increase (Radjenović et al. 2008).

3. Biofouling, organic fouling, and inorganic fouling

A definition is based on fouling components. The fouling in MBRs can be classified into three major categories: **biofouling, organic fouling, and inorganic fouling**, although, in general, all of them occur simultaneously during membrane filtration of activated sludge.

Biofouling

Biofouling refers to the deposition, growth, and metabolism of bacteria cells or flocs on the membranes. Biofouling may start with the deposition of individual cell or cell cluster on the membrane surface, after which the cells multiply and form a biofilm (Hai and Yamamoto 2011).

Organic fouling

Organic fouling in MBRs refers to the deposition of biopolymers on membrane surface. Due to their small sizes, soluble biopolymers deposit onto membranes readily in comparison to large particles (e.g., colloids and sludge flocs). Powerful analytical tools have confirmed that soluble microbial product (SMP) or EPS is the origin of organic fouling in MBR. Inorganic elements such as Mg, Al, Fe, Ca, Si, etc. and metals can enhance the formation of biofouling and organic fouling, and can together form a recalcitrant cake layer (Hai and Yamamoto 2011).

Inorganic fouling

In general, membrane fouling in MBRs is mainly governed by biofouling and organic fouling rather than by inorganic fouling, although all of them take place simultaneously during membrane filtration of activated sludge. On inorganic membranes inorganic fouling may occur more easily. In general, a cake of inorganic matter can be irremovable due to the cohesive properties. Sometimes, the fouling caused by inorganic scaling is not easy to be eliminated even by chemical cleaning (Meng et al. 2009).

The inorganic fouling can form through two ways: chemical precipitation and biological precipitation. A great number of cations and anions such as Ca^{2+}, Mg^{2+}, Al^{3+}, Fe^{3+}, CO_3^{2-}, SO_4^{2-}, PO_4^{3-}, OH^- and others are present in MBRs. Concentration polarisation will lead to higher concentration of retained salts on the membrane surface. Chemical precipitation occurs when the concentration of chemical species exceeds the saturation concentrations due to concentration polarisation. Additionally, the fouling layer on membranes can protect the surface layer from shear stress as biofilm or biocake is elastic in nature leading to greater concentration polarisation and precipitation of inorganics. Carbonates are one kind of the predominant salts in inorganic fouling. The aeration and the CO_2 produced by microorganisms can affect the super-saturation of carbonates and the pH of the sludge suspension. The carbonates of metals such as Ca, Mg, and Fe can increase the potential of membrane scaling (Meng et al. 2009).

Biological precipitation is another contribution to inorganic fouling. The biopolymers contain ionisable groups such as COO^-, CO_3^{2-}, SO_4^{2-}, PO_4^{3-}, OH^-. Metal ions can be easily captured by these negative ions. In some cases, calcium and acidic functional groups (R–COOH) can form complexes and build a dense bio-cake layer or gel layer that may exacerbate flux decline. When the metal ions in treated water pass through the membranes, they could be caught by the bio-cake layer via complexing and charge neutralisation and then accelerate membrane fouling. Metal ions play a significant role in the formation of fouling layers, which can bridge the deposited cells and biopolymers and then form a dense cake layer. There exists a synergistic interaction among biofouling, organic fouling and inorganic fouling (Meng et al. 2009).

3.1.3 Long term fouling development

It is evident that even sub-critical operation inevitably leads to fouling. This fouling is often reported to follow a two-stage fouling pattern, which includes slow TMP increase over a long period of time, followed by a rapid increase after some critical time period. Fouling in MBR under sub-critical conditions three stages include (Radjenović et al. 2008):

1. Initial conditioning fouling
2. Slow fouling
3. Sudden TMP jump

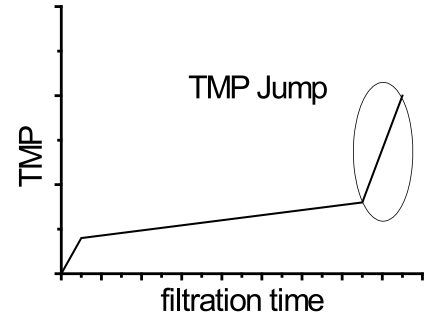

Figure 3.2 schematic illustration of the occurrence of TMP jump (Meng et al. 2009).

Initial fouling takes place between the membrane surface and soluble components of a mixed liquor. This fouling is usually rapid (measured in hours. In the second stage, slow fouling, the membrane surface is gradually covered by biopolymers such as EPS, which changes the properties of the membrane surface and makes attachment of the microbial flocs to the membrane surface easier. Thus, biofilm growth on the membrane surface may be promoted. Over time, complete or partial pore blocking takes place. This blocking is expected to be inhomogeneous since the air and the liquid flow are distributed unevenly in MBR. With regions of membrane more fouled than others, flux locally varies, thus exceeding the critical flux in some areas of the membrane surface, which then leads to a sudden TMP jump characteristic for operation above the critical flux. The other explanation for the sudden TMP jump may be the change of properties of the fouling cake on the membrane surface due to its compression (Radjenović et al. 2008). Figure 3.2 shows the schematic illustration of the occurrence of TMP jump.

It can be concluded that the interactions between TMP jump and these operating parameters are very complex, and TMP jump occurs inevitably during long-term operation of MBRs. Thus, the overall goal of fouling control is to retard the occurrence of the TMP jump via modifying sludge characteristics or decreasing membrane flux (e.g., operation below critical flux) (Meng et al. 2009).

3.1.4 Parameters influencing MBR fouling

All the parameters involved in the design and operation of MBR processes have impacts on membrane fouling. While some of these parameters have a direct influence on MBR fouling, many others result in subsequent effects on phenomena exacerbating fouling propensity. However, three main categories of factors can be identified – membrane and module characteristics, feed and biomass parameters, and operating conditions (Hai and Yamamoto 2011). Figure 3.3 summarize a number of factors that can influence the membrane fouling. The factors will be further briefly discussed below.

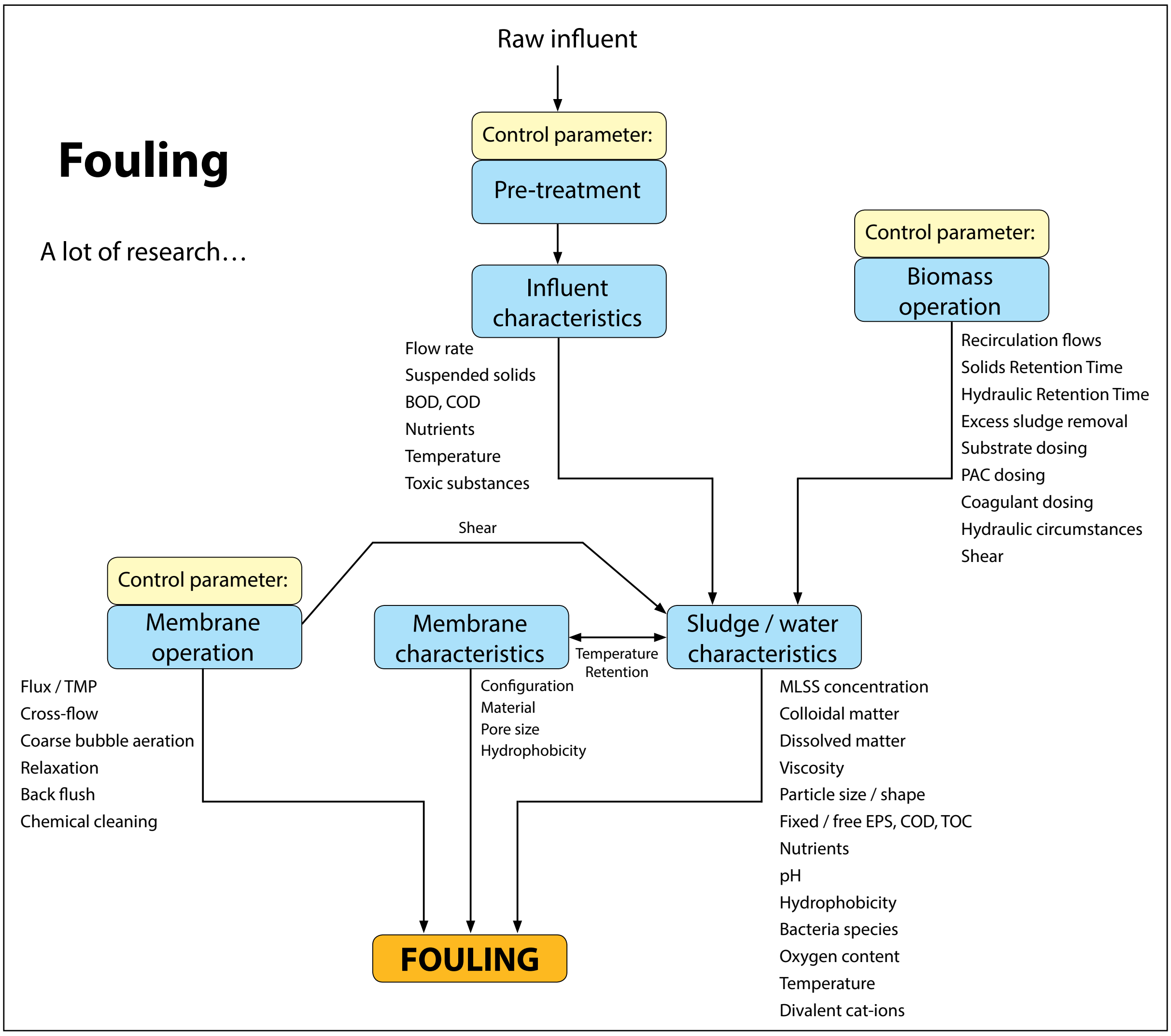

Figure 3.3 parameters influencing membrane fouling.

1. Membrane characteristics

1) Physical parameters

- Pore size and distribution

Studies revealed that the pore size alone could not predict hydraulic performances. The effects of pore size (and distribution of pore size) on membrane fouling are strongly related to the feed solution characteristics and in particular the particle size distribution. The complex and changing nature of the biological suspension present in MBR systems and the large pore-size distribution of the membrane generally used in MBR systems are the main reasons for the undefined general dependency of the flux propensity on pore size. It is generally expected that smaller-pore membranes would reject a wider range of materials, and the resulting cake layer would feature a higher resistance compared to large-pore membranes. However, this type of fouling is easily removed during the maintenance cleaning than fouling due to internal pore clogging obtained in larger-pore membrane systems. The chemically removable fouling, due to the deposition of organic and inorganic materials onto and into the membrane pores, is the main cause of the poor long term performances of larger pore-size membranes. However, the opposite trend is sometimes reported. The duration of the experiment and other operating parameters such as cross-flow velocity and constant pressure or constant flux operation have a direct influence on the determination of the optimization of the membrane pore size and are responsible for contradictory reports in the literature (Hai and Yamamoto 2011).

- Porosity/roughness.

Membrane roughness and porosity along with membrane microstructure, material, and pore-size distribution were suggested as potential reasons for the different observed fouling behaviours. For instance, a track-etched membrane, with its dense structure and small but uniform cylindrical pores, featured the lowest resistance due to pore fouling in contrast to the other membranes having interwoven sponge-like highly porous network. Other studies have pointed out the importance of pore-aspect ratio (mean major-axis length/mean minor-axis length) or roughness on fouling in an MBR (Hai and Yamamoto 2011).

- Membrane configuration.

In submerged MBR processes, the membrane can be configured as vertical flat plates, vertical or horizontal hollow fine fibers (filtration from out to in) or, more rarely as tubes (filtration from in to out). Each of hollow-fiber and flat-sheet membrane types has specific footprint and air scouring and chemical cleaning requirement, which may favour one process over another for a given application. Nevertheless, hollow-fiber modules are generally more economical to manufacture, provide high specific membrane area, and can tolerate vigorous backwashing. For low-flux operation, hollow fibers are attractive due to their high packing density. A higher fiber-packing density would increase productivity; however, increasing the packing density may lead to severe interstitial blockage due to the impeded propagation of air bubbles toward the core, limiting their effect on fouling limitation (Hai and Yamamoto 2011).

2) Chemical parameters.

- Hydrophobicity.

The influence of the membrane hydrophobicity on the early stage of the fouling formation may be significant; however, this parameter is expected to play only a minor role during extended filtration periods in MBRs. Once initially fouled, the membrane's chemical characteristics would become secondary to those of the sludge materials covering the membrane surface. Nevertheless, because of the hydrophobic interactions occurring between solutes, microbial cells and membrane material, membrane fouling is expected to be more severe with hydrophobic rather than hydrophilic membranes, although different results have also been reported. In many reported studies, change in membrane hydrophobicity often occurs with other membrane modifications such as pore size and morphology, which make the correlation between membrane hydrophobicity and fouling more difficult to assess (Hai and Yamamoto 2011).

- Materials.

The large majority of the membranes used in MBRs are polymeric based. A direct comparison between polyethylene (PE) and polyvinylidene fluoride (PVDF) membranes clearly indicated that the latter leads to a better prevention of physically irremovable fouling and that PE membrane fouled more quickly. Although featuring superior chemical, thermal, and hydraulic resistances, ceramic and stainless steel membrane modules are not the preferred option for MBR applications due to their high cost (around an order of magnitude more expensive than the polymeric materials) (Hai and Yamamoto 2011).

2. Feed–biomass characteristics.

1) Nature of feed and concentration

Fouling in the MBR is mostly affected by the interactions between the membrane and the biological suspension rather than wastewater itself. Nevertheless, the fouling propensity of the wastewater has to be indirectly taken into consideration during the characterization of the biomass, as the wastewater nature can significantly influence the physicochemical changes in the biological suspensions, which in turn may aggravate fouling (Hai and Yamamoto 2011).

2) Biomass fractionation.

Although the relatively low fouling role played by the suspended solids (biofloc and the attached EPS) compared to those of the soluble and colloids (generally defined as soluble microbial products or SMP) is usually reported, the reported relative contribution of the SMP to overall membrane fouling ranges from 17% to 81%. These wide discrepancies may be explained by the different operating conditions and biological states of the suspension used in the reported studies. Although an interesting approach for studying MBR fouling, the fractionation experiments neglect any coupling or synergistic effects which may occur among the different components of the biomass (Hai and Yamamoto 2011).

3) MLSS concentration.

Although the increase in MLSS concentration has often been reported to have a mostly negative impact on the MBR hydraulic performances, controversies exist. The existence of threshold values above or below which the MLSS concentration has a negative influence was also reported. Figure 3.4 depicts the influence of shift in MLSS concentration on flux as reported in different studies. Nowadays, information on additional biomass characteristics (e.g., composition and concentration of EPS) is deemed necessary to furnish a comprehensive picture. On the other hand, it was shown that the extent of fouling was independent of MLSS concentration itself, and was rather more influenced by the efficiency of the fouling prevention strategies adopted (Hai and Yamamoto 2011).

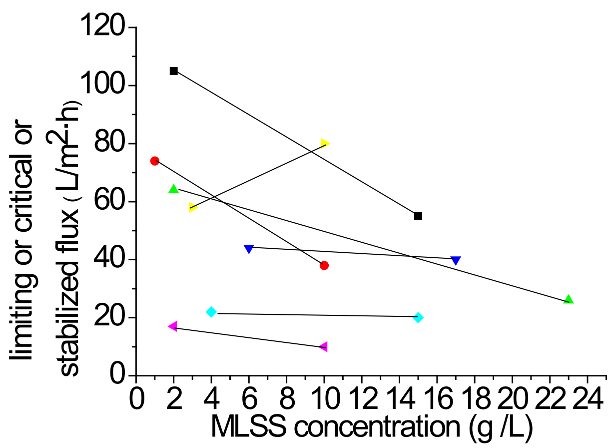

Figure 3.4 Influence of shift in MLSS concentration on flux (fouling) (Hai and Yamamoto 2011).

4) Viscosity.

The importance of MLSS viscosity is that it modifies bubble size and can dampen the movement of hollow fibers in submerged bundles. The net result of this phenomenon would be a greater rate of fouling. Increased viscosity also reduces the efficiency of mass transfer of oxygen and can therefore affect DO; fouling, as discussed later, tends to be worse at low DO. Critical MLSS concentrations have been reported in the literature above which, suspension viscosity tends to increase exponentially with the solid concentration (Hai and Yamamoto 2011).

5) Temperature.

Temperature may impact membrane filtration by increasing fluid viscosity, causing deflocculation of biomass and higher EPS secretion, reducing biodegradation rate, etc (Hai and Yamamoto 2011).

6) Dissolved oxygen.

The effects of DO on MBR fouling are multiple and may include changes in biofilm structure, SMP levels, and floc-size distribution. The average level of DO in the bioreactor is controlled by the aeration rate, which not only provides oxygen to the biomass but also tends to limit fouling formation on the membrane surface. Optimum aeration would result in lower specific cake resistance of the fouling layer featuring larger particle sizes and greater porosity. Therefore, in general, higher DO tends to lead to better filterability, and lower fouling rate (Hai and Yamamoto 2011).

7) Floc size

The floc-size distribution obtained with the MBR sludge is lower than the results generally obtained from conventional activated sludge plants. Unlike in CAS systems, the effective separation of suspended biomass from the treated water is not critically dependent on aggregation of the microorganisms, and the formation of large floc. However, independent of their size, biological floc play a major role in the secretion of EPS and formation of the fouling cake on the membrane surface (Hai and Yamamoto 2011).

It is generally accepted that fouling caused by suspended solids is less than that of supernatant. With regards to the fouling mechanism, it is generally assumed that soluble and colloid materials are responsible for pore blocking, while suspended solids contribute mainly to cake layer resistance. Again, one has to be aware of the fact that biomass itself is responsible for a composition of soluble and colloid material in the liquid phase, and that composition of supernatant is a function of the physiological state of the suspended phase, i.e., biomass (Radjenović et al. 2008).

8) Extracellular polymeric substances.

The term EPS is used as a general and comprehensive concept for different classes of macromolecules such as polysaccharides, proteins, nucleic acids, (phosphor-) lipids, and other polymeric compounds which have been found at, or outside, the cell surface and in the intercellular space of microbial aggregates. EPS are the construction materials for microbial aggregates such as biofilms, flocs, and activated sludge liquors. The functions of EPS matrix are multiple and include aggregation of bacterial cells in flocs and biofilms, formation of a protective barrier around the bacteria, retention of water, and adhesion to surfaces. With its heterogeneous and changing nature, EPS can form a highly hydrated gel matrix in which microbial cells are embedded. Therefore, they can be responsible for the creation of a significant barrier to permeate flow in the membrane processes. Contemporary literature is replete with reports identifying EPS as a major fouling parameter. On the other hand, since the EPS matrix plays a major role in the hydrophobic interactions among microbial cells and thus in the floc formation, it was proposed that a decrease in EPS levels may cause floc deterioration and may be detrimental for the MBR performances. This indicates the existence of an optimum EPS level for which floc structure is maintained without featuring high fouling propensity. Many parameters including gas sparging, substrate composition, and loading rate affect EPS characteristics in the MBR, but SRT probably remains the most significant of them. A functional relationship between specific resistance, mixed liquor volatile suspended solids (MLVSS), TMP, and permeate viscosity, and EPS is believed to exist (Hai and Yamamoto 2011).

3. Operation conditions

1) Aeration, cross-flow velocity

Since the origin of the MBR, bubbling has been defined as the strategy of choice to induce flow circulation and shear stress on the membrane surface. Aeration used in MBR systems has three major roles: providing oxygen to the biomass, maintaining the activated sludge in suspension, and mitigating fouling by constant scouring of the membrane surface. However, an optimum aeration rate, beyond which a further increase has no significant effect on fouling suppression, has been observed on many occasions. It is also important to note that too intense an aeration rate may damage the floc structure reducing their size, and release EPS into the bioreactor, and thereby aggravate fouling (Hai and Yamamoto 2011).

2) Solid retention time

SRT, which greatly controls biomass characteristics, is regarded as the most important operating parameter influencing fouling propensity in MBRs. Considering the advantages of this process over the conventional activated sludge process, the early MBRs were typically run at very long SRTs to minimize excess sludge. But unlike in bench-scale studies employing simpler synthetic feed, the progressive accumulation of non-biodegradable materials (such as hair and lint) in an MBR fed with real sewage definitely leads to clogging of the membrane module.

Operating an MBR at higher SRT leads inevitably to increase of MLSS concentration. The increase in aeration intensity to retain high MLSS levels in suspension and maintain proper oxygenation may not be a sustainable option for the treatment process. In this scenario, the increased shear provided to control fouling could cause biofloc deterioration as well as cell lysis and enhanced EPS secretion, and lead to fatal fouling. On the other hand, 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 explain the low fouling propensity observed for high SRT operation in certain studies. It is likely that there is an optimal SRT, between the high fouling tendency of very low SRT operation and the high viscosity suspension prevalent for very long SRT (Hai and Yamamoto 2011).

3) Unsteady state operation

In practical applications, unsteady state conditions such as variations in operating conditions (flow input/HRT and organic load) and shifts in oxygen supply could occur regularly. The start-up phase can also be considered as unsteady operation and data collected before biomass stabilization (including the period necessary to reach acclimatization) may become relevant in the design of MBRs. Such unsteady state conditions have also been defined as additional factors leading to changes in MBR fouling propensity. For instance, the addition of a spike of acetate in the feed water significantly decreased the filterability of the biomass in an MBR due to the rise in SMP levels resulting from the feed spike (Hai and Yamamoto 2011).

3.2 Membrane fouling reduction methods

The complex interactions between the fouling parameters complicate the perception of MBR fouling and it is therefore crucial to have a complete understanding of the biological, chemical, and physical phenomena occurring in MBRs to assess fouling propensity and mechanisms and thereby formulate mitigation strategies. As membrane fouling increases with increasing flux in all membrane separation processes, the operating flux should be lower than the critical flux. When the operating flux is below the critical flux, particle accumulation in the region of membranes can be effectively prevented. However, due to physicochemical solute-membrane material interactions, the membrane permeability decreases over time, even when MBRs are operated in subcritical (below critical flux) conditions (Hai and Yamamoto 2011). Some approaches, that can be used to reduce membrane fouling during MBR operation, are briefly discussed below.

3.2.1 Membrane modification

1. Optimization of membrane characteristics

Many studies have shown that chemical modifications of membrane surface can efficiently improve antifouling properties. Recent examples comprise (1) increasing membrane hydrophilicity by NH_3 and CO_2 plasma treatments and ultraviolet (UV) irradiation, (2) TiO_2 entrapped membrane, and (3) applying precoating of TiO_2, GAC, ferric hydroxide, polyvinylidene fluoride-graft-polyoxyethylene methacrylated (PVDF-g- POEM), polyvinyl alcohol (PVA), etc. Improved performance in case of precoated membrane has been attributed to the adsorption of soluble organics on the precoat, limiting the direct contact between the organics and the membrane. Self-forming dynamic membrane-coupled bioreactors, utilizing coarse pore-sized substrates and allowing cake and gel layers to deposit on the surface, have been reported to obtain high flux and good removal in certain studies, although stable performance cannot be expected with such a filtration barrier (Hai and Yamamoto 2011).

2. Membrane module design

The membrane module design by optimizing the packing density of hollow fibers or flat sheets, the location of aerators, the orientation of fibers, and diameters of fibers remains another important parameter in the optimization of the MBR operation. In a specially designed module in which air bubbles were confined in close proximity to the hollow fiber (rather than diffusing in the reactor), higher permeability was obtained. Two major design approaches are adopted in case of the commercially available hollow-fiber bundles. One of these approaches relies on partitioning of bundles of fibers, which are fixed at both ends, to secure flow path of air bubbles introduced from the centre of the bundle at the base, thereby leading sludge out of the module. In another approach, bundle of one-end free fibers are allowed to float freely under the scouring action of air bubbles introduced from the core of the bundle to avoid accumulation of sludge. In order to utilize high packing density without encountering severe fouling, a new approach to hollow-fiber module design was explored by. Spacer was introduced within usual hollow-fiber bundles with the aim of minimizing the intrusion of sludge into the module. The little amount of intruded sludge was then backwashed through the bottom end, while the sludge deposited on the surface was effectively cleaned by air scouring. In this way, efficient utilization of cleaning solution and air for backwashing and surface cleaning, respectively, were possible. Recent approaches such as novel fiber sheet membrane, multi-module flat-sheet concept, and vacuum rotation membrane are also noticeable (Hai and Yamamoto 2011).

3.2.2 Optimization of operation conditions.

1. Aeration

As mentioned earlier, bubbling is an established strategy to induce flow circulation and shear stress on the membrane surface. The aeration intensity (air/permeate ratio, m^3/m^3) applied by MBR suppliers may vary between 24 and 50, depending on the membrane configuration (flat sheet vs. hollow fiber) and the MBR tank design (whether the membrane and aerobic zone combined into a single tank or not). However, recent large-scale studies revealed these original ratios are quite conservative. The specific design of bubble size, airflow rate and patterns, and location of aerators have been defined as crucial parameters in fouling mitigation. As the energy involved in providing aeration to the membrane remains a significant cost factor in MBR design. Therefore, the optimization of aeration both from the points of view of fouling mitigation and reducing energy requirement should be done. Recent developments in aeration design include cyclic aeration systems, intermittent aeration, air pulsing , air sparging, improved aerator systems, etc (Hai and Yamamoto 2011).

2. Back flushing

Cleaning protocol has been studied intensively by many researchers where the key parameters of interest are duration and frequency of the cleaning and the back-flush flux. Less frequent, longer back-flushing (600 s filtration/45 s back-flushing) has been found to be more efficient then more frequent but shorter back-flushing (200s filtration/15 s back-flush). Membrane relaxation, which is the most common practice for fouling control, encourages the diffusive back-transport of foulants away from the membrane surface, which is enhanced by air scouring. Relaxation allows longer filtration periods between chemical cleanings, and despite some reports that it may not be a feasible practice for large-scale MBRs, it is widely used in practice. Although intensive research has been done on this subject, membrane fouling in MBRs needs further attention in order to understand the complex interactions among biologically active and constantly changing filtration media, hydrodynamic conditions of the filtration process, and the membrane itself (Radjenović et al. 2008).

3.2.3 Modification of biomass characteristics.

1. Coagulant/flocculant.

Due to back transport and shear induced fouling control mechanisms, large microbial flocs are expected to have a lower impact on membrane fouling. Based on this expectation, studies have explored addition of coagulants such as alum, ferric chloride, zeolite, etc. and have shown permeability enhancement. Pre-treatment of the effluent is also possible and studies based on the pre-coagulation/ sedimentation of effluent before its introduction in the bioreactor revealed the fouling limitation offered by this technique (Hai and Yamamoto 2011).

A cationic polymer-based membrane performance enhancer (MPE 50) has been commercialized by Nalco recently. The interaction between the polymer and the soluble organics was reported as the main mechanism responsible for performance enhancement. The potential impacts of coagulants or adsorbents on biomass community or biomass metabolism need to be taken into account, and the discharge of some chemicals that are used as coagulants or adsorbents might be a potential environmental risk. Such flux enhancers are probably best suited for solving occasional upsets rather than their continuous addition (Hai and Yamamoto 2011).

2. Adsorbent/flux enhancers

Lower fouling propensity is observed in MBR processes when biomass is mixed with adsorbents in that addition of adsorbents into biological treatment systems decreases the level of pollutants, and more particularly organic compounds. In view of saturation of PAC during long term studies, researchers have suggested periodic addition of PAC. Certain studies have proposed pre-flocculation and PAC addition (Hai and Yamamoto 2011).

3.3 Membrane cleaning

Physical cleaning in MBRs is normally achieved either by back-flushing or by relaxation (stopping the permeate flow and continuing to scour the membrane with air bubbles). Physical cleaning is a simple and short method (usually lasting less than 2 min) of fouling suppression which demands no chemicals and generally it is less likely that it will affect the membrane material. The latest published data on the cleaning regime of full-scale plants suggests that presently most of MBR facilities use relaxation rather than back flushing. However, by means of physical cleaning it is not possible to remove all the material deposited on the membrane (Hai and Yamamoto 2011).

Chemical cleaning is a more effective method, which is able to remove more strongly the adsorbed deposits. Chemical cleaning is carried out mostly with sodium hypochlorite and sodium hydroxide for organic deposits removal, or with acidic solutions for removal of lime or other inorganic deposits. Cleaning is performed by soaking the membrane in the cleaning solution or by adding the cleaning agent into the back flush water. Most MBRs employ chemical maintenance cleaning on a weekly basis, which lasts 30–60 min, and recovery cleaning when filtration is no longer sustainable, which occurs once or twice a year (Hai and Yamamoto 2011).

The effectiveness of physical cleaning tends to decrease with operation time as more recalcitrant fouling accumulates on the membrane surface. Therefore, in addition to physical cleaning, different types/intensities of chemical cleaning are applied in practice. A combination of the following types of cleaning is usually applied (Hai and Yamamoto 2011).

Maintenance cleaning with moderate chemical concentration (weekly) is applied to maintain design permeability and it helps to reduce the frequency of intense cleaning. This may be replaced by a more frequent (e.g., on a daily basis) chemically enhanced backwash utilizing mild chemical concentration.
• Intensive (or recovery) chemical cleaning (once or twice a year) is generally carried out when further filtration is no longer sustainable because of an elevated TMP (Hai and Yamamoto 2011).

The MBR suppliers propose their own chemical cleaning recipes, which differ mainly in terms of concentration and methods, and often site-specific protocols are followed. Mainly, sodium hypochlorite (for organic foulants) and citric acid (for inorganics) are used as chemical agents (Hai and Yamamoto 2011).

Some pitfalls of chemical cleaning are worth noting. The detrimental effect of cleaning chemicals on biological performance has been reported. It has also been mentioned that the level of pollutants (measured as total organic carbon) in the permeate rises just after the chemical cleaning step. This raises concern especially in case of MBRs used in the reclamation process trains (e.g., upstream of RO). Chemical cleaning may also shorten the membrane lifetime and disposal of spent chemical agents causes environmental problems (Hai and Yamamoto 2011).

4 Biological performance of membrane bioreactors

The aim of this chapter is to introduce the biological performance of membrane bioreactor.

4.1 Removal of organic matter and suspended solids

Improved COD removal to the avoidance of biomass washout problems commonly encountered in activated sludge process, as well as to complete particulate retention by the membrane. Membrane rejection of a significant amount of soluble organic molecules and colloids makes their removal more effective due to a higher lyses activity in the reactor induced by elevated concentrations of these compounds. Higher sludge ages that are achieved by long SRTs allow more complete mineralization of biodegradable raw water organics, but also an adaptation of microorganisms to less biodegradable compounds. Therefore, biomass can acclimatize to wastewater without being restricted to fast-growing and floc-forming microorganisms (Radjenović et al. 2008).

Better performance of MBR operated at long SRTs can also be explained by the presence of dispersed bacteria that are advantageous in the overall population competition when substrate concentration becomes very low, i.e., at low F/M ratio and high sludge age. Flocs in a bioreactor were found to be smaller, which can explain enhanced mass transfer for both oxygen and carbon, thus enabling a higher removal rate and more adaptability to changes in the influent quality and quantity. In another study it was demonstrated that the flocs were more active and displayed greater species diversity (Radjenović et al. 2008).

The overall capacity of biomass to degrade different carbon substrates does not change significantly at different SRTs, which confirms that MBR is capable of degrading a wide variety of carbon substrates in a similar fashion. This robustness of MBR treatment regarding turbidity and organic matter removals was confirmed in several studies. In another study, in spite of large fluctuations in the influent, COD effluent COD was always low and extremely stable, because upon the addition of organic substrates, biomass responded immediately with increased respiration activity. It is assumed that there is an upper limit for organic loading rate in an MBR under which degradation performance is independent of biomass concentration and organic loading rate (Radjenović et al. 2008).

Concerning turbidity removal, due to a complete retention of particulate matter by the membrane, there are no suspended solids found in the MBR effluent, unlike the effluent of a conventional process. The UF/MF membrane can capture all SS in the reactor because of its fine pore size. Therefore, non-biodegradable organic compounds are removed through filtration of particulates and discharged with the sludge (Radjenović et al. 2008).

4.2 Nitrification/Denitrification and Phosphorus Removal

An irrational use of agricultural fertilizers and pesticides and the discharge of incompletely treated industrial and municipal wastewater results in high nitrogen and phosphate concentration in surface water and groundwater. This enrichment usually leads to an excessive eutrophication of lakes and other water bodies, promoting an excessive growth of certain weedy species. Since both nitrates and phosphates are rate-limiting for the process of eutrophication (extraordinary growth of algae), their removal is of crucial importance for successful wastewater treatment. Nitrate and nitrite-contaminated water supplies are also related to several diseases such as methemoglobinemia occurring in infants, also called 'blue baby disease'. Moreover, these two compounds can induce mutations of deoxyribonucleic acid (DNA), causing gastric cancer (Radjenović et al. 2008).

Biological nitrification is an oxic process of conversion of ammonia to nitrite (NO_2) and then to nitrate (NO_3^-). Following nitrification, nitrogen can be removed from wastewater by reducing nitrate to nitrogen gas (N_2) in the process of anoxic denitrification. Because of the low growth rate and poor cell yield of nitrifying bacteria, nitrification is generally a rate-limiting step in biological nitrogen removal performance. The key requirement for nitrification to occur is that the net rate of accumulation of biomass (and hence the net rate of withdrawal of biomass from the system) is less than the growth rate of nitrifying bacteria. Long SRTs applied in MBR prevent nitrifying bacteria from being washed out from the bioreactor, improving the nitrification capability of the activated sludge. Moreover, nitrifiers are less endangered by faster-growing heterotrophic bacteria, which are better competitors for the ammonia nitrogen (NH_3-N). Many studies have proven that MBR can operate as a high-rate nitrifying technology that can be applied in the nitrification of wastewater containing a high concentration of ammonia nitrogen. Chiemchaisri and Muller found that more than 80% of the influent total Kjeldahl nitrogen (TKN) could be nitrified to NO_3^- in an MBR. On the other hand, the denitrification process requires anoxic conditions in order to occur. To enhance denitrification, usually an anoxic tank is added upstream from the aerated tank. Anoxic conditions can also be introduced by operating MBR in an intermittent aeration mode, even when regarding submerged MBR, which needs permanent bubbling. In the intermittently aerated MBR, ammonium is nitrified mostly to nitrate and most phosphates are removed during the aerobic period (aeration), where the accumulated nitrate is completely denitrified during the anoxic period (non-aeration), and phosphorus (P) is taken up. The net P removal is achieved by wasting sludge after the aerobic period when the biomass contains a high level of polyphosphates (polyP) (Radjenović et al. 2008).

P is found in wastewater as phosphates (orthophosphates, condensed phosphates, organic phosphate fractions), and it can be eliminated either by precipitation and/or adsorption, or by luxury uptake. Only a small amount of phosphorus is used for cell metabolism and growth (1–2% of the total suspended solids (TSS) mass in the mixed liquor). Precipitation and adsorption processes require an appropriate pH, the presence of iron or calcium ions, etc. In WWTPs, luxury uptake of P is accomplished by the introduction of an anaerobic phase in the wastewater treatment line ahead of the aerobic phase and recycling of sludge through the anaerobic and aerobic phase. Exposing mixed liquor to an anaerobic/aerobic sequence selects phosphate accumulating microorganisms (PAOs) due to a competition between PAOs and other aerobic organisms. This competition mechanism is based on a complete anaerobic uptake of the lower fatty acids by the polyp organisms (i.e., PAOs), which assures that in the aerobic phase, no fatty acids are left. The polyP organisms use the stored internal substrate during aerobic conditions while other aerobic organisms are lacking substrate. This process is usually referred to as the EBPR process. EBPR process can be established in MBR treatment unit by operating it in intermittent aeration mode. Moreover, phosphorus removal will be significantly improved in an MBR by a physical retention of PAOs, whose size is typically larger than 0.5 μm. Since an MF membrane (0.2 μm) will act as a physical barrier to retain the PAOs in the reactor, sufficient biomass is provided for the EBPR mechanism to take place (Radjenović et al. 2008).

Intermittently aerated MBR can achieve nitrogen and phosphorus removal by a simultaneous nitrification and denitrification, P-uptake and P-release in the same reactor in accordance with time cycle of aeration and non-aeration. However, even though intermittent aeration was successful in removing nitrogen, P removal was difficult to achieve at a higher level. This is probably due to the inhibition by nitrate. In the anaerobic stage, nitrate reduces phosphate release, and in the aerobic stage it diminishes its uptake. Denitrification has more capability than phosphorus release with respect to the competition of substrate. This is because nitrate will be utilized as a final electron acceptor in the growth of non-poly-P heterotrophs. Thereby, the amount of substrate available for poly-P organisms is reduced and hence the removal of phosphorus is lowered. There are some studies that confirm the ability of poly-P organisms for denitrification, however, not all PAOs can use nitrate as an electron acceptor (Radjenović et al. 2008).

In addition, intermittently aerated MBR showed an unstable nitrogen removal in its application to treat domestic sewage of rural settlements because of incomplete denitrification. Stable nitrification can be maintained as long as the oxygen concentration is high enough. It was found that by lowering the DO concentration, nitrification was significantly inhibited, although it recovered completely after raising the DO concentration to 1 mg/L. Nitrosomonas and Nitrospira-like bacteria were identified as the predominant ammonium and nitrate-oxidizing bacteria (NOB) in an MBR, respectively. Both of these are obligate aerobic and under anoxic conditions are unable to store or utilize their substrate. The absence of oxygen may provoke stress and damage of their metabolism, NOB being the more sensitive (Radjenović et al. 2008).

There is much research on the effect of SRT on MBR performance as far as nitrification/denitrification and phosphorus removals are concerned. There is a decrease in nitrification rate at very low SRT (2 days), supposedly due to a partial loss of nitrifying microorganisms. On the other side, a decreasing trend of nitrifiers was observed when increasing the sludge concentration, i.e., solids retention time. Another study confirmed a negative influence of long SRT on nitrification performance, which was explained by impeded oxygen and substrate transfer owing to an increase in MLSS concentration (Radjenović et al. 2008).

These findings indicate that a compromise should be found between a sufficiently long SRT necessary to prevent the washout of nitrifiers, and a negative influence of too long SRT (decreased mass transfer due to poor aeration, competition with GAOs, etc.). However, over 90% removal of NH_3-N is usually achieved in MBR systems, almost independent on the SRT. The performance of an MBR system which start-up was done without any sludge inoculum was investigated. The system was fed on municipal sewage in order to favour biomass selection based on the imposed operating conditions. Biodegradation of the influent chemical oxygen demand (COD) and complete nitrification were consistently obtained already in the first days of operation. The ammonium oxidation performance over the whole experiment showed a typical nitrification start-up curve with initial N-NO_2 production followed by complete nitrification that occurred only 10 days after the start-up of the plant. As far as HRT is concerned, several studies noted a complete nitrification in an MBR operating with a HRT as low as 2 h (Radjenović et al. 2008).

Other important factors that are to be considered for nitrogen removal are alkalinity, temperature, and organic and nitrogen loads (C/N ratio, i.e., BOD to total nitrogen (TN) ratio). The BOD/TN ratio must be high enough to denitrify the nitrogen to be nitrated. Though this ratio is an important factor to be considered for successful nitrogen removal, it depends upon the components of organic matter that were readily degradable, such as volatile fatty acids. As far as temperature is concerned, it is considered that it has to be maintained below 40 oC to ensure sufficient nitrification. If the temperature is controlled, a nitrification rate usually over 99% can be gained in spite of variations of inflow TN concentration (Radjenović et al. 2008).

Aerated MBR offers two major advantages in the elimination of phosphorus: complete removal of all particles (containing usually up to 0.1 mg of P per mg of total solids (TS), and aeration, which prevents the phosphate release that occurs under anoxic conditions. Furthermore, there is an increasing interest for the application of MBR as a technology for phosphorus recycling, since the P-content of sludge is expected to increase when prolonging SRT (Radjenović et al. 2008).

Much research has confirmed that MBR is a highly viable wastewater treatment technology regarding nitrification-denitrification and phosphorus removal. With optimized design and operating parameters it warrants high effluent quality in terms of ammonia, nitrates, and phosphates present in wastewater. Current European regulation describes guidelines for total phosphorus and nitrogen in treated effluent to 1–2 mg L^{-1} and 10–15 mg L^{-1}, respectively. More stringent regulations are expected to come into force soon in some countries, which fulfilment will require improvements in the existing treatments and the implementation of additional ones (Radjenović et al. 2008).

4.3 Removal of trace organic compounds by a membrane bioreactor

WWTPs treating wastewater from municipalities and industries have been shown as major sources of many environmental pollutants. These pollutants usually originate from synthetic chemicals that have been used widely for industrial, agricultural, and household purposes. Compounds like pharmaceutically active compounds), industrial chemicals, and pesticides are produced worldwide on a 100 000 t scale. After their usage for the intended purpose, a large fraction of these substances will be discharged into the wastewater unchanged or in the form of degradation products that are often hardly eliminable in conventional WWTPs. Depending on the efficiency of the treatment and chemical nature of a compound, they reach WWTP effluents and surface waters in certain concentration. In the worst case, they are present in drinking water, in spite of expensive treatment steps (Radjenović et al. 2008).

Although the exact effect of consistent exposure to trace organic contaminants is still unclear, there is no more doubt that it has significant adverse consequences for public health. For example, antibiotics and their metabolites can significantly increase antibiotic resistance in the population. Synthetic hormones can act as endocrine disruptors by mimicking or blocking hormones and disrupting the body's normal functions. Due to their polarity, they can be eliminated during wastewater treatment only incompletely. Polar poorly degradable compounds were detected in high and comparable concentrations in the effluents of numerous WWTPs all over Europe. A proper wastewater treatment as mandatory in the European Union due to the Urban Wastewater Treatment Directive will not eliminate polar pollutants completely. Therefore, to avoid such contaminants, emissions with WWTP effluents would have to be reduced by their advanced treatment or by avoidance and replacement measures for the respective pollutant. One of the most promising technologies is MBR technology. The potential of MBR to efficiently remove hazardous substances from wastewater is often highlighted. Besides the fact that there is a physical retention of all the molecules larger than the molecular weight cut-off (MWCO) of the membrane, hydrophobic substances also tend to accumulate onto the sludge and therefore they are removed from the effluent. Furthermore, as all the bacteria are held back, there are better adapted to mineralizing of micro pollutants present in the reactor (Radjenović et al. 2008).

4.4 Removal of hormones

Estrogenic substances have been identified and quantified in a wide variety of environments associated with industrial and municipal effluents, as well as urban and agricultural runoffs. Negative adverse health effects on aquatic organisms which could be attributed to endocrine disrupting compounds (EDCs) are reported by several authors. EDCs are substances that interfere with the hormone system of animals and human beings. When absorbed into the body, they either mimic or block hormones and disrupt the body's normal functions. This disruption can happen through altering normal hormone levels, halting or stimulating their production, or changing the way they travel through the body, thus affecting the functions that these hormones control. According to a description by the European Commission, an endocrine disruptor is "an exogenous substance or mixture that alters function(s) of endocrine system and consequently causes adverse health effects in an intact organism, or its progeny, or (sub) populations". Two different classes of substances causing endocrine disruption can be identified: natural substances, including natural sexual hormones (estrogens, progesterone and testosterone) and phytoestrogens (chemicals produced by plants that act like estrogens in animal cells and bodies), xenobiotic substances, including synthetic hormones as the contraceptive 17α- ethinylestradiole as well as man-made chemicals and their by-products (e.g., pesticides, cleaning agents, flame retardants, etc.) (Radjenović et al. 2008).

Membrane bioreactor technology is a possibility to enhance the removal of EDCs.Factors like high sludge ages and low organic loads could not yet be correlated to improved degradation capacity, although biomass compositions might influence hormone removal in MBRs. Generally, the micro and ultrafiltration membranes do not display a barrier effect to hormones, but compared to conventional secondary and tertiary systems, a high removal can be expected due to full particle retention promoting the adsorption onto the sludge flocks. Although a certain estrogenic activity could be expected in MBR effluent (due to the cleavage of estrogen conjugates, or dissolution of particles due to digestion process that may release estrogens by desorption), membrane technology represents a good combination of different mechanisms for removal of endocrines. On one side, smaller floc sizes and higher sludge activity enhance biodegradation, and on the other particle size exclusion enables retention of the adsorbed compounds (Radjenović et al. 2008).

5 Operation cost of membrane bioreactor

Although membrane bioreactor can provide much better effluent water quality than conventional bioprocesses, the operation of membrane bioreactor generally requires higher energy cost. This chapter is to introduce the cost of the operation of membrane bioreactor.

5.1 Energy cost

A cost comparison in 2007 showed that MBR was 15% more expensive (15 million litters a day), whereas a study by Zenon in 2003 gave MBR 5% lower costs. The differences were due to the design fluxes assumed and the capital charge rate for the project. Neither study allocated a cost advantage from the reduced footprint, which could typically translate to a treated water cost saving of up to 5%. It is interesting to evaluate the development in cost estimates over the past several years. A cost comparison was made for two wastewater treatment plants, with capacities of 2350 and 37 500 p.e. Based on a assumptions (e.g., a membrane lifetime of 7 years), it was conclude that depending on the design capacity (i.e., 2 times DWF to be treated) MBR is competitive with conventional treatment up to a treatment capacity of 12 000 m^3 d^{-1} (Hai and Yamamoto 2011).

Based on a survey of conventional wastewater-treatment facilities in the US, the energy usage range was 0.32–0.66 kW $h^{-1}m^{-3}$. Long-term monitoring of wastewater-treatment systems has shown usages as low as 0.15 kW $h^{-1}m^{-3}$ for activated sludge, increasing to 0.25 kW $h^{-1}m^{-3}$ if a biological aerated filter (BAF) stage is included (Hai and Yamamoto 2011).

Membrane filtration after conventional treatment is estimated to add 0.1–0.2 kW $h^{-1}m^{-3}$ to the energy. Table 5.1 lists typical energy-use rates of different biological based treatment combinations .Experience in large-scale commercial MBRs shows an energy usage of around 1.0 kW $h^{-1}m^{-3}$, although smaller-scale facilities typically operate at 1.2–1.5 kW h^{-1} m^{-3} or higher (Hai and Yamamoto 2011) .

Treatment option	Energy use (kW h^{-1} m^{-3})
CAS	0.15
CAS-BAF	0.25
CAS-MF/UF	0.35–0.5
MBR	0.75–1.5[a]

Power consumption range for large to smaller scale plants.

Table 5.1 comparative typical energy consumption by different treatment option (Hai and Yamamoto 2011)

5.2 Aeration demand

Energy usage for membrane aeration is a significant operating cost for any membrane bioreactor facility. It was calculated the total variable operational cost of MBR by summing the decreasing sludge-treatment cost and increasing aeration cost (see Figure 5.1). Since minimized sludge production implies maximized aeration cost, and vice versa, they considered the existence of an optimum point between these two extreme cases, where the total operational cost is minimized. They concluded that for reasonable ranges of HRT and MLSS sludge treatment cost overwhelms aeration cost, so the most adequate strategy for MBR cost reduction would be maintenance of low sludge production conditions. High SRT in an MBR means a high MLSS concentration and low F/M ratio, which enables application of short HRT. However, sludge production is obviously inversely proportional to HRT when MLSS is mixed. The shortest HRT and the minimum sludge production cannot be achieved simultaneously (Radjenović et al. 2008).

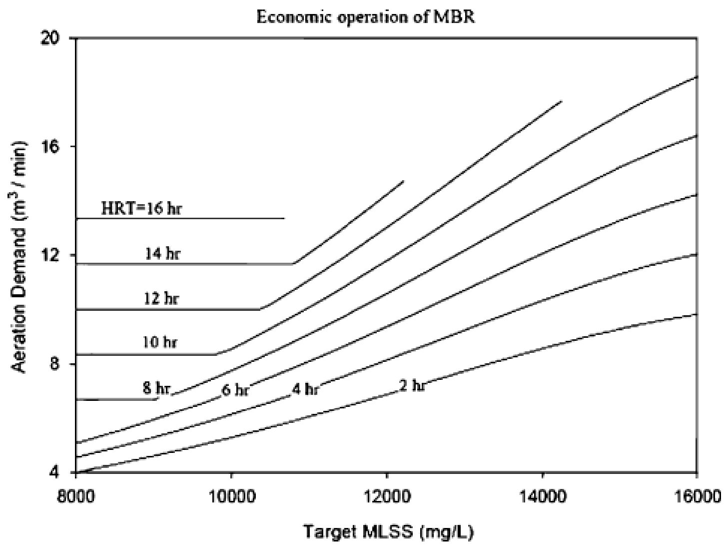

Figure 5.1 aeration demands for biodegradation of organic matters as a function of target MLSS and HRT. Flow rate and COD of influent were 1000 m3/d and 400 mg/L, respectively (Radjenović et al. 2008).

6 Design of membrane bioreactor

The aim of this chapter is to introduce design considerations of MBR, which is important to ensure the successful operation of membrane bioreactors.

6.1 Pre-treatment

6.1.1 Screen

MBR suppliers normally specify fine screening requirement of <2 mm mesh or hole opening (<1 mm preferred), while side stream MBRs will typically have tighter requirement for fine screening. Fine screens are sized for peak flow with one screen out of service to prevent overflow or bypass of unscreened wastewater. Fine screens in many different configurations are available, each uniquely fitting a particular need and application. Typical fine screen configurations include rotating brush screens, internally-fed rotary drum screens, in-channel rotary drum screens and traveling band screens (AMTA 2007). In the case of hollow fiber membranes, use of 0.8 mm to 1.5 mm fine screen upstream of membranes is recommended to protect the membranes from hair and other stringy materials that can result in excessive cleaning frequencies. A fine screen of 2–3 mm is usually employed for flat sheet membrane systems (Radjenović et al. 2008). Hollow fiber membranes have a tendency for aggregates of hair and other debris to collect at the top of the membrane elements. These aggregates usually cannot be significantly removed by back-flushing. Flat sheet modules are somewhat less prone to such clogging, but they too need a feed-water pre-treatment, though with coarser screens (Radjenović et al. 2008).

In the operation of a MBR, even if a 1 mm screen was applied, two severe problems occurred (Melin et al. 2006):

1. The wedge wire sieve (slots) showed quite low removal efficiencies. Later it was replaced by the much better performing woven mesh type sieve. The type of sieve installed is very important regarding the total screening removal, especially towards hair and fibres.
2. After replacement of the drum, severe and frequent clogging of the mesh type sieve occurred, caused by an inadequate automated cleaning system. Frequent manual high pressure cleaning by the operators was necessary to guarantee a continuous operation. Note that the investment in a 1 mm pre-treatment drum screen is of little use if the bioreactor is open and surrounded by trees as it is the case in Schilde. After the start-up it was decided to cover the bioreactor completely in order to prevent leafs, needles and other debris from falling into the bioreactor.

6.1.2 Oil and grease

Oil and grease in the concentrations typically found in municipal sewage have little or no impact on the operation of an MBR. However, free oil and grease needs to be removed as this can prematurely foul membranes (AMTA 2007).

6.1.3 Equalization Tank

As membranes are costly and the membrane surface area has to be designed for the maximum flow, it has to be checked whether hydraulic peaks can be levelled out. Equalization tanks and/or an aeration tank operation mode with alternating water levels are measures to be considered in the design phase. At municipal plants, the maximum storm water flow at low temperatures has to be taken in account (Cornel and Krause 2008).

6.1.4 Inorganic Chemicals

High calcium concentrations (>200 mg/L Ca^{2+}) may cause scaling problems, depending on aeration and pH. With the high specific aeration at the membrane surface of submerged membranes, CO_2 is stripped out, which leads to an increase in the pH and, as a consequence, may cause $CaCO_3$ precipitation at or in the membrane; pH control helps avoid precipitation as has been demonstrated at an industrial site in Germany. In tubular membrane systems without scour aeration, the pressure drop across the module and/or the membrane can shift the solubility equilibrium and cause $CaCO_3$ precipitation. Typically, only industrial applications are affected by calcium precipitation. The membranes can be cleaned with (weak) acids (Cornel and Krause 2008).

6.2 Factors influencing performance

6.2.1 Membrane selection and applied flux

An MBR membrane needs to be mechanically robust, chemically resistant to high Cl_2 concentrations used in cleaning, and non-biodegradable. Clean-water permeability is not as important in an MBR as in membrane-filtration applications, since the membrane transport properties are strongly influenced by the accumulation of foulant particles at the membrane surface. However, process flux in treating a wastewater feed is important since it directly affects capital cost, due to its effect on membrane area and footprint, and operating costs due to the effect of membrane area on chemical and air use. Most MBRs operate at an average flux rate between 12.5 and 25 L $m^{-2} h^{-1}$. The key flux rates that determine the number of membranes required are associated with the peak flow rates. For plants with peaking factors of less than two, an MBR can handle the plant flow variation without having a significantly impact on the average design flux rate. Otherwise, equalization needs to be provided with either a separate tank at the head of the facility or within the aeration basin, allowing side water depth variations during peak flow (Hai and Yamamoto 2011).

The leading MBR providers propose several MBR designs. In each case, the process proposed is very specific. Not only are the membrane material and configuration used different, but the operating conditions, cleaning protocols, and reactor designs also change from one company to another. For example, the flat-sheet membrane provided by Kubota does not require backwash operation, while hollow-fiber membranes have been especially designed to hydraulically backwash the membrane on a given frequency (Hai and Yamamoto 2011).

In the 1990s, PVDF became established in MBRs through the reinforced capillary fiber in Zenon's ZW 500 module. PVDF has impressive performance in terms of strength and flexibility, but is significantly more expensive as a polymer. Nevertheless, PVDF membranes can achieve substantially higher flux, thereby overcoming price disadvantage. Recently, MRE also developed a PVDF-based membrane system. This membrane, designated as SADF, promises to be very competitive in both capital and operating costs, and despite it having a lower packing density than the PE product, it operates at much higher flux. With several companies now offering PVDF products in both capillary and flat-sheet formats, this is the dominant membrane polymer in the MBR market (Hai and Yamamoto 2011).

A significantly used membrane polymer in MBR is a reinforced PES, used by Koch-Puron. Although PES is an important polymer in water treatment, in wastewater applications, its lack of flexibility limits the possibility of using air scour. Reinforcing the capillary does allow air scour, but at the expense of permeability. The Puron product uses reinforced PES rather than the PVDF, favoured by its rivals. However, its main distinguishing feature is that the membrane fibers are potted at only one end. This overcomes the problem of fouling below the potting interface by hairs and fibers, which is a problem for the other hollow-fiber technologies (Hai and Yamamoto 2011).

Norit is the one major MBR Company that offers a system based on a sidestream format with tubular membranes rather than a submerged format. Cross flow is only used for small scale applications, with feeds that are difficult to treat, whereas airlift is cost effective for larger-scale municipal applications (Hai and Yamamoto 2011).

6.2.2 Sludge retention time

In the past, most MBR systems were designed with extremely long SRTs, of the order of 30–70 days, and very few were operated at less than about 20 days. Two reasons prompted such practice: (1) the drive to minimize sludge production or eliminate it all together and (2) the concern over the reduced flux resulting from short SRT operation, presumably due to the fouling effect of extracellular excretions from younger sludge. Currently, the selection of SRT is based more on the treatment requirements and SRTs as low as 8–10 days can now be contemplated (Hai and Yamamoto 2011).

6.2.3 Mixed liquor suspended solids concentration

From the point of view of bioreactor volume reduction and minimization of excess sludge, submerged MBR systems have been typically operated with MLSS concentrations of more than 12 000 mg L^{-1}, and often in the range of 20 000 mg L^{-1}. Hence, they offer greater flexibility in the selection of the design SRT. However, excessively high MLSS may render the aeration system ineffective and reduce membrane flux. A trade-off, therefore, comes into play. Current design practice is to assume the MLSS to be closer to 10 000 mg L^{-1} to ensure adequate oxygen transfer and to allow for higher membrane flux. With larger systems, it is more cost effective to reduce the design MLSS because of the high relative cost of membranes when compared to the cost of additional tank volume (Hai and Yamamoto 2011).

It is observed that the coefficient for oxygen mass transfer for clean water and waste water (a-value) decreases strongly with an increase of the MLSS in the aeration basin which leads at the same time to a higher process air demand and a higher energy consumption for aeration (Alfred and Möbius 2009). Our recommendation for MLSS design according to the available information is a moderate mixed liquor concentration of 10–12 g/l with a a-value of 0.6 for calculation of the oxygen requirement in the bioreactor. However, the results come from tests with municipal wastewater and have to be verified with papermill effluent (Alfred and Möbius 2009).

6.2.4 Aeration

At high MLSS concentrations, the demand for oxygen can be significant. In some cases, the demand can exceed the volumetric capacity of typical oxygenation systems. The oxygen transfer capacity of the aeration system must also be carefully analysed. Submerged membranes are typically provided with shallow coarse bubble air to agitate the membranes as a means to control fouling. Such aeration provides some oxygenation, but at low efficiency. In compact systems, fine bubble aeration may be placed at greater depth below the membrane aeration; however, the combined efficiency and the bubble coalescing effects require further consideration during design. The lower operating cost obtained with the submerged configuration along with the steady decrease in the membrane cost encouraged an exponential increase in MBR plant installations from the mid-1990s onward. Since then, further improvements in the MBR design and operation have been introduced and incorporated into larger plants. The key steps in the recent MBR development are summarized below (Hai and Yamamoto 2011):

1. The acceptance of modest fluxes (25% or less of those in the first generation), and the idea of using two-phase bubbly flow to control fouling.
2. While early MBRs were operated at SRTs as high as 100 days with MLSS up to 30 g L^{-1}, the recent trend is to apply a lower SRT (around 10–20 days), resulting in more manageable MLSS levels (10–15 g L^{-1}).
3. Thanks to these new operating conditions, the fouling propensity in the MBR has tended to decrease and overall maintenance has been simplified, as less-frequent membrane cleaning is necessary.

It is important to install enough aeration capacity in the first cascade (s) to avoid an oxygen deficit and minimise the risk (of a massive growth) of filamentous microorganism (bulking sludge) which might affect the operation of the membrane stage. This has to be mentioned because of the higher sludge content and high oxygen consumption compared to classical activated sludge systems (Alfred and Möbius 2009).

The aeration basin (and membrane containers in the case of submerged membranes) should be fit with spray water system for foam control. However, defoamer should only be used when required to avoid costs and the risk of an accumulation of substances which might affect the membranes. In addition the possibility of installing a skim device on the aeration surface for the evacuation of foam out of the system can be proved if the effluent tends to have massive foam development and the foam control requires too much de-foamer chemicals (Alfred and Möbius 2009).

6.2.5 Flow variation

Many small-scale MBR plants are purposely base loaded in order to optimize the membrane design, e.g. the peak day flow is limited. While this design approach has worked well for satellite facilities, most large scale plants are end of the line facilities and the plant must accommodate the flow fluctuation from both diurnal flows and storm flows. Because the membrane sizing is hydraulically driven, alternatives to increasing the number of membranes should be considered if the peak flow is more than twice the average flow as the economical upper flow limit for membranes in most MBRs is approximately 1.5 to 2 times the average flow rate. Designing membranes to accommodate higher peak flows typically results in flux rates at the average flow which are below the optimized point and significantly increases equipment cost. Operation and management costs are also increased due to increased air scour requirements. To allow membrane optimization, equalization should be considered.

There are two options for equalization: external and internal. External equalization consists of a separate basin ahead of the MBR basin. The external equalization basin easily satisfies design requirements, but if the driver behind the selection of an MBR is footprint, an additional facility may be difficult to incorporate on space constrained sites. Internal equalization, e.g. side water depth variation within the aeration tanks, can be used if the variation isn't too large. Most plants will be limited to approximately 2 to 4 feet of variation before the aeration blowers are significantly impacted unless less efficient positive displacement blowers are used. Typically internal equalization is best suited for dampening the diurnal pattern because the level variation required to effectively manage storm flows tends to be significant and could adversely impact the blower design. For some facilities a combination of both external and internal flow equalization provides a cost effective solution with the external equalization basin used for off-line storage to handle storm events and internal equalization used to handle the daily diurnal variation. If the expected rainwater flow / dry weather flow is smaller than three, the membrane surface area can be sized based on the average flow rate. In constant, the cross flow MBR system will have to be sized based on the peak flow rate, resulting in additional investment costs (van Haandel and van der Lubbe 2007).

Table 6.1 shows the operation conditions for submerged MBR, which is useful to design an MBR.

Parameter	value
Flux	
Instantaneous, L/m^2.h	25–35
Sustainable in long term operation, L/m^2.h	15–30
Transmembrane pressure, kpa	20
Biomass concentration, g MLSS/L	5–25
Solids retention time ,d	>20
Sludge production, kg SS/kg COD.d	<0.25
Hydraulic retention time, h	1–9
F/M, kg COD/kg MLSS.d	0.2
Volumetric load, kd COD/m^3.d	Up to 20
Air flow rate, N m^3/h per module	8–12
Operation temperature °C	10–35
Operation pH	7–7.5
Backwash frequency, min	5–16
Backwash duration, s	15–30
Energy consumption for filtration, kWh/m^3	0.2–0.4
For membrane aeration,%	80–90
Pumping for permeate extraction,%	10–20

Table 6.1 Operation conditions for submerged MBR (Melin et al. 2006)

7 Long term experience on membrane bioreactors operation

This chapter shows the start-up of a MBR and its cleaning in practice

7.1 Star-up of a MBR

A MBR pilot system was installed at the premises of the Sanitary Engineering Research and Development Centre. Primary treated municipal wastewater after passing through 1 mm screens was fed to a 2 m^3 equalisation tank and then to a 250 L pilot scales MBR where the membrane module was immersed. Permeate was filtered directly from the aeration tank. Coarse bubble aeration was supplied at a constant rate of 4.0 m^3/h to minimise membrane fouling, while fine bubble aeration was supplied through air diffusers in order to maintain the DO level higher than 2 mg/L. The high DO concentration ensured that the DO fluctuation in the bioreactor had a minimal effect on membrane fouling. The filtration pattern took place at 10-min cycles during which 9.5 min of filtration was followed by 0.5 min of backwash. The MBR was left to develop its own biomass for a period of 3 months (start-up period) and was subsequently operated under steady-state conditions. The system was left for a period of at least 2×SRT to attain the required steady state conditions. Four experimental periods were conducted and each period was characterised by a different solids retention time (SRT), namely 10, 15, 20 and 33 days. The hydraulic retention time (HRT), filtration flux, backwash flux and membrane aeration remained constant under all operating conditions. The hollow fibre membrane module was supplied by GE Water and Process Technologies. It was made of polyvinylidene fluoride (PVDF) and had a nominal pore size of 0.04 mm. Table 7.1 summarises the operating characteristics (Malamis et al. 2011).

Parameter	1st Period	2nd period	3rd period	4th period
SRT, h	10	15	20	33
HRT, h	11.2	11.1	10.7	11.3
Operation volume, L	190	210	190	210
Excess sludge, L/d	19.0	14.0	9.5	6.4
MLSS, g/L	2.96	5.08	6.27	11.40
	2.06–3.78	4.90–5.34	4.67–7.55	9.66–12.82
MLVSS, g/L	2.53	4.27	5.19	9.13
	1.79–3.64	4.13–4.41	3.86–6.21	7.70–10.38
DO, mg/L	>2.0	>2.0	>2.0	>2.0
pH	7.15	7.09	7.10	7.15
	6.77–7.86	6.89–7.38	6.76–7.51	6.92–7.42
Membrane aeration (m^3/h)	4	4	4	4
Filtration flux ($L/m^2.h$)	22.3	22.3	22.3	22.3
Backflush flux, $L/m^2.h$	21.5	21.5	21.5	21.5
F/M	0.475	0.298	0.252	0.161

Table 7.1 operating characteristics of MBR (Malamis et al. 2011)

The MBR pilot system operated at SRT = 10, 15, 20 and 33 days. Figure 7.1 shows the variation of MLSS and MLVSS in the bioreactor. Two major sludge loss events were experienced, one on day 358 when the system was operating at SRT = 20 d and one on day 562 when the system was operating with SRT = 33 d. In both cases the system was left to regain its former steady state conditions, while no experiments were conducted. The increase of SRT for constant HRT resulted in an increase of the MLSS concentration from 3 g/L at SRT = 10 days, to 11–13 g/L at SRT = 33 days (Malamis et al. 2011).

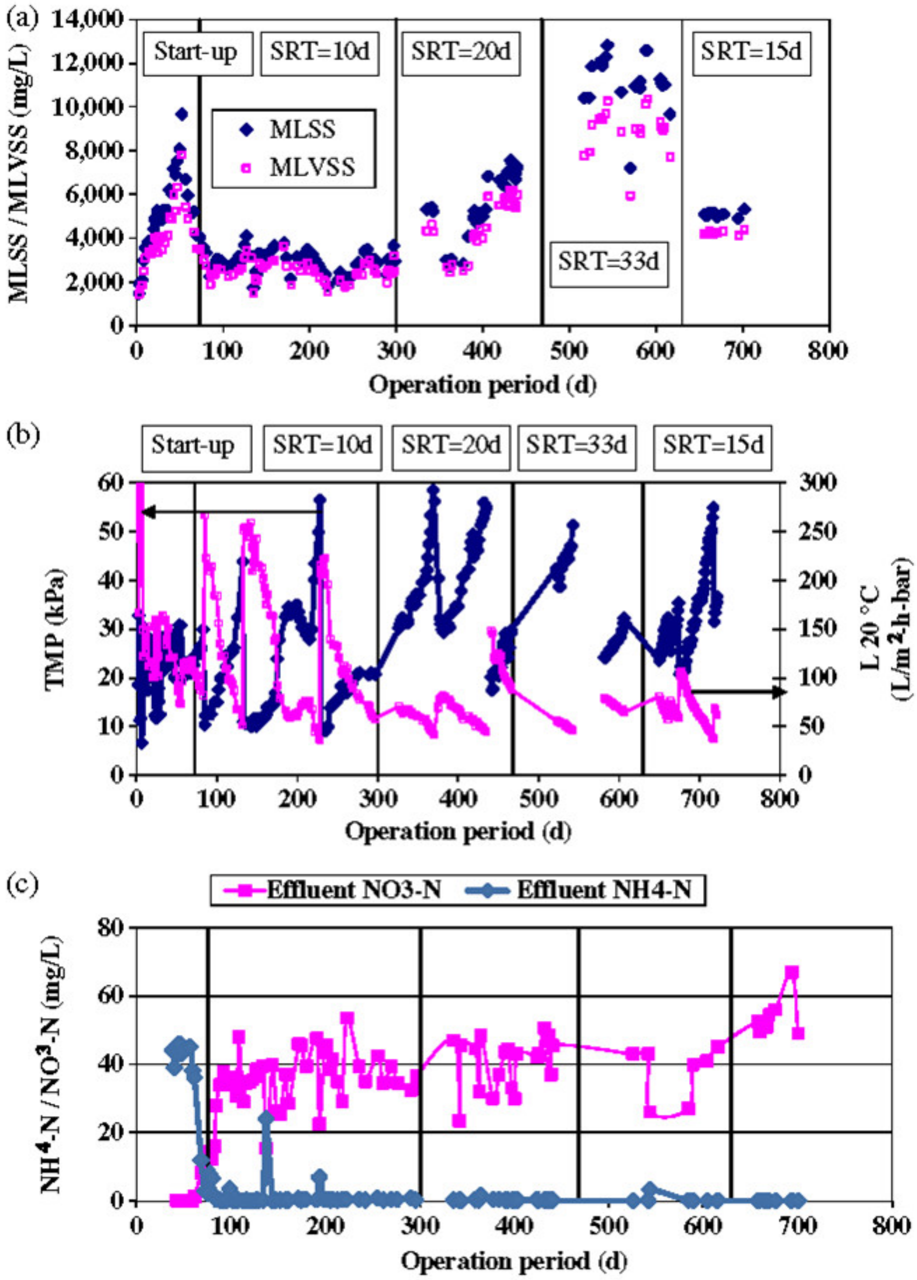

Figure 7.1 variation of (a) MLSS and MLVSS, (b) transmembrane pressure (TMP) and membrane permeability,(c) treated effluent NH_4-N and NO_3-N with time (Malamis et al. 2011).

Figure 7.1 shows the transmembrane pressure (TMP) and the membrane permeability fluctuation during the whole operation period of the system. The gradual increase of TMP and the corresponding decrease of permeability occurred due to gradual membrane fouling taking place. The rapid decrease of TMP and the resulting increase in membrane permeability occurred due to chemical cleanings that were performed to restore the membrane permeability. The increase of SRT from 10 to 20 days resulted in a reduction of the rate of membrane permeability decrease from 2.56 L/m^2.h-bar-d to 0.60 L/m^2.h-bar-d (80% reduction), while the further increase of SRT to 33 days did not significantly impact on the fouling rate, which was 0.53 L/m^2-h-bar-d. The system performance stabilised at SRT Z20 d, which is associated with lower concentration of extracellular polymeric substances (Malamis et al. 2011).

The effluent was characterised by very low COD for all operating conditions. The average effluent COD was lower than 30 mg/L for all operating conditions and was lower than 20 mg/L for SRT Z 20 days, while the percentage COD removal was higher than 95%. Organics removal was significant even during the first 10–15 days of system operation, despite the fact that the biomass was not fully developed. This is attributed to the effective filtration process which resulted in the removal of suspended solids and colloidal substances, since the membrane MWCO was 200 kDa. Full nitrification (100%) was observed for all the examined operating conditions. This was the case even at relatively low temperature (12–13 ^{O}C) and high MLSS concentrations (11–13 g/L). After the start-up period the NH_4-N concentration in the bioreactor and in the final effluent was negligible and high nitrate levels were always observed (Figure 7.1(c)) (Malamis et al. 2011).

7.2 Membrane fouling and cleaning in a long term operation

Figure. 7.2 shows the evolution of TMP and flux in the long-term operation of a MBR. In Run 1 of 200 d, TMP and its rising rate were in the low range of 4.6–17.6 kPa and 0.15–2.09 kPa/d respectively, indicating steady filtration was successfully achieved in the long-term operation. This depressed TMP evolution was mainly attributed to both synergistic effectiveness of chemical cleaning-in-place (CIP), with a TMP controlling mode, and sub-critical flux operation. Sub-critical flux operation could reduce the cake layer caused by sludge flocs, resulting in slow TMP rise (such as the average of 0.79 kPa/d in Run 1 in this study) with time. After every chemical CIP, TMP decreased almost down to the original value (5–7 kPa), indicating that chemical CIP could effectively remove the fouling in terms of membrane pore blockage and gel layer caused by colloids and soluble organic substances. Cleaning intervals were from 2 to 3 weeks to nearly 3 months due to fluctuation of sludge characteristics. In Run 2, TMP and its rising rate were also in the low range of 5.1–18.2 kPa and 0.50–1.48 kPa/d respectively under effective membrane fouling control by sub-critical flux operation and chemical CIP, but with a time controlling mode. Although TMP did not decrease to the original value after weekly cleaning with low concentration NaClO and rose, to some extent, in one month, TMP decreased to the original value after monthly cleaning with high concentration NaClO. From the low average TMP and its rising rate in Run 1 and Run 2 with the similar flux, steady filtration in the long-term operation could be achieved under sub-critical flux operation and chemical CIP (with TMP or time controlling mode) (Wei et al. 2011).

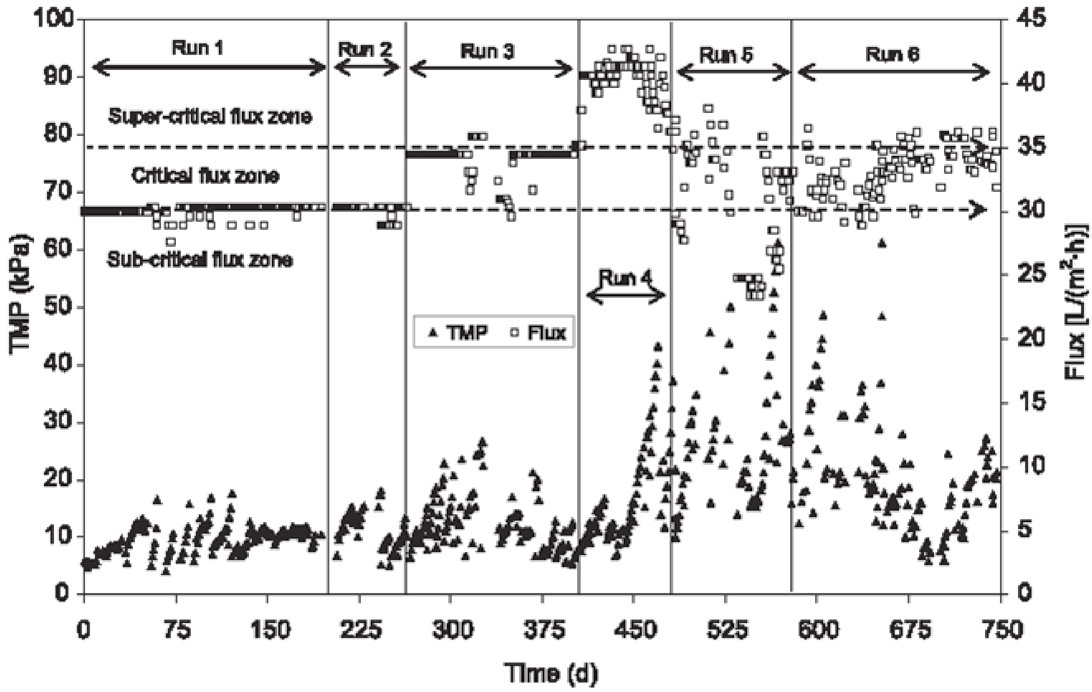

Figure 7.2 TMP and flux changes of the pilot SMBR during the long term operation (Wei et al. 2011)

Compared with Run 1 and Run 2, TMP (5.3–26.6 kPa) and its rising rate (0.39–3.05 kPa/d) between two cleanings in Run 3 were higher due to critical flux operation. Although filtration stability was not as good as in Run 1 and Run 2, TMP could be kept below 30 kPa in Run 3, showing a possible steady long-term filtration under critical flux operation and chemical CIP with combined TMP and time controlling modes. Especially TMP increase in the second half of Run 3 with flux of 29.7–35.9 L/(m^2 h) was comparable to that in initial Run 1 with flux of 27.6–30.3 L/(m^2 h), indicating the possible seasonal effects on membrane fouling. Both initial Run 1 and the second half of Run 3 were in summer with good sludge filterability. High temperature not only decreased the sludge viscosity but also affected microbial activity especially under low SRT, thus resulting in less fouling than low temperature. In addition, the relative steady filtration in Run 3 also showed that the criterion for identifying critical flux in this study was a little conservative (Wei et al. 2011).

In Run 4, under super-critical flux operation and chemical CIP with time controlling mode, both TMP and its rising rate were in the higher range of 6.7–43.2 kPa and 0.61–4.69 kPa/ d compared to Run 1–3. Especially at the second month of Run 4, TMP rising rate was higher than 3 kPa/d, showing characteristics of cake layer fouling. After weekly cleaning with low concentration NaClO, virtually no TMP decrease was observed. Although TMP could be depressed after monthly cleaning with high concentration NaClO, TMP rose up to higher than 40 kPa at the end of Run 4 and flux decreased with increased TMP due to the limit of suction pump. This indicated that steady filtration under super-critical flux operation could not be achieved even adopting chemical CIP (Wei et al. 2011).

In Run 5, variable flux operation was conducted in order to restore membrane permeability. Although enhanced chemical CIP (NaClO and HCl or NaOH) was performed, low TMP (less than 20 kPa) was only adequate under low flux (about 25 L/(m^2 h)) due to accumulative fouling in Run 4. In addition, the relative poor filterability of activated sludge under winter temperature (10–15 ^{O}C) in Run 5 might be another factor for low permeability recovery (Wei et al. 2011).

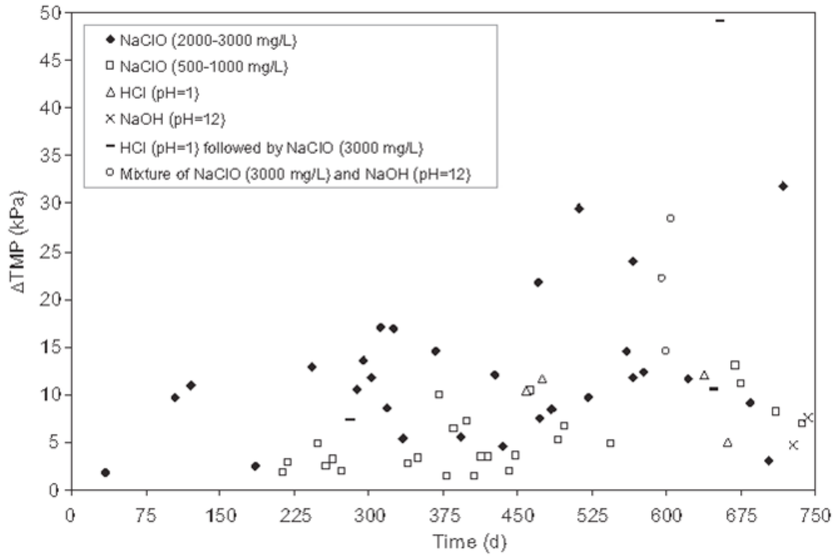

Figure 7.3 TMP decrease after chemical CIP with different cleaning reagents.

In Run 6 under critical flux operation, TMP showed a gradual decrease followed by a gradual increase due to enhanced chemical CIP and fluctuation of sludge characteristics. In the second half, TMP was obviously lower than in Run 4 and Run 5 and was close to that in Run 3, indicating membrane permeability could be recovered from serious fouling, caused by super-critical flux operation in Run 4, to some extent by chemical CIP (Wei et al. 2011).

From all runs, long-term steady filtration was successfully achieved under sub-critical flux operation (30 L/(m^2 h)) and chemical CIP with both cleaning modes. Critical flux operation (30–35 L/(m^2 h)) appeared to be feasible for achieving long-term steady filtration under chemical CIP with combined TMP and time controlling mode because this combined cleaning mode could provide the more intensive protection for membrane. However super-critical flux operation (35–42 L/(m^2 h)) was not possible for achieving long-term steady filtration even adopting chemical CIP. The results from the long-term operation of the pilot-scale SMBR in this study also demonstrated that the critical flux should be a key parameter for realizing long-term steady filtration for real SMBR application and the method for critical flux measurement through short-term constant-flux filtration test used in this study was feasible for long-term operation (Wei et al. 2011).

Reagent	TMP decrease after cleaning (kPa)
NaClO (2000–3000 mg/L)	1.8–31.9 (12.7), n=31
NaClO (500–1000 mg/L)	1.4–13.0 (5.1), n=25
HCl (pH = 1)	5.0–12.1 (9.8), n=4
NaOH (pH = 12)	4.7–7.6 (6.2), n=2
HCl (pH = 1) followed by NaClO (3000 mg/L)	7.4–49.2 (22.4), n=3
Mixture of NaClO (3000 mg/L) and NaOH (pH = 12)	14.6–28.4 (21.7), n=3

Value in parentheses are the average values, n is the number of cleaning.

Table 7.2 cleaning performance of different reagents (Wei et al. 2011)

Figure.7.3 and Table 7.2 show chemical CIP performance of various cleaning reagents in terms of TMP decrease after cleaning under the same flux, i.e. the simple index for permeability recovery. From Table 7.2, average TMP decrease after cleaning for high concentration NaClO (2000-3000 mg/L), low concentration NaClO (500–1000 mg/L), HCl (pH=1) and NaOH (pH = 12) was 12.7 kPa, 5.1 kPa, 9.8 kPa and 6.2 kPa respectively, indicating high concentration NaClO was the best among all of the reagents. It was also found that high concentration NaClO (2000 mg/L) was better than other chemicals (such as H_2O_2, NaOH, HCl, citric acid and enzymes) used in their study. The recovery of membrane permeability by NaClO cleaning has also been reported in many publications. HCl and NaOH showed results similar to that of low concentration NaClO. The combination of NaClO and HCl or NaOH appeared to be better than NaClO only, especially when serious fouling occurred. In addition, for the performance of each cleaning in terms of TMP decrease, it depended not only on cleaning reagent but also on fouling condition before cleaning. Under the same cleaning reagent, generally the higher TMP before cleaning was, the higher TMP decrease after cleaning was. Serious fouling occurred at the second half of Run 4, Run 5 and initial Run 6, was partially the reason for the large fluctuation of cleaning performance. Finally, it should be noted that HCl, NaOH, combined NaClO and HCl, and combined NaClO and NaOH cleaning were performed only 4, 2, 3 and 3 times, respectively, in this study. Therefore, further investigation is necessary to optimize CIP strategies for MBR (Wei et al. 2011).

During the long-term (750 d) operation of a pilot-scale submerged membrane bioreactor for municipal wastewater treatment, steady filtration for at least 265 d under high flux (30 L/(m^2 h)) was successfully achieved due to effective membrane fouling control by sub-critical flux operation and chemical CIP with NaClO. Sub-critical flux operation prevented rapid fouling caused by cake layer formation. NaClO CIP, in both a TMP controlling mode (cleaning with high concentration NaClO (2000–3000 mg/L) was performed when TMP rose to 15 kPa) and a time controlling mode (cleanings were performed weekly and monthly respectively with low concentration NaClO (500–1000 mg/L) and high concentration NaClO (3000 mg/L)), effectively controlled slow fouling caused by pore blockage and gel layer. Critical flux operation (30–35 L/ (m^2.h)) under chemical CIP with combined TMP and time controlling mode achieved steady filtration up to 140 d and appeared promising for achieving longer steady filtration although it was less stable than sub-critical flux operation. However super-critical flux operation (35–42 L/(m^2.h)) was not attainable for achieving long-term steady filtration even adopting chemical CIP. In addition, the results also demonstrated that the method for critical flux measurement through short-term constant-flux filtration test used in this study was feasible for long-term operation (Wei et al. 2011).

From macroscopic cleaning performance, high concentration NaClO was the best. HCl (pH = 1) and NaOH (pH = 12) showed results similar to low concentration NaClO. The combination of NaClO and HCl or NaOH appeared to be better than NaClO alone, especially when serious fouling occurred. Microscopic analysis on membrane fibers before and after high concentration NaClO CIP was conducted in order to investigate the mechanism of NaClO CIP for removing membrane fouling. SEM and Atomic force microscope images showed that NaClO cleaning could effectively remove the gel layer, the dominant fouling under sub-critical flux operation. Porosity measurements indicated that NaClO cleaning could partially remove pore blockage fouling. Fourier transform infrared spectroscopy and EDS analyses demonstrated that protein-like macromolecular organics and inorganics were the important components of the fouling layer. The analysis of effluent quality before and after NaClO CIP showed no obvious negative effect on effluent quality (Wei et al. 2011).

8 Applications of membrane bioreactors

In this chapter, the world wide applications of MBR is introduced and some full scale MRBs are also presented.

8.1 Installations Worldwide

Kubota (Japan) has installed most of the world's MBRs while Zenon (Canada) dominates in regard to installed capacity having almost four times more water treated through their membranes than Kubota. Zenon has installed about 85% of North American installations, which comprise about 11% of the world's MBR market. Asian markets (mostly in Japan and South Korea) have employed MBR technology mostly for small-scale domestic applications. In general, most of the MBRs in operation are medium or small scale plants. More than 85% of Kubota's MBRs have flows less than 200 m^3 d^{-1} while out of 219 MBR plants that treat municipal wastewater in North America, only 17 exceed 10 000 m^3 d^{-1}. The largest capacity plant in operation is in North America, which operates at 26 900 m3 d^{-1}, while the largest MBR worldwide currently is in Kaarst, Germany (48 000 m^3 d^{-1}) with total membrane area of 84 480 m^2. Both of these plants operate with Zenon membranes. Leading manufacturers have exponential growth in the number of installed MBRs and their cumulative capacity in the last decade. Although the market is still dominated by Zenon and Kubota, there is a wide range of MBR systems available; however most are still at the development stage. The photographs of typical Kubota and Zenon membranes are presented in Figure 8.1 and 8.2, respectively (Radjenović et al. 2008).

Figure 8.1 Kubota flat sheets MBR

The side-stream MBRs that were predominant before the 1990s are still present on the market but they hold a smaller share. The main manufacturers of side-stream MBR systems are Norit X-Flow, Millennium pore and Novasep-Orelis. Most of the side-stream MBRs today treat industrial wastewaters or landfill (Radjenović et al. 2008).

Municipal wastewater treatment is both the earliest and largest application of MBR, and it is predicted that this will continue to be its primary use. Due to its small footprint and potential for reuse of high-quality effluent, MBR is capable of coping with population growth and limited space. For industrial applications where more stringent regulations are imposed, it provides an effluent that can be safely discharged into the environment. The main applications of membrane technology reported in industry are for treatments of heavily loaded wastewaters such are oily wastewaters, or discharges from tanneries and textile industries (Radjenović et al. 2008).

Figure 8.2 Zenon hollow fiber MBR

8.2 Plant size

Earlier MBR technology was favoured in difficult applications or those applications where compactness was important and reuse was the target; and it usually involved smaller plants. As the demand for MBR technology grows globally, both the number of installations and the capacity of the installed plants are increasing dramatically. The most optimistic industry estimates suggest that up to 1000 new MBR plants will be built annually during the survey period. The size of the constructed plants has grown from facilities treating hundreds to thousands of gallons of wastewater per day to those treating tens of millions of gallons per day in just a few years. However, the most common capacity for current worldwide MBR installations ranges from the 50 000 gpd (200 m^3 d^{-1}) to 500 000 gpd systems (Hai and Yamamoto 2011).

The largest MBR plant in the world is set to be operational in 2010/11 in King County, Washington State. When completed, the facility will have an initial peak flow capacity of 495 000 m^3 d^{-1} (average 136 000 m^3 d^{-1}), rising to a daily 645 000 m^3 (average 205 000 m^3) by 2040 (Hai and Yamamoto 2011).

8.3 Full scale membrane bioreactors

The above sections mainly introduce the knowledge about MBR. The section 8.3 tries to introduce some full scale MBR to readers, which is helpful for the readers to obtain more information about MBR. Information in this section can be refer to (Pinnekamp and Friedrich 2006).

8.3.1 Municipal wastewater

8.3.1.1 *Nordkanal wastewater treatment plants*

When the expansion of the Nordkanal wastewater treatment plant became necessary, the original site had to be given up due to the spatial development of the city of Kaarst. A new plant had to be built at another site. The Erftverband decided on the membrane bioreactor process because positive experience had been acquired with this process at the Rodingen wastewater treatment plant. The wastewater treatment concept was developed in close coordination with the Ministry for Environment and Nature Conservation, Agriculture and Consumer protection of the North-Rhine Westphalia. Due to its size, this plant represents new planning dimensions and has demonstration character throughout Europe.

The plant is designed for a capacity of 80,000 PE and a combined water flow of 1881 m^3/h. It was commissioned in 2003. The demands on the effluent quality are compiled in Table 8.1.

Parameter	Unit	Minimum requirements	Discharge consent
COD	mg/L	90	90
BOD_5	mg/L	20	20
NH_4-N	mg/L	10	10
N_{tot}	mg/L	18	18
P_{tot}	mg/L	2	2

Table 8.1 minimum requirements and discharge consent of the Nordkanal waste treatment plant

Figure 8.3 Rotary screen of the fine screen installation

At the site of the old wastewater treatment plant, the wastewater is pre-treated by a coarse screen and pumped to the new Nordkanal plant, situated a distance of 2.5 km away, where the wastewater is mechanically pre-treated by two step screens (5 mm spacing) operated in parallel and two aerated grit and grease traps also operated in parallel. Then the wastewater is treated by two rotary screens operated in parallel with an aperture size of 0.5 mm (figure 8.3) to protect the membranes in the nitrification stage. The emergency circuit of the rotary screens is made safe by a fine screen with an aperture size of 1mm, so that the membranes are protected from the input of coarse material into the activated sludge tank. Figure 8.4 shows the flow sheet of the Nordkanal wastewater treatment plant.

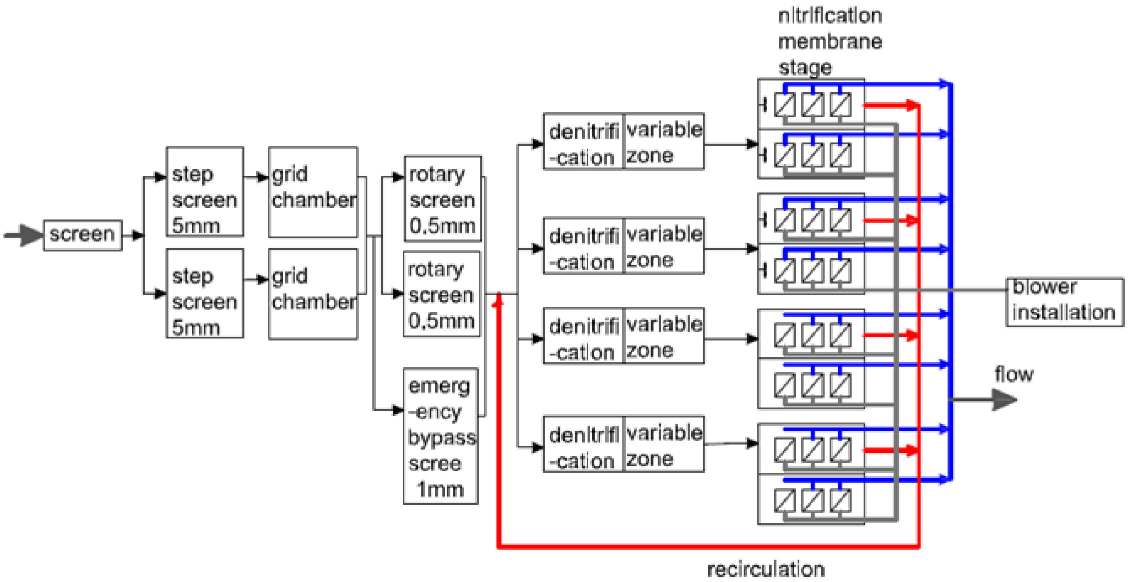

Figure 8.4 flow sheet of the Nordkanal waste water treatment plant

Figure 8.5 membrane installation at the WWTP Nordkanal

The activated sludge stage has four lines, each of which consists of upstream denitrification tanks, a variable tank zone for either denitrification or nitrification, and of the nitrification tanks with immersed membrane modules, designed as activated sludge tanks with circulating flow. The nitrification tanks are cased. The total volume of the activated sludge tanks is 9200 m^3. The sludge is stabilized aerobically. As a result of flow simulations, agitators and baffles were integrated into the activated sludge tanks with circulating flow.

The membrane installation has been realized with eight lines and equipped with capillary membranes from company ZENEN (ZE 500c). A total filter surface area of approximately 85000 m^3 has been installed because the District Government Dusseldorf demanded to provide a reserve of 25% for the membrane filtration. For external chemical cleaning, s separate cleaning chamber is available.

The investment for the new construction of the Nordkanal wastewater treatment plant was 21.5 million euro. Approximately, 6.6 million euro of this amount had been taken over by the federal state North Rhine-Westphalia.

8.3.1.2 *Monheim wastewater treatment plant*

The wastewater treatment plant of the city of Monhelm is situated in the sensitive karstland of the district Donauries. It treats not only the wastewater from the city of Monheim, but also from the municipalities of Rogling and Tagmersheim. The treated effluent is discharged into the Gailach which infiltrates into the karst 6 km downstream of Monheim. In 1998 and 1999, first concepts for the discharge of wastewater into the karst subsoil were developed.

Within the scope of the large scale pilot project 'wastewater treatment Gailach valley', the Free State of Bavaria supported the financing of the construction of a membrane bioreactor at the site of the Monheim wastewater treatment plant. The investment for the membrane bioreactor was approx. 7.6 million euro, of which 5.8 million euro were granted as subsidy by the state Bavaria.

The Monheim wastewater treatment plant is designed for a capacity of 9700 PE, based on a peak flow of 288 m^3/h and an average daily wastewater flow of 2400 m^3/d.

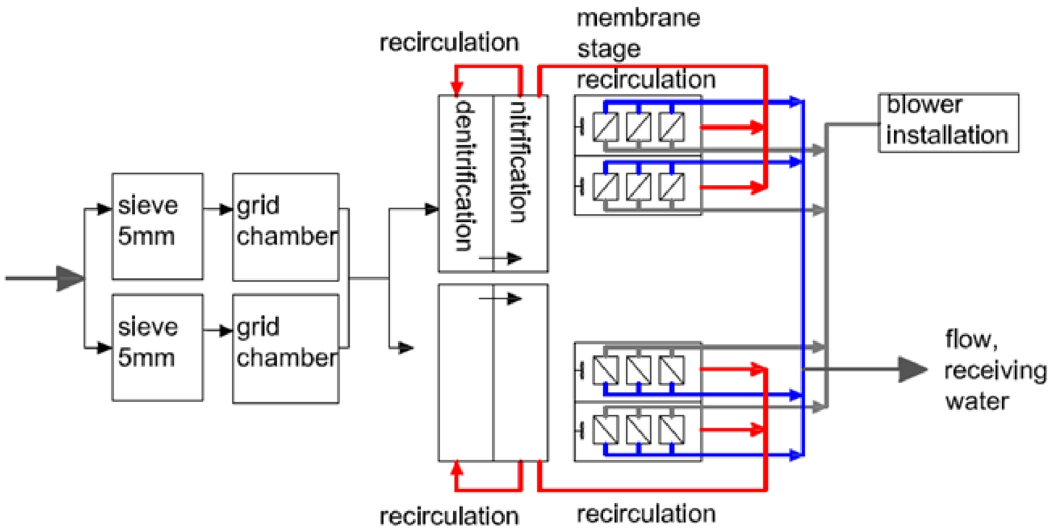

Figure 8.6 flow sheet of the Monheim wastewater treatment plant

As shown in the flow sheet of the Monheim wastewater treatment plant (figuer 8.6), the mechanical pretreatment stage has two lines. Each line consists of a fine sieve with an aperture size of 1mm and a grit chamber. 75% of the maximum inflow can be treated by each line. The mechanically pretreated wastewater flows into the activated sludge stage with a total volume of 1660 m^3, which is also built in two lines. Each line consistof an upstream denitrification and a nitrification tank as well as two membrane chambers which have been provided with a coating resistant to chemicals to protect the concrete. The tanks for the denitrification and nitrification have a volume of 240 m^3 each, which each of the four membraen chambers has a volume of 75 m^3. The sludge is stabilized aerobically.

The membarne stage was designed for a specific filtration capacity of 22–24 L/m^2.h of combined flow. This volume can be increased at short notice up to 31 L/m^3.h when one membrane chamber is shut down. According to this design, the membrane stage contains 28 module cassettes from the company ZENON (type ZW 500c) with a total membrane surface area of 12320 m^2 filter. Filtration lines are installed in four separate chambers, chemical cleaning of the modules can be realized by pumping off the activated sludge without removing the modules (on air).

The specific energy demand of the wastewater treatment plant is about 1 kWh per m^3 of wasteater. The manpower requriemnt corresponds to that of a conventional plant.

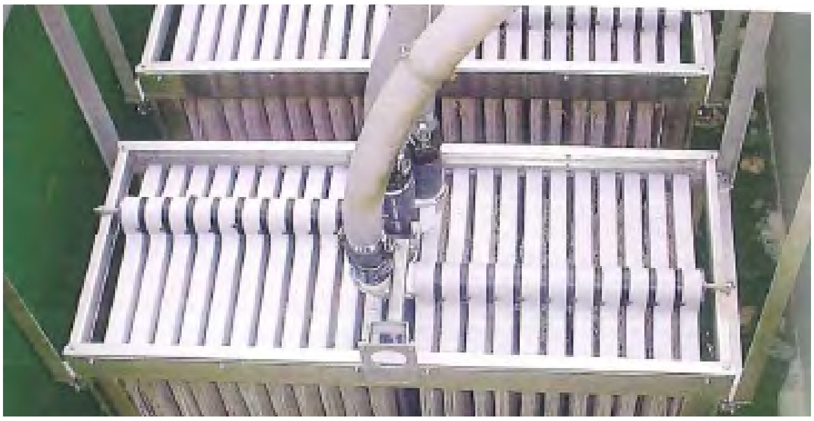

Figure 8.7 module cassettes during In-air cleaning

Using the membrane process at the Monheim wastewater treatment plant, the requirements for the effluent quality are safely met, as can be taken from table 8.2.

Parameter	Unit	Minimum requirement	Discharge consent	Operating values
COD	mg/L	90	75	5
BOD_5	mg/L	20	15	1.2
NH_4-N	mg/L	10	5	0.1
N_{tot}	mg/L	-	18	10
P_{tot}	mg/L	-	1	0.6

Table 8.2 minimum requirements, discharge consent and operating values of the Monheim wastewater treatment plant

At present, the operation of the Monheim membrane bioreactor is accompanied by a research program. Main items of this study include testing and optimization of the membrane bioreactor process and investigating the effects of wastewater discharge on the Gailach and the groundwater.

8.3.1.3 *Markranstadt wastewater treatment plant*

The Markranstadt wastewater treatment plant is situated in the southwest of Leipzig. It is one of more than 30 wastewater treatment plants of Kommunale Wasserwerke Leipzig (municipal waterworks). It was designed for 12000 PE; the actural degree of capacity utilization is approx..8000 PE.

The reason for a new construction of this plant was the planned closure of the obsolete wastewater treatment plant which no longer complied with the requirements. The deciding factors for the construction of a membrane bioreactor were the limited surface area of the site and increased demands on the effluent quality (table 8.3) due to a 'weak' receiving water.

Parameter	Unit	Minimum requirement	Discharge consent	Operating values
COD	mg/L	90	50	25
BOD_5	mg/L	20	10	5
NH_4-N	mg/L	10	5	1
N_{tot}	mg/L	18	18	15
P_{tot}	mg/L	2	2	1
Filterable solid	mg/L	No information	No set target	No information

Table 8.3 minimum requirements, discharge consent and operating values of the Markranstadt wastewater treatment plant

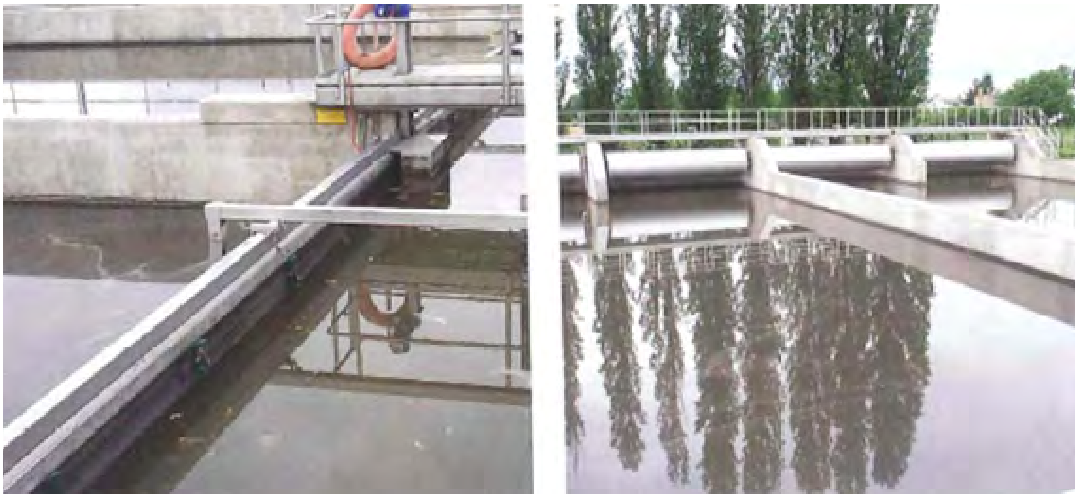

Figure 8.8 process stages at the Markranstadt wastewater treatment. Left: inflow chamber to the membrane bioreactor with overflow edge to combined water treatment, right: combined water treatment tank.

The plant has a hydraulic capacity of 180 m^3/h. From the intercepting sewer of the combined sewer system, the wastewater is fed by a lifting pump via the inflow chamber (figure 8.8, left) to the mechanical pre-treatment stage.

The two line mechanical pre-treatment stage consists of a step screen (3 mm spacing) (figure 8.9, left) and a grit and grease trap. By a distributor the wastewater flows into the two line activated sludge. It is operated as upstream denitrification (V_{DN} = 2.435 m^3) with downstream nitrification (VN=2.435 m^3). All tanks are equipped with agitators. In addition, aggregates for fine bubble aeration are installed over the whole surface area of the bottom of the nitrification tanks.

The membrane modules for biomass separation from the company ZENON are arranged at te inner longitudinal sides in the upper zone of the nitrification tanks with a depth of 7 m. The total filter surface area of 7360 m^2 is distributed in four lines, two each in both nitrification zones. Between the longwise arranged nitrification tanks a cleaning shaft for external module cleaning is installed. The modules can be removed by a fixed crane.

Besides the wastewater treatment plant, a combined water treatment plant was built in parallel at the same site. The wastewater quantities which exceed the capacity of the membrane stage during combined water flow are stored temporarily and pretreated in parallel in two tanks which serve as settling and storage tanks. These wastewater quantities are fed to the membrane installation during periods with smaller inflow volumes. Thanks to the combined water treatment plant, the necessary membrane surface area could be considerably reduced because it had to be designed for the maximum inflow quantity, but only for 1.1 $Q_{T.}$

Since the plant was commissioned in 2000, much knowledge has been acquired concerning the optimizaiton of process engineering and control. Improvement of mechanical pretreatment was especially important. The screen installed initally was replanced by a combinaiton of coarse screen (5 mm spacing) and fine sieve with a aperture size of <1 mm.

Figure 8.9 process stages at the Markranstadt wastewater treatment plant.
Left: step screen, right: nitrification and de-nitrification tanks

8.3.2 Industrial wastewater: examples from food industry

The generic term food industry is comprised of a large number of branches, such as the milk or meat processing industry, processing of vegetables, finished products, the beverage industry etc. Correspondingly, the wastewaters of the individual braches vary in their composition. They have in common only high organic loads.

In the following the use of membrane technology for the treatment of wastewater from the food industry is described for three branches: potato starch production, delicatessen production and malt production, and presented with the help of concrete examples.

Besides wastewater water treatment, membrane technology is also sued in the food industry for other purposes, such as concentration (e.g. of juice, milk, whey, egg whites), filtration (e.g. of juice, wine, beer) and alcohol removal from beer. The alcohol fraction resulting from alcohol removal is a suitable substrate for denitrification in wastewater treatment (as a methanol substitute).

8.3.2.1 *Potato starch production*

In the Federal Republic of Germany, starch is produced from maize, potatoes, wheat and rice. Of these raw materials the potato has the highest water content. For starch production, the potatoes are carefully prewashed and then ground, separated from the pulp water (0.76 m^3 of pulp waster per ton of potatoes) and washed out. The starch is produced from the ground potatoes, and the pulp water is generally used to produce potato protein. The residual pulp water is used for irrigation of farmland or evaporated. Potato pulp, which contains fine-ground peelings, cell walls, starch residues and pulp water, is dewatered. In Germany and the Benelux Countries, potato pulp has been used for many years as fodder for dairy cattle and young stock and also partly for fat stock.

The entire production process results in sweeping and washing water, pulp water and starch washing water. The amount of washing water is about 1.8 to 2.8 m^3 per ton of starch. Characteristic constituents are potato pulp water ingredients, fibres and mineral components (earth sand etc).

Potato starch is only produced only during a certain season. The Fertilizer Ordinance (1996) dictates the storage of potato pulp and irrigation waster between 15th November and 15th January (even longer in the case of frost) and limits the application in autumn to a maximum of 80 kg N_{tot}/ha.

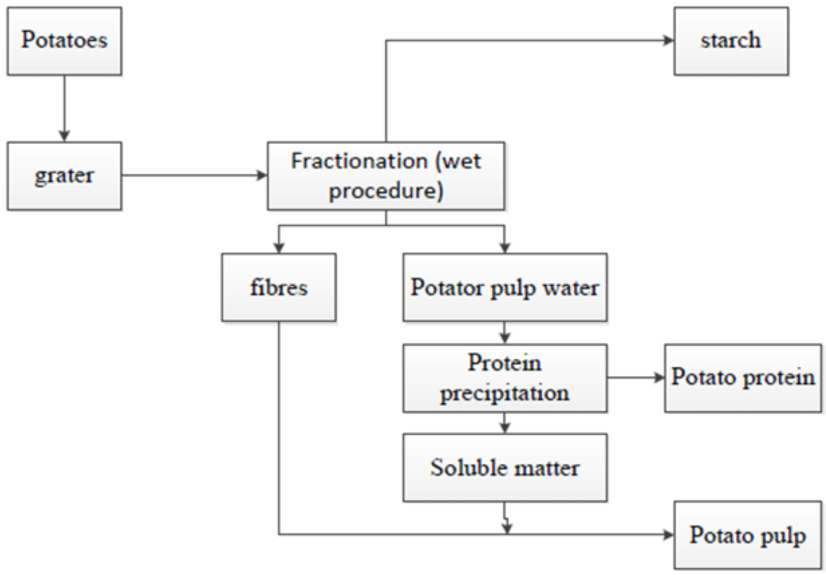

Figure 8.10 flow chart of potato starch production

The concentration of the pulp water and closing of the internal water cycle are suitable measures to manage the production limitations defeine above. This can be obtained with the help of different procedures (e.g. membrane technology).

8.3.2.2 *Food industry,emsland Starke GmbH*

The company Emsland Starke GmbH is the most important producer of potato starch in Germany and is one of the world's leading manufacturers of finished starch products, potato protein, amino acids and sugar products such as glucose syrup. The parent plant in Emlichheim, established in 1928, has 405 employees. For the production of potato starch, starch and potato protein, water with drinking water quality is used and ultimately disposed of as wastewater (sweeping and washing water, derivate waste water)

In 1997, a reverse osmosis installation from the company Stork was commissioned in the Emlichheim plant in order to reduce the quantity of wastewater t obe disposed of, to conserve drinking water and to recover more protein. The potato pulp wster from starch procution is separated in the crossflow mode at an operating pressur eof 40 bar. Tube modules with a total membrane surface area of 5000 m^2 filter a feed volume flow of 140 m^3/h. With daily backwashing and cleaning with commercial enzymatic cleaning agents, the servive life of the membrane is about 6000 hours. During the production campaign (about 120 days per year), the instalation works 24 hours perday, so that the membranes have to be replanced after approx.two campains.

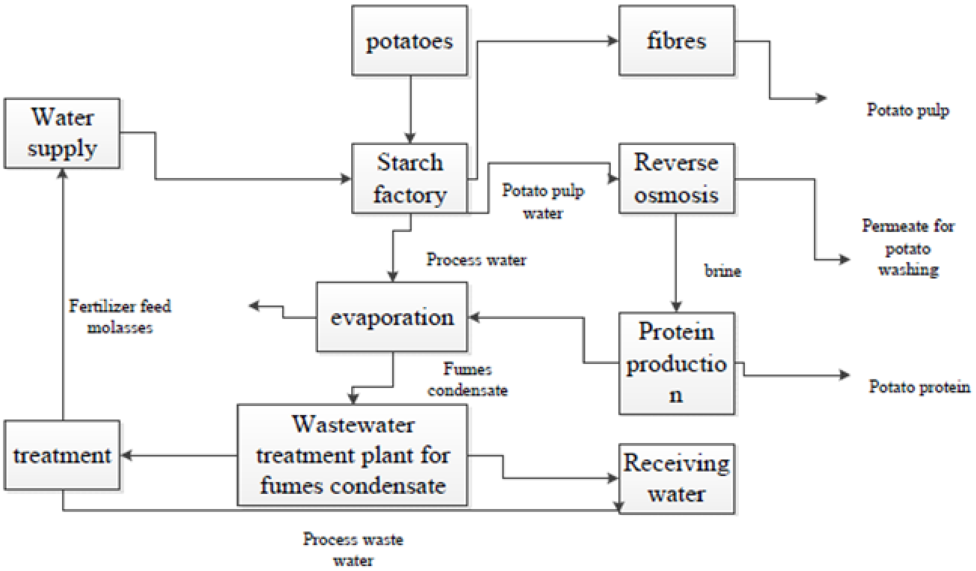

Figure 8.11 Flow chart of the treatment of process and potato pulp water at Emsland Starke GmbH

The permeate (about 62 m^3/h) from the reverse osmosis system is used to wash the potatoes, while potato protein is produced from the brine. The remaining potato pulp waster is evaporated. The vapour condenstes are fed to a dedicated waste water treatment plant and recycled after fruther treatment at drinking water quality into the production process (see also figure 8.11).

Thanks to the closed water cycle, more than 500,000 m^3 of water are saved per year (250,000 m^3 of washing water by the permeate of the reverse osmosis installation and more than 250000 m^3 by cosing the cycel with the vapour consenstes). Other advantages of the membrane installation are a dreastic reduction in the wastewater volume and a higher protein yield.

8.3.2.3 *Malt house*

Malt is used as a raw material to produce alcohol from starch-containing materials. Today a large number of breweries get malt from commercial malthouse which predominantly use barley (about 2.5 million tons per year) and partly wheat for malt production. The malt productionprocess can be broadly subdivided into the steps cleaning, soaking, germinating and kilm-drying. Due to high waster consumption for washing and soaking, malthouses have to pay high costs for freshwater and waste water disposal.

For wastewater whose pollution load comes mainly from the production of malt from cereals and which is discharged directly into a receiving water. Depending on the production process applied, the wasteater quantities and concentration may vary significantly from one malthouse ot the other. Wastewater constituents include suspended substances (dust, earthy constituents, residues from cereals and husks), sugar, nitrogen-contaning substances (soluble protein, vegetable fibrin), inorganic matter, and possibly rubber and polyphenols.

For the treatment of malhouse wastewater, membrane technology can be used in various combinations which have to be adapted to the specific case. Besides the example described below, treatemnt by microfiltration in low pressure operation combined with a biological stage and a closed process water cycle is also possible.

The Heidelsheim company H.Durst Malzfabriken GmbH & Co.KG is specialized in malt production for Pilsner bear, but also furnishes special malt for other types fo brewing. 25 people are employed in the works in Gernsheim, one of the four plants which combined produce about 230000 tons of malt per year.

Large wasewater quantities and disposal costs were motives for Durst Malz to cooperate with Schwander GmbH at Bad Vilbel, which together with Frings Recycling-Anlagen GmbH developed the patented FriSsh-Verfahen for the treatment of process water in the malt and beverage industry.

Promoted by Hessische Landes-und Treuhandgesellschaft (HLT) Wiesbaden, today Investbank Hessen (IBH), an installaiton for the treatment of the malthouse wsteater was commissined in 1997 at Gernsheim. Since that time a daily amoutn of 700 m^3 of water, consisting of the barley soaking water and the washing water fo the production plants, is treated.

The malthouse wastewater has a high COD content of approx 2500 to 3000 mg/L. By biological (SBR process) and physical-chemical (ferric chloride precipiation) treatment, followed by fine filtration, this concentratioin is reduced to 30 mg/L in the influent of the reverse osmosis installation. Subsequently, spiral-wound modules in the reverse osmosis installation (imb+frings water systems gmbh) with a total filter surface area of 1333 remove all undesirable malting residues from thewater at an operating pressure of approx 10 bar. The permeate complies with the requriments of the Drinking Water Ordinance and is recycled into the barley soaking process. The remaining brine (about 25–30% fo the total inflow to the membrane installation) is treated int the municipal wastewater treatemnt plant. In order to ensure the oneartion of the spiral-wound modules, and antiscaling agent is used and th emodels are backwashed daily with citric acid.

The employement of the memrbaen installation led to a reduction in the water demand and to considerable cost saving concerning the wasteater surfachge. Another advantage is the modular construciton of the installaiton, as it can be adapted without causing problems to varying production parameters.

9 Anaerobic digestion and anaerobic membrane bioreactor

Anaerobic membrane bioreactor (AnMBR) is a developing technology. The AnMBR is the combination of anaerobic digestion technology and membrane technology. In order to have a good understanding of AnMBR, readers should have basic understanding of anaerobic digestion. A introduction of anaerobic digestion is summarized in the section 9.1 (van Lier et al. 2008), and a brief introduction of AnMBR and one of its application are provided in the section of 9.2 and 9.3, respectively.

9.1 Anaerobic digestion

The anaerobic degradation pathway of organic matter is a multi-step process of series and parallel reactions. Methanogenic bacteria are located at the end of the anaerobic food chain and, partly thanks to their activity; no large quantities of organic matter accumulate in anaerobic environments, where this matter is inaccessible to aerobic organisms. The anaerobic digestion process involves a complex food web, in which organic matter is sequentially degraded by a wide variety of micro-organisms. The microbial consortia involved jointly convert complex organic matter and ultimately mineralize it into methane (CH_4), carbon dioxide CO_2, ammonium (NH_3), hydrogen sulphide (H_2S) and water (H_2O).

The digestion process may be subdivided into the following four phases:

9.1.1 Hydrolysis

Since bacteria are unable to take up particulate organic matter, the first step in anaerobic degradation consists of the hydrolysis of polymers. This process is merely a surface phenomenon in which the polymeric particles are degraded through the action of exo-enzymes to produce smaller molecules which can cross the cell barrier. During the enzymatic hydrolysis process, proteins are hydrolysed to amino acids, polysaccharide to simple sugars and lipids to long chain fatty acids (LCFA). Hydrolysis is in most cases, notably with (semi-) solid substrates and wastewaters with a high suspended solids (SS)/COD ratio, rate-limiting for the overall digestion process. Moreover, the hydrolysis process is very sensitive to temperature and temperature fluctuations. For that reason, the design of anaerobic digesters for (semi-) solid substrates and wastewaters with a high SS/COD ratio, such as distillery slops and low temperature sewage, is usually based on the hydrolysis step.

Hydrolysis can be defined as a process in which complex polymeric substrates, particulate or dissolved, are converted into monomeric and dimeric compounds which are readily accessible for the acidogenic bacteria. During anaerobic digestion of complex substrates hydrolysis is usually the first step. Although in some cases a preparatory step, i.e. physical-chemical pre-treatment, is needed to make hydrolysis possible. With the digestion of biological sludges, such as waste activated sludge, the hydrolysis of the sludge is preceded by death and lysis of the biomass. The hydrolysis is accomplished by exo-enzymes which are produced by the acidogenic bacteria. The products of the hydrolysis are the substrates for the acidogenic bacteria.

As mentioned, hydrolysis is generally considered to be the rate-limiting step during the anaerobic digestion of complex substrates. However, usually this is not due to a lack of enzyme activity but to the availability of free accessible surface area of the particles and the overall structure of the solid substrate. Even in dilute wastewaters such as low temperature domestic sewage, hydrolysis may determine the overall process and thereby determining the required reactor design. It must be noted that 45–75% of domestic sewage, and 80% in primary sludge consists of suspended matter. The main biopolymers in sewage are proteins, carbohydrates and lipids.

9.1.2 Acidogenesis

During the acidogenesis step, the hydrolysis products (amino acids, simple sugars, LCFAs), which are relatively small soluble compounds, are diffused inside the bacterial cells through the cell membrane and subsequently fermented or anaerobically oxidized. Acidogenesis is a very common reaction and is performed by a large group of hydrolytic and non-hydrolytic microorganisms. About 1% of all known bacteria are (facultative) fermenters. The acidification products consist of a variety of small organic compounds, mainly VFAs, i.e. acetate and higher organic acids such as propionate and butyrate, as well as H_2, CO_2, some lactic acids, ethanol and ammonia.

Characteristically, neutral compounds such as sugars and proteins are converted into VFAs and carbonic acid, being the main end products. Therefore, fermentative organisms are usually designated as acidifying or acidogenic microorganisms, and the process is therefore indicated by acidogenesis. Table 9.1 lists several acidogenic reactions starting from sucrose and generating different amounts of VFAs, HCO_3^-, H_2, H^+. Apparently, the type of end products depends on the conditions in the reactor medium. From Table 9.1 it follows that the ΔG° of the less energetic acidogenic reactions with sucrose as the substrate strongly depends on the prevailing H_2 concentrations. If H_2 is effectively removed by H_2 scavenging organisms such as methanogens, acetate will be the main end product. However, if methanogenesis is retarded and H_2 accumulates, more reduced products such as propionate and butyrate are likely to appear and possibly the even more reduced compounds lactate and alcohols. Therefore, effluents of overloaded or perturbed anaerobic reactors (or reactors designed as acidifying reactors in an anaerobic two-step process) often contain these more reduced intermediate products.

Reactions	ΔG°' (kJ/mol)	Eq.
$C_{12}H_{22}O_{11} + 9H_2O \rightarrow 4CH_3COO^- + 4HCO_3^- + 8H^+ + 8H_2$	- 457.5	(16.1)
$C_{12}H_{22}O_{11} + 5H_2O \rightarrow 2CH_3CH_2CH_2COO^- + 4HCO_3^- + 6H^+ + 4H_2$	- 554.1	(16.2)
$C_{12}H_{22}O_{11} + 3H_2O \rightarrow 2CH_3COO^- + 2CH_3CH_2COO^- + 2HCO_3^- + 6H^+ + 2H_2$	- 610.5	(16.3)

Table 9.1 Acidogenic reactors with sucrose as the substrate and the corresponding free energy change (ΔG°) at °C

Process	Conversion rate gCOD/gVSS.d	Y gVSS/gCOD	K_s mgCOD/l	μ_m 1/d
Acidogenesis	13	0.15	200	2.00
Methanogenesis	3	0.03	30	0.12
Overall	2	0.03-0.18	-	0.12

Table 9.2 Averaged kinetic properties of acidifiers and methanogens

Acidogenesis is the most rapid conversion step in the anaerobic food chain. The ΔG° of acidifying reactions is highest of all anaerobic conversions, resulting in ten to twentyfold higher bacterial growth rates, and fivefold higher bacterial yields and conversion rates compared to methanogens (Table 9.2). For that reason, anaerobic reactors are subjected to souring, i.e. a sudden pH drop, when reactors are overloaded or perturbed by toxic compounds. Once alkalinity is consumed by the produced acids the pH starts to drop, resulting in a higher concentration of non-dissociated VFAs, leading to a more severe inhibition of methanogens. The latter, obviously leads to an even quicker accumulation of VFAs and subsequent pH drop (Figure 9.1).

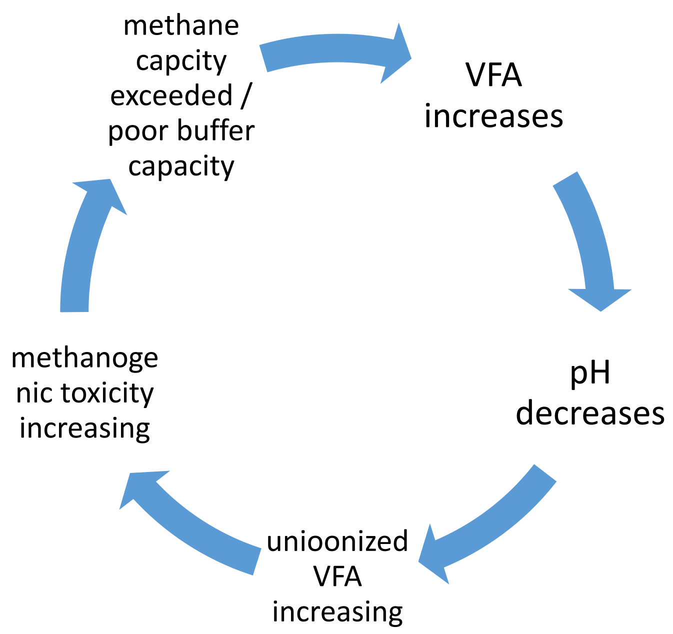

Figure 9.1 reactor pH drop as result of methanogenic overloading and accumulating VFAs

The fact that acidifiers are active even at low pH (4), means the reactor souring to pH 4 to 5 can and will occur when the methanogenic capacity of the system is trespassed.

The acidogenic conversion of amino acids generally follows the Stickland reaction, in which an amino acid is de-ammonified by anaerobic oxidation yielding also VFA and H_2, in conjunction with the reductive deammonification of other amino acids consuming the produced H_2. From both reactions NH_3 is released and subsequently acts as a proton acceptor, thus leading to a pH increase. In this reaction there is no net proton production and there is no chance of reactor pH drop.

9.1.3 Acetogenesis

The short chain fatty acids (SCFA), other than acetate, which are produced in the acidogenesis step are further converted to acetate, hydrogen gas and carbon dioxide by the acetogenic bacteria. The most important acetogenic substrates are propionate and butyrate, key intermediates in the anaerobic digestion process. But also lactate, ethanol, methanol and even H_2 and CO_2 are (homo) acetogenically converted to acetate as shown in Figure 9.2 and Table 9.3. LCFAs are converted by specific acetogenic bacteria following the so-called β-oxidation in which acetate moieties are split from the aliphatic chain (Table 9.3). LCFAs with uneven C atoms also yield propionate next to acetate. Non-saturated LCFAs like oleate and linoleate are firstly saturated by H_2 addition prior to the β-oxidation. The acetogenic bacteria are obligate hydrogen producers and their metabolism is inhibited by hydrogen, which immediately follows from the stoichiometric conversion reaction, such as for propionate:

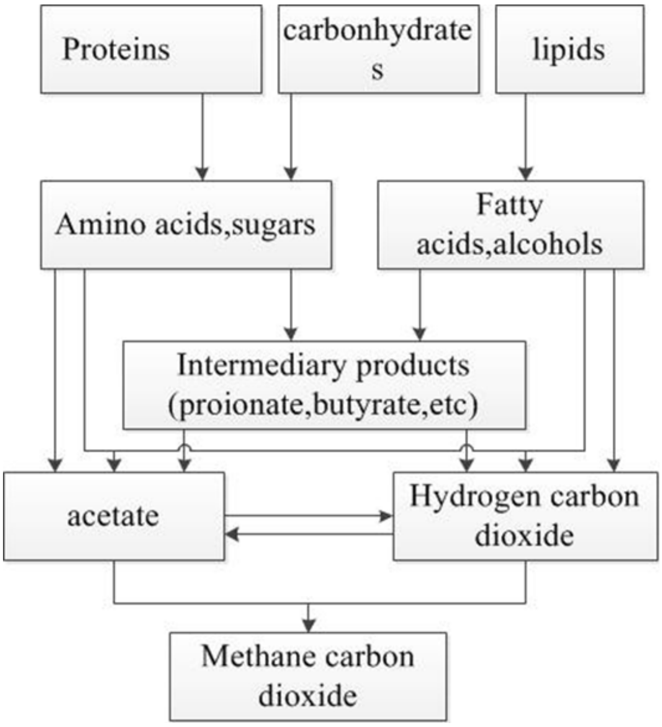

Figure 9.2 Reactive scheme for the anaerobic digestion of polymeric materials. Numbers indicate the bacterial groups involved: 1. Hydrolytic and fermentative bacteria, 2. Acetogenic bacteria, 3. Homo-acetogenic bacteria, 4. Hydrogenotrophic methanogens, 5. Aceticlastic methanogens

Compound	Reaction	ΔG°' (kJ/mol)	Eq.
Lactate	$CH_3CHOHCOO^- + 2H_2O \rightarrow CH_3COO^- + HCO_3^- + H^+ + 2H_2$	-4.2	(16.5)
Ethanol	$CH_3CH_2OH + H_2O \rightarrow CH_3COO^- + H^+ + 2H_2$	+9.6	(16.6)
Butyrate	$CH_3CH_2CH_2COO^- + 2H_2O \rightarrow 2CH_3COO^- + H^+ + 2H_2$	+48.1	(16.7)
Propionate	$CH_3CH_2COO^- + 3\,H_2O \rightarrow CH_3COO^- + HCO_3^- + H^+ + 3H_2$	+76.1	(16.8)
Methanol	$4\,CH_3OH + 2\,CO_2 \rightarrow 3CH_3COOH + 2H_2O$	-2.9	(16.9)
Hydrogen-CO_2	$2\,HCO_3^- + 4\,H_2 + H^+ \rightarrow CH_3COO^- + 4\,H_2O$	-70.3	(16.10)
Palmitate	$CH_3\text{-}(CH_2)_{14}\text{-}COO- + 14H_2O \rightarrow 8CH_3COO- + 7H+ + 14H_2$	+ 345.6	(16.11)

Table 9.3 Stoichiometry and change of free energy (ΔG°)for some acetogenic reactions, assuming neutral pH, a temperature of 25°C and a pressure of 1 atm (101 kPa). Water is regarded as a pure liquid, and all soluble compounds have an activity of 1 mol/kg

Studies carried out on acetogenic conversions have elucidated the required narrow associations between the H_2-producing acetogenic bacteria and the H_2-consuming methanogenic bacteria, thereby regulating the H_2 level in their environment. This is of vital importance as these reactions are thermodynamically unfavourable, indicated by the positive ΔG° in Table 9.3. From this table it follows that the reactions for ethanol, butyrate, propionate and the LCFAs palmitate will not occur under standard conditions, as the ΔG° is positive, and thus the bacterial energy yield is negative.

However, under stabilised digestion conditions the hydrogen partial pressure is maintained at an extremely low level. This can be achieved by an effective uptake of the hydrogen by methanogens or sulphate reducing bacteria. Methanogenic bacteria usually utilize molecular hydrogen in the anaerobic digester so rapidly that the hydrogen partial pressure drops below 10^{-4} atm, which is enough to ensure the actual occurrence of the hydrogen producing acetogenic reaction (Figure 9.3).

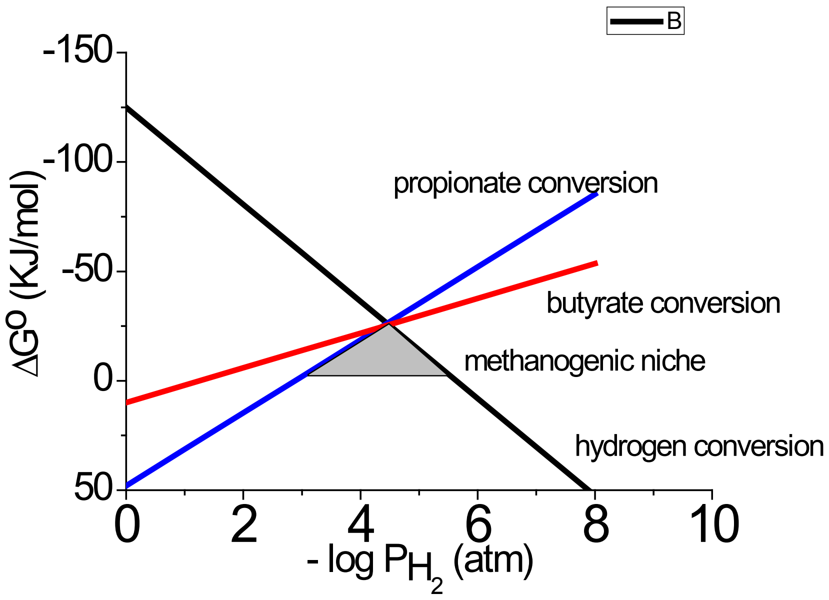

Figure 9.3 Free energy change as a function of the H_2 partial pressure.
A negative ΔG° indicates possible occurrence of the mentioned reaction

This interdependence means that the degradation of higher fatty acids and alcohols largely depends on the activity of electron scavenging organisms such as methanogenic bacteria. Microbial associations in which a H_2-producing organism can grow only in the presence of a H_2-consuming organism are called syntrophic associations. The coupling of formation and use of H_2 is called interspecies hydrogen transfer. In a properly functioning methane-producing installation, the partial hydrogen pressure will not exceed 10^{-4} atm and is usually between 10^{-4}–10^{-6} atm. At such a low hydrogen concentration, the degradation of ethanol, butyrate or propionate becomes exergonic and will yield energy for the acetogens.

Similar to the other acetogenic substrates, LCFA conversion is highly endergonic and often limits the entire digestion process. Trials with upflow anaerobic sludge blanket (UASB) reactors were only partly successful as LCFA tend to absorb to the sludge forming fatty clumps of biomass with little if any methanogenic activity. Expanded bed reactors, in which the LCFA is more evenly distributed over the available biomass were more successful. Other authors propose in fact to use the absorptive capacity of the sludge and periodically load the sludge with LCFA after which solid state digestion will convert the absorbed matter to CH_4. Such a sequencing bed mode of operation requires multiple reactors to treat a continuous flow wastewater.

9.1.4 Methanogenesis

Methanogenic bacteria accomplish the final stage in the overall anaerobic conversion of organic matter to methane and carbon dioxide. During this fourth and last stage of anaerobic degradation of organic matter, a group of methanogenic archea both reduce the carbon dioxide using hydrogen as electron donor and decarboxylate acetate to form CH_4. It is only in this stage when the influent COD is converted to a gaseous form that automatically leaves the reactor system. Methanogens are obligate anaerobes, with a very narrow substrate spectrum. Some can only use certain determined substrates such as acetate, methylamines, methanol, formate, and H_2/CO_2 or CO. For engineering purposes, methanogens are classified into two major groups: the acetate converting or aceticlastic methanogens and the hydrogen utilising or hydrogenotrophic methanogens (Table 9.4). Generally, about 70 % of the produced methane originates from acetate as the main precursor. The rest mainly originates from H_2 and CO_2. The growth rate of the aceticlastic methanogens is very low, resulting in doubling times of several days or even more. The extremely low growth rates explain why anaerobic reactors require a very long start-up time with unadapted seed material and why high sludge concentrations are pursued. Hydrogenotrophic bacteria have a much higher maximum growth rate than the acetoclastic bacteria with doubling times of 4 to 12 hours. Because of this feature and despite the very delicate acetogenic reaction step discussed in the previous section, anaerobic high rate reactor systems exert a remarkable stability under varying conditions.

Functional step	Reaction	ΔG, KJ/mol	μ_{max}, 1/d	T_d d	K_s mg COD/l
Acetotrophic methanogenesis	CH_3COO-+H_2O—— CH_4+HCO_{3-}	-31	0.12a 0.71b	5.8[a] 1.0[b]	30[a] 300[b]
Hydrogenotrophic methanogenesis	CO_2+4H_2——CH_4+2H_2O	-131	2.85	0.2	0.06

Table 9.4 Most important methanogenic reactions, the corresponding free energy change (ΔG°) and some kinetic properties

Table 9.4 lists two types of aceticlastic methanogens with very different kinetic characteristics. Also the morphological characteristics of both methanogenic genera are very different as indicated by Figure 9.4.

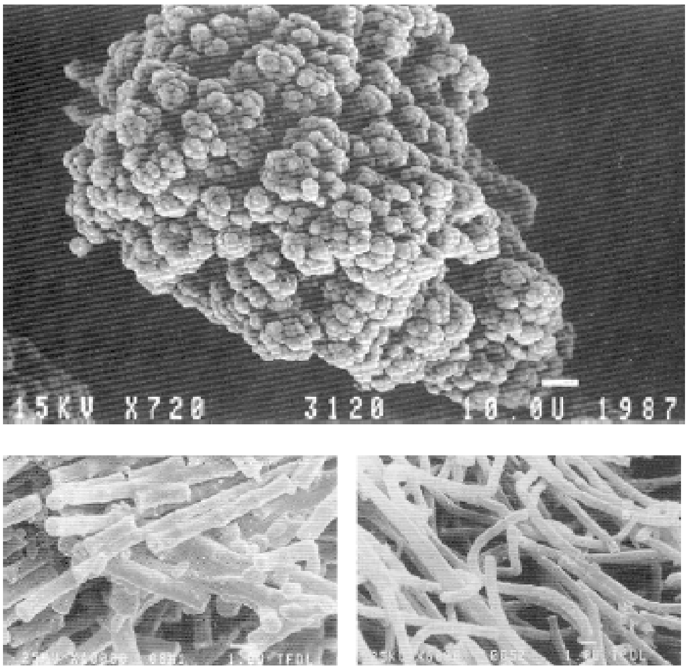

Figure 9.4 Morphology and appearance of the most important acetotrophic methanogens belonging to the genera Methanosarcina (above) and Methanosaeta (below)

Methanosarcina spp. are characterised by a coccoid shape, appearing in small grape-like clumps, and have a relative wide substrate spectrum as they can convert a.o. acetate, H_2/CO_2, methylamines, methanol, and formate. They have a relatively high μ_{max} and relative low substrate affinity. *Methanosaeta* spp. are filamentous, appear in large spaghetti like conglomerates can only convert acetate and are kinetically characterised by a low μ_{max} and a very high substrate affinity. Although the μ_{max} of the latter organism is significantly lower, *Methanosaeta* spp. are the most common acetotrophic methanogens in anaerobic high rate systems based on high solids retention times, such as sludge bed systems and anaerobic filters. The reason for this phenomenon can be attributed to the fact that wastewater treatment systems always aim at the lowest possible effluent concentrations, while substrate concentrations inside biofilms or sludge granules of the mentioned anaerobic systems approaches 'zero' when bulk liquid concentrations are low. Under such conditions, Methanosaeta spp. species have a clear kinetic advantage over the Methanosarcina spp. (Figure 9.5).

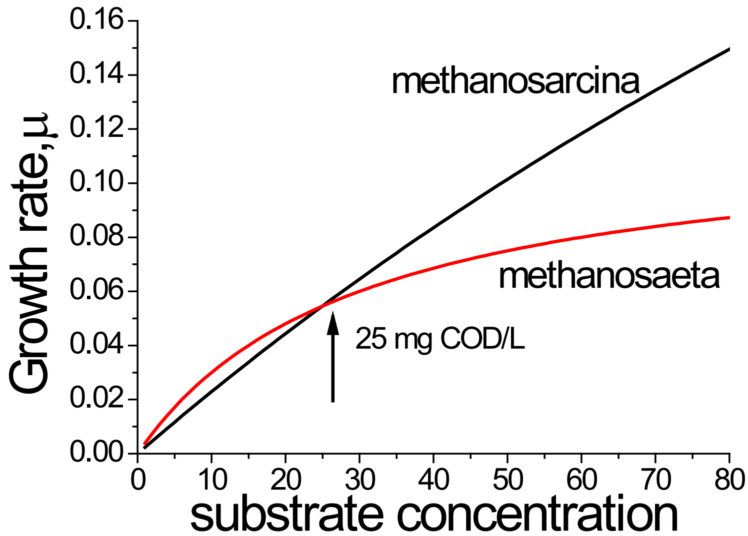

Figure 9.5 Monod growth curves of the acetotrophic methanogens Methanosarcina spp. and Methanosaeta spp. Both μ_{max} and the Monod half saturation constant (Ks) of both genera is given in Table 9.4

Once the Methanosaeta spp. dominates the sludge bed, a very effective wastewater treatment system is obtained, reaching extremely low effluent acetate concentrations. Considering the inferior kinetic properties at low substrate concentrations and the inferior adherence properties of Methanosarcina spp., it is advised to keep the effluent acetate concentrations at a very low level during the first start-up of an anaerobic reactor with un-adapted seed material.

9.2 Anaerobic MBR

The anaerobic membrane bioreactor (AnMBR) process was developed and tested in the mid 1990's and been used in full-scale since 2000. It was concluded that in 2008 there were fourteen full-scale AnMBR installations operating in Japan treating alcohol production stillage, organic wastes, septage wastewater treatment sludge, and a variety of residues from food processing (e.g., dairy, potato, confectionery, etc.). Pilot-scale AnMBR studies were also carried out on potato processing wastewater and has shown very favourable results providing maximum biogas production, producing less waste sludge and a superior effluent quality. The first full-scale AnMBR system in North America was installed at Ken's Foods in Massachusetts, USA in 2008. It provides waste treatment and energy production from salad dressing and BBQ sauce wastewater (Liao et al. 2006).

The AnMBR process provides numerous process advantages over conventional anaerobic wastewater treatment processes including (Liao et al. 2006):

1. A superior quality effluent is produced on a consistent basis. Typically, the AnMBR effluent quality is sufficient to significantly reduce, and sometimes even completely avoid, aerobic post-treatment.
2. High-rate organic loading rates (10 to 15 kg COD/m^3·d) are achievable, minimizing reactor size and footprint of the treatment plant.

3. Suspended solids (TSS) removal is not required ahead of the process. In some cases, fat/oil/grease (FOG) removal is not required either. This allows TSS and FOG to be digested which simplifies the overall system; eliminates primary treatment; increases biogas yield; and reduces waste sludge production, handling and disposal, and associated costs.
4. Waste activated sludge from a downstream aerobic process can be digested in the AnMBR system. This, again, simplifies the overall process; increases biogas yield; and further reduces sludge production, handling and disposal, and associated costs.
5. Complete retention of biomass (by use of membranes) assures a consistently high degree of treatment with negligible effluent TSS concentrations and superior process stability.
6. Granular sludge is not required, eliminating costs associated with obtaining granular seed sludge and maintaining this type of special sludge in the system.

9.3 A application of anaerobic MBR

An AnMBR pilot study was operated on site at Ken's Foods from September 2007 to March 2008 to evaluate the AnMBR treatment process as a means of upgrading their existing anaerobic treatment system which was space-limited. The pilot study results concluded that the AnMBR process was suitable for meeting the final effluent limits on a consistent basis (without post-aerobic biological treatment for BOD/TSS removal). Also, the system would easily increase operating capacity with minimal additional footprint due its ability to operate at higher biomass concentrations and organic loading rates. Based on the successful pilot study, Ken's Foods contracted ADI Systems to design and construct the full-scale AnMBR system in early 2008; it was commissioned in July 2008 (Christian et al. 2010).

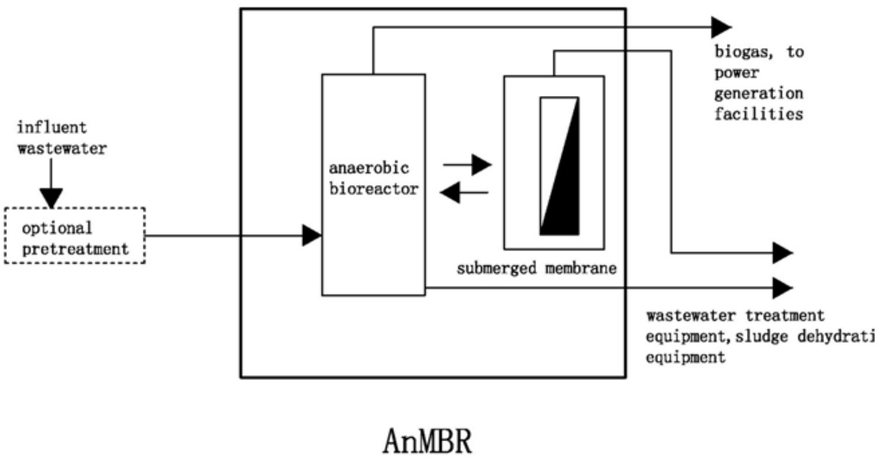

Figure 9.6 general process flow diagram of the AnMBR process

Figure 9.6 presents a process flow diagram of the AnMBR system. The AnMBR system consists of an existing proprietary 8300 m^3 reactor and four new anaerobic membrane tanks (each with a working volume of 100 m^3) equipped with Kubota submerged membrane cassettes. The membrane tanks operate in conjunction with the existing BVF reactor and have a continuous recycle loop in between. Seed sludge for the four anaerobic membrane tanks consisted of effluent TSS from the existing anaerobic BVF reactor and was concentrated by managing sludge recycle flow from the AnMBR tanks to the existing BVF reactor. Using the existing anaerobic system biomass as seed sludge enabled the AnMBR to begin treating 100 percent organic load and flow during the first week of commissioning (Christian et al. 2010).

The overall system also includes sludge recycle and effluent pumps, biogas scour blowers, process piping and valving, membrane cleaning chemical system, and instrumentation and control. A removable geomembrane cover system on each membrane tank provides a gas tight seal with biogas collection capabilities (Christian et al. 2010).

The AnMBR system was commissioned in July 2008 and is operated at mesophilic conditions with an average operating temperature of 33 °C for all four membrane tanks. The temperature is controlled through a spiral heat exchanger, with heat supplied by a biogas-fuelled hot water boiler. The continuous recycle of biomass between the BVF and the AnMBR tanks also provides the means of maintaining the temperature at the mesophilic condition. The average pH in the AnMBR is maintained well at 6.9 ± 0.1 for anaerobic sludge acclimatization (Christian et al. 2010).

Table 9.5 presents the raw wastewater characteristics and average effluent quality over the past 20 months.

Parameter	Raw wastewater	AnMBR effluent	% removals
Flow rate,m^3/d	300	300	-
BOD, mg/L	18000	20	99.9
COD,mg/L	33500	210	99.4
TSS, mg/L	10900	<1	100
pH	-	7.05	--

Table 9.5 raw wastewater and AnMBR effluent characteristics (Christian et al. 2010)

The AnMBR system was designed to treat 475 m^3/d of raw wastewater with wastewater characteristics of 39,000 mg/l COD, 18,000 mg/l BOD, and 12,000 mg/l TSS. Due to the fluctuating influent flow and varying wastewater characteristic, the actual influent flow is 300 ± 70 m^3/d with 33,500 ± 9,350 mg/l COD and 10,900 ± 3,200 mg/l TSS. The AnMBR system consistently produces an effluent with non-detectable TSS concentrations and COD and BOD concentrations of 209 ± 39 mg/l and 16 ± 5 mg/l, corresponding to COD and BOD removals of 99.4 and 99.9 percent, respectively (Christian et al. 2010).

Figures 9.7 and 9.8 presents the AnMBR influent and effluent COD and TSS concentrations and percent removals throughout the first 20 months of full-scale operation (Christian et al. 2010).

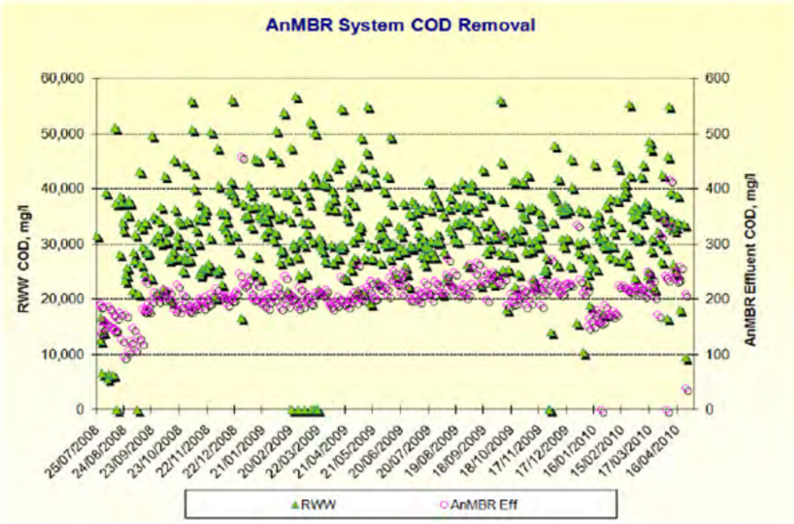

Figure 9.7 influent and effluent COD concentrations (Christian et al. 2010)

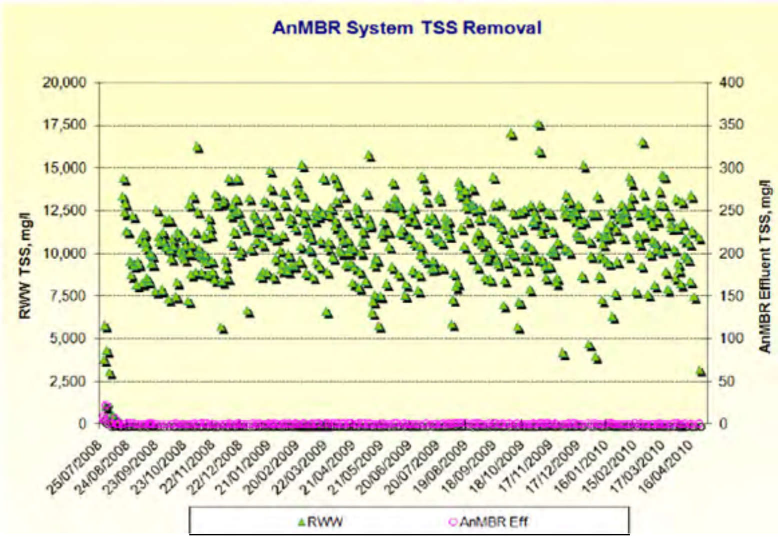

Figure 9.8 influent and effluent TSS concentration (Christian et al. 2010)

The system also provides complete digestion of the raw wastewater fat, oil, and grease concentrations providing for further biogas generation and better management of biomass and suspended solids concentration (Christian et al. 2010).

The AnMBR MLSS concentration ranges from 20,000 to 45,000 mg/l with an average of 23,000 mg/l in the four membrane tanks. The submerged Kubota membrane units provide a near-absolute barrier to block all suspended solids from escaping to the effluent, resulting in a very high quality effluent compared to conventional anaerobic technologies that rely on gravity settling in the reactor. The MLSS concentration is a function of sludge viscosity. Continuous sludge recycle from the anaerobic membrane tanks to the pre-existing BVF reactor minimizes the occurrence of sludge thickening or rapid increases in viscosity in the anaerobic membrane tanks. The average viscosity of the mixed liquor in four membrane tanks is 145 cP (Christian et al. 2010).

Since the AnMBR effluent contains no suspended solids, a waste sludge metering pump with piping connected to the AnMBR tanks sludge recycle piping to the final effluent discharge point allows Ken's Foods to discharge anaerobic sludge to the sewer freely while keeping within its TSS discharge limit of less than 230 kg/d. Therefore, no sludge has been intentionally wasted from the AnMBR system to the existing sludge dewatering system (resulting in a significant savings in dewatering chemicals) (Christian et al. 2010).

The AnMBR system operating flux rate ranged from 0.06 to 0.10 $m^3/m^2.d$; the design flux is 0.10 $m^3/m^2.d$. Figure 9.9 presents the transmembrane pressure (TMP) trend for each membrane tank over the past 20 months. Membrane cleaning is required when the TMP reaches 10 kPa (40 inches of water column). Some membranes in the system (three of four membrane compartments) were cleaned in situ on one occasion during the first 20 months of operation, as per scheduled maintenance activities; other membranes (in the fourth compartment) have still not been cleaned after 20 months of continuous operation. Operation to-date has resulted in an operating TMP of 12 inches of water column or less indicated that the membranes perform very well and that membrane fouling is not an issue (Christian et al. 2010).

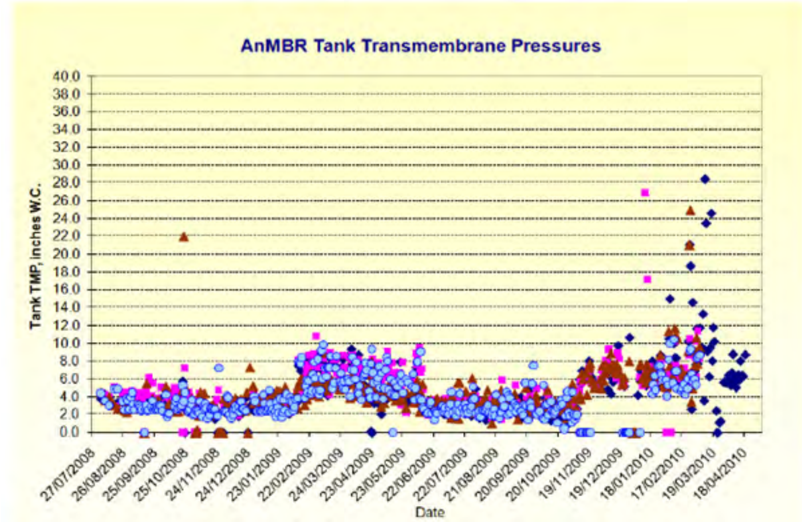

Figure 9.9 membrane performances (TMP) for each AnMBR tank (Christian et al. 2010)

The first 20 months of AnMBR system operation resulted in over a 50 percent reduction in operating expenses for the wastewater treatment plant compared to the prior 12 month fiscal period. This reduction was directly attributable to the increased system capacity, ability to treat wastewater with higher biomass in the AnMBR system, and elimination of the need to dewater and dispose of dewatered solids, and significant reduction in aeration power requirements for the existing SBR (Christian et al. 2010).

10 Abbreviations

AnMBR: anaerobic membrane bioreactor
BAF: biological aerated filter
BOD: biodegradable oxygen demand
b_H : specific rate of endogenous mass loss of OHOs
CAS: conventional activated sludge
CH_4: methane
CIP: chemical cleaning-in-place
COD: chemical oxygen demand
CO_2: carbon dioxide
C_c: solute concentration in concentrate.
C_f: solute concentration in feed flow;
C_p: solute concentration in permeate;
DNA: deoxyribonucleic acid
DO: dissolved oxygen
EBPR: enhanced biological phosphorus removal
EDCs: endocrine disrupting compounds
FOG: fat/oil/grease

EPS: extracellular polymeric substance
F/M: sludge organic load
$f'_{s'up}$: particulate un-biodegradable fraction of total influent COD
f_{cv} : COD to VSS ratio of the sludge
f_h : un-biodegradable fraction of the OHOs
$f_{i,OHO}$: inorganic contents of OHO
GAOs: glycogen accumulating organisms
H_2S: hydrogen sulphide
H_2O: water
J: Permeate flux
LCFA.: long chain fatty acids
MBR: membrane bioreactor
MF: Microfiltration
MLSS: mixed liquid suspended sludge
MWCO: molecular weight cut-off
NF: Nanofiltration
NH_3: ammonium
TN: total nitrogen
NOB: nitrate-oxidizing bacteria
OHOs: ordinary heterotrophic organisms
PAOs: phosphate-accumulating organisms
PE: polyethylene
PES: polyethylsulphone
PHAs: poly-b-hydroxyalkanoates
PVDF: and polyvinylidene fluoride
Q: flow rate
Q_c: solute concentration in concentrate;
Q_f: feed flow rate;
Q_p: permeate flow rate;
RO: Reverse osmosis
SCFA: short chain fatty acids
SEM: scanned electronic microscope
SMP: soluble microbial product
S_I: soluble and non-biodegradable
Ss; soluble and biodegradable
TKN: Kjeldahl nitrogen
TMP or ΔP: transmembrane pressure
TN: total nitrogen
TSS: total suspended solids

UASB: upflow activated sludge bed

UF: Ultrafiltration

UV: ultraviolet

VFAs: volatile fatty acids

VSS: volatile suspended solids

X_I: non-soluble and non-biodegradable

X_S: non-soluble and biodegradable

Y_{HV}: yield of OHO in terms of VSS

11 References

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12 About the author

Jixiang Yang obtained his Bachelor and Master's degree of municipal engineering at Harbin Institute of Technology and doctor's degree at Delft University of Technology. He worked with membrane technology for drinking water production and wastewater treatment in the past six years. He is now working in Chinese Academy of Sciences.

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