3 Reverse Osmosis Membrane

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3.1 INTRODUCTION

The fastest growing desalination process is a membrane separation process called reverse osmosis (RO). The most remarkable advantage of RO is that it consumes little energy since no phase change is involved in the process. RO employs hydraulic pressure to overcome the osmotic pressure of the salt solution, causing water-selective permeation from the saline side of a membrane to the freshwater side as the membrane barrier rejects salts [1–4]. Polymeric membranes are usually fabricated from materials such as cellulose acetate (CA), cellulose triacetate (CTA), and polyamide (PA) by the dry–wet phase inversion technique or by coating aromatic PA via interfacial polymerization (IFP) [5].

In the early 1960s, Loeb and Sourirajan developed asymmetric CA membrane by the phase inversion method for seawater desalination [6]. The fundamental principle underlying their RO membrane development is the preferential sorption-capillary flow mechanism. According to the Gibbs adsorption isotherm, a thin (about 0.5 nm) layer of pure water exists at the surface of seawater. By analogy, a pure water layer should also exist at the membrane/salt water interface, which can be collected through the membrane pore by applying a pressure gradient, as depicted in Figure 3.1 [7]. Later, it was found that the Loeb–Sourirajan membrane had an asymmetric structure with a top thin layer that governs the salt rejection and the water flux and a porous sublayer that provides the mechanical strength (see Figure 3.2) [7]. In 1968, Cadotte was able to prepare a new type of RO membrane called a thin-film-composite (TFC) membrane that consisted of a thin layer of PA formed by an in situ polycondensation reaction of branched poly(ethylene imine) and 2,4-diisocyanate on a porous polysulfone (PS) membrane [8]. Since then, the focus has been shifted to the development of TFC membranes with a thin skin layer, mostly made of aromatic PA material, and it has provided the major platform to fabricate novel RO membranes for seawater desalination. Nowadays, TFC membranes are manufactured by IFP of trimesoyl chloride (TMC) and m-phenylene diamine (MDP) at the surface of a porous PS support. Although the first practical PA-TFC membrane appeared nearly 40 years ago, there is still considerable uncertainty regarding the molecular mechanisms of their function, selectivity, and molecular structure–performance relationship. The main feature of the TFC membrane is that the material of the top thin layer (PA) is different from that of the porous support layer (PS), while, for the CA membrane, the material of both layers is the same. Therefore, the membrane of the latter type is called an integrally skinned asymmetric membrane to be distinguished from the TFC membrane, which also has an asymmetric structure.

Another noticeable development in RO membrane fabrication was that of the PA membrane with an integrally skinned asymmetric structure. It was made by DuPont in 1971 both in flat-sheet and hollow-fiber configurations [9,10], which established a novel membrane fabrication technology of hollow-fiber spinning. According to the patent disclosure, approximately 10% of the linear aromatic PA rings are substituted by sulfonic acid groups. As the aromatic ring is easily attacked by the chlorine in water, the presence of the sulfonate group is necessary to render the material chlorine resistant.
3.2 Membrane Preparation

3.2.1 Integrally Skinned Asymmetric Membrane and Thin-Film-Composite Membrane

The phase inversion technique is used to prepare integrally skinned asymmetric membranes [12]. This is a process in which a polymer is transformed from a liquid to a solid state. The dry–wet phase inversion technique and the temperature-induced phase separation (TIPS) are most commonly used in the industrial membrane manufacturing. Loeb and Sourirajan applied the dry–wet phase inversion technique in their development of the first CA membrane for seawater desalination, which is often called the Loeb–Sourirajan method. According to the Loeb–Sourirajan method, polymer solutions are prepared by mixing polymer, solvent, and sometimes even nonsolvent. The solution is then cast on a suitable surface with a doctor blade to a thickness of about 250 μm. After partial evaporation of the solvent, the cast polymer solution film is immersed in a bath of nonsolvent medium, often called coagulation (gelation) medium. Solidification of the polymer film takes place in two steps, that is, evaporation of the solvent and the solvent–nonsolvent exchange in the gelation bath. It is often preferable to choose a solvent of strong dissolving power with high volatility. During the first step of desolvation by solvent evaporation, a thin skin layer of solid polymer is formed instantly at the top of the cast film due to the loss of solvent. In the next solvent–nonsolvent exchange step, the nonsolvent diffuses into the polymer solution film, while the solvent diffuses out through the thin solid layer. The composition of the polymer solution changes during the solvent–nonsolvent exchange process.

As already mentioned, TFC membranes are fabricated by in situ polymerization. This method, developed by Cadotte and coworkers of Filmtec in the 1970s, is currently most widely used to prepare high-performance RO and nanofiltration (NF) membranes [8]. A thin selective layer is deposited on top of a porous substrate membrane by interfacial in situ polycondensation. There are a number of modifications of this method primarily based on the choice of the monomers [13]. However, for the matter of simplicity, the polycondensation procedure is described by a pair of diamine and diacid chloride monomers.

A diamine solution in water and a diacid chloride solution in hexane are prepared. A porous substrate membrane is then dipped into the aqueous solution of diamine. The pores at the top of the porous substrate membrane are filled with the aqueous solution in this process. The membrane is then immersed in the diacid chloride solution in hexane. Since water and hexane are not miscible, an interface is formed at the boundary of the two phases. Polycondensation of diamine and diacid chloride will take place at the interface, resulting in a very thin layer of PA. The preparation of composite membranes by the interfacial in situ polycondensation is schematically presented in Figure 3.4.
Reverse Osmosis Membrane

There are a number of combinations for the choice of diamine and acid chloride monomers. For example, if TMC, which has three –COCl groups in an aromatic ring, is mixed with phthaloyl chloride, which has two –COCl groups, cross-linking will form between two main chains. Unreacted –COCl will become –COOH upon contact with water and the membrane will become negatively charged. Monomers with reactive groups other than amine and acid chloride can also be used.

3.2.2 Membrane Surface Modification

Since both the surface chemistry and morphology of the membrane play a crucial role in determining the membrane performance, its enhancement has been attempted through modification of the membrane surface [14,15]. In fact, the latest researches on TFC membranes are mostly on the membrane surface modification as summarized [16]. To the authors’ knowledge, most of them remain research attempts and yet to be commercialized.

Soaking the freshly prepared TFC membranes in solutions containing various organic species, including glycerol, sodium lauryl sulfate (SLS), and salt of triethylamine and camphorsulfonic acid (TEACSA), can increase the membrane flux in RO applications by 30%–70% [17]. The physical properties of TFC (abrasion resistance) and flux stability can also be improved by applying an aqueous solution composed of polyvinyl alcohol (PVA) and a buffer solution as a posttreatment step during the preparation of the TFC membranes [18]. On the other hand, using PVA-based amine compounds having a side-chain amino group at the aqueous phase monomer instead of MPD can produce high-flux TFC membranes for low-pressure applications [19,20]. The addition of alcohols, ethers, sulfur-containing compounds, monohydric aromatic compounds, and more specifically dimethyl sulfoxide (DMSO), in the aqueous phase, can produce TFC membranes with an excellent performance [21–24].

It was also found that the water flux of TFC-RO membranes could be doubled without affecting the salt rejection by incorporating zeolite nanoparticles in the thin layer of the

![FIGURE 3.3](image-url) Representative chemical structures of the conventional TFC of (a) Dow, (b) DuPont, (c) Toray, and (d) Toyobo membranes.

![FIGURE 3.4](image-url) Schematic of in situ IFP.
TFC-RO membranes [25]. To prevent microbial fouling, a new type of antifouling membrane was developed by introducing TiO₂ nanoparticles to TFC membranes in order to reduce the loss of RO permeability [26,27].

Water permeability of PA-TFC-RO membranes could be increased using an oxygen plasma treatment by introducing carboxylic groups, which increases the hydrophilicity of the treated membrane. On the other hand, the chlorine resistance of the TFC-RO membrane can be enhanced using an argon plasma treatment that causes cross-linking to take place at the nitrogen sites on the membrane surface [28].

Since hydrophilic surfaces are believed to attract more water than organic contaminants of high hydrophobicity, many attempts have been made to increase membrane surface hydrophilicity by surface modification techniques, each having its own advantages and disadvantages. Hydrophilization by treating the membrane surface with water-soluble solvents (acids, alcohols, and mixtures of acids, alcohols, and water) is one of the surface modification techniques, as mentioned earlier. This method increases the flux without changing the chemical structure, but one of its disadvantages is that the water flux decreases with time because of the leaching of the hydrophilizing agent by water permeation [29]. Using a mixture of acid and alcohol in water for the surface treatment can improve the surface properties since acid and alcohol in water cause partial hydrolysis and skin modification, which produces a membrane with a higher flux and a higher rejection. It was suggested that the presence of hydrogen bonding on the membrane surface encourages the acid and water to react on these sites producing more charges [29,30]. Kulkarni et al. hydrophilized a TFC-RO membrane by using ethanol, 2-propanol, hydrofluoric acid, and hydrochloric acid [29]. They found that there was an increase in hydrophilicity, which led to a remarkable increase in water flux with no loss in ion rejection.

A hydrophilic, charged TFC could be produced by using radical grafting of two monomers, methacrylic acid and poly(ethylene glycol) (PEG) methacrylate, onto a commercial PA-TFC-RO membrane [31]. It was found that the use of amine containing ethylene glycol blocks enhanced the performance of the membrane and highly improved membrane water permeability by increasing hydrophilicity [32]. PEG and its derivatives have been used for surface modification. TFC membrane resistance to fouling could be improved by grafting PEG chains onto the TFC-RO membranes [33,34]. Louie et al. coated the surface of commercial TFC-RO membranes with a solution of polyether–PA (PEBAX 1657) to produce antifouling membranes [35].

Another alternative and less common approach for the membrane surface modification is the introduction of an active additive. The basis of this technique is the idea that these additives can move toward the top film surface during membrane formation and alter membrane surface chemistry, while keeping bulk properties unchanged. According to this method, only a very small quantity of the additives is enough to change the surface chemistry of the membranes [36]. Blending is a conventional technique used for membrane surface modification, and recently, much attention has been given to utilize this technique, in which hydrophobic surface-modifying macromolecules (SMMs) are blended to a base polymer for membrane surface modification [37]. Preparation of PA-TFC membranes for desalination using novel hydrophilic SMMs improved flux stability of PA-TFC-RO membrane by using hydrophilic SMMs formed simultaneously by in situ polymerization reaction when the polycondensation reaction takes place within the organic solvent of the TFC system [38,39].

### 3.2.3 Development of Membranes for Boron Removal, Chlorine Tolerance, and Antifouling

Boron occurs in seawater at an average concentration of 4–6 mg/L. Toxicological effects of exposure of the human body to boron, primarily by consumption of desalinated seawater, were also reported [40]. A provisional guideline of the WHO suggests a maximum boron concentration of 2.4 mg/L in drinking water (see, e.g., [41]). The California Department of Public Health (CDPH) set a notification level of 1.0 mg/L for the drinking water. It has been found that the RO membrane is not very effective in boron removal. According to the study for the two-stage RO desalination of Atlantic Ocean water under the conditions of feed pressure 802–966 psi, a product recovery of 38%–42%, for the boron concentration of 4.2–4.7 mg/L and with pH 7.0–8.2, was achieved. Boron removal in the first stage was 80.0%–83.4%, while in the second stage, it was 64.4%–66.0%. These low values of boron rejection cannot satisfy the WHO standard for the drinking water. The difficulty of boron removal arises from the fact that boron exists in seawater either as boric acid (H₃BO₃) or borate ion (H₂BO₃⁻) with their respective concentration depending on pH. Based on the first dissociation constant, pKₐ₁ = 8.7, boron is primarily present as boric acid in neutral seawater. It was found that the charged H₂BO₃⁻ is rejected at 97.3%–99.6% by commercial spiral-wound modules, while the rejection of H₂BO₃ is much lower (85.7%–96.7%) [42]. A recent paper published by Koseoglu et al. reported boron rejection of 85%–90% from seawater of pH = 8.2 (average seawater pH) and >98% at pH of 10.5 [43]. The experiments were carried out at 700 psig at ambient temperature using two commercial seawater reverse osmosis (SWRO) modules, that is, Toray™ UTC-80-AB and Filmtect™ SW30HR (Dow Chemical). SWRO membranes with better boron rejection are still being sought. Further improvements in boron removal by RO membranes can be found in the literatures [44,45].

There are still drawbacks in the commercial integrally skinned CA and PA-TFC-RO membranes. CA membranes are susceptible to microbiological attack, undergo compaction at high temperatures and pressures, and are workable only within a narrow range of pH (3–7). PA membranes, on the other hand, can tolerate a wider range of pH but suffer from poor resistance to continual exposure to oxidizing agents such as chlorine that is most widely used as an oxidizing biocide in water treatment. Namely, performance of...
PA membranes deteriorates at a continuous exposure to water containing more than a few parts per billion (ppb) of chlorine. A number of attempts have been made to change the macro-molecular structure of PA, as already mentioned in surface modification section, but the problem has not yet been solved completely.

Recently, an attempt was made to synthesize a sulfonated PS using a disulfonated monomer. Namely, a random copolymer is synthesized by direct polymerization of 3,3′-disulfonato-4,4′-dichlorodiphenyl sulfone (SDCDPS), 4,4′-dichlorodiphenyl sulfone (DCDPS), and 4,4′-biphenol (BP), or another copolymer is synthesized from SDCDPS, BP, and 4,4′-difluorotritylphenylphosphine oxide (DFIPPO) [46]. The reaction scheme is presented in Figure 3.5 [46]. In contrast to sulfonation of PS, sulfonated PS of a given ion exchange capacity (IEC) is obtained by this method in a more controlled manner. Using these polymers, RO membranes were prepared and their chlorine resistance was examined. One laboratory-made membrane (BPS-40H, based on the second type of synthesized copolymer with 40% sulfonic acid component in H form) was compared with SW30HR SWRO membrane from Dow with a feed solution containing 2000 ppm NaCl and 500 ppm chlorine at pH = 9.5. While the solute rejection of commercial RO membranes decreased by 20% within 10,000 ppm-hours of chlorine exposure, there was hardly any change in salt rejection in the laboratory-made membrane. This chapter indicates that there are still many things to be done in RO membrane material development, a research topic that has been neglected for many years.

3.2.4 Membrane Development with Low Pressure and High Flux

Kurihara summarized the improvement in RO membrane performance up to 1999 [47]. The membrane productivity in terms of water flux normalized by operating pressure has increased by an order of magnitude. This allowed the operation of RO for brackish water desalination and ultrapure-water production at pressures lower than 1 MPa, resulting in significant saving in energy cost. Kurihara attributed the high productivity of the newly developed membrane to an increase in surface roughness. Currently, there seems no limit in sight in further improvement of membrane productivity. Table 3.1 also shows similar advancement made for the Dow RO membranes [48].

3.2.5 Membrane Development for High Operating Pressure

The present seawater desalination technology aims to increase the pure water recovery by a membrane module from the conventional 40% to 60%. Since the osmotic pressure of the retentate will increase from 4.5 to 7.0 MPa when
the water recovery increases from 40% to 60%, the development of a high-pressure vessel, as well as the development of a membrane that will show little compaction under a high pressure, is necessary. Kawada reported the development of an RO membrane that was suitable for operation at 9 MPa [49]. The compaction of membrane usually takes place at the porous sublayer rather than at the skin layer. An attempt was therefore made to reduce the compaction by making a large number of uniform pores of small sizes at the surface of the porous sublayer on which an aromatic PA skin layer was coated by in situ IFP. The stability of the membrane module productivity increased significantly as compared to the conventional seawater desalination membrane.

3.3 REVERSE OSMOSIS TRANSPORT

3.3.1 SINGLE COMPONENT SYSTEM

Currently, the mainstream of RO membrane transport theory is the solution–diffusion model [50]. According to the model, mass transfer occurs in three steps: absorption to the membrane, diffusion through the membrane, and desorption from the membrane. The chemical potential gradient from the feed side of the membrane to the permeate side of the membrane is the driving force for the mass transfer. When the difference in hydrostatic pressure is greater than the difference in osmotic pressure between the upstream and downstream sides of the membrane, a chemical potential difference of water across the membrane drives water against the natural direction of water flow.

Thus, the water transport through the membrane can be described by

\[ N_A = L(\Delta p - \Delta \pi) \] (3.1)

where

- \( N_A \) is the water flux through the membrane (subscript \( A \) denotes water)
- \( L \) is the water permeability coefficient
- \( \Delta p \) is the transmembrane pressure difference
- \( \Delta \pi \) is the difference in osmotic pressure between the upstream and downstream sides of the membrane

The omotic pressure can be given by an equation similar to the van’t Hoff equation:

\[ \pi = i \phi CRT \] (3.2)

where

- \( i \) is the dissociation parameter, which is the number of ions produced by the dissociation of the salt
- \( \phi \) represents a correction factor
- \( C \) is the salt concentration
- \( R \) is the universal gas constant
- \( T \) is the absolute temperature

The permeability coefficient can be given by

\[ L = \frac{DSV}{RTl} \] (3.3)

where

- \( D \) is the diffusivity of water in the membrane
- \( S \) is the water solubility in the membrane
- \( V \) is the partial molar volume of water
- \( l \) is the thickness of the skin layer of the membrane

The osmotic pressure of seawater is typically 2300–2600 kPa and can be as high as 3500 kPa [51]. Osmotic pressures of brackish water are much smaller than those of seawater. Corresponding to the concentration range of 2000–5000 mg/L, the osmotic pressure ranges from 100 to 300 kPa [51]. To overcome these osmotic pressures, the pressures applied to RO seawater desalination are typically 6000–8000 kPa, and for the desalination of brackish water, they are 600–3000 kPa.

The recovery of product water is an important parameter of RO operation. It is defined as

\[ R_w = \frac{Q_p}{Q_f} \] (3.4)

where \( Q_p \) and \( Q_f \) are the volumetric flow rates of the permeate and feed streams, respectively. Typically, the recovery

---

**TABLE 3.1**

<table>
<thead>
<tr>
<th>Year</th>
<th>Production Capacity (gpd)</th>
<th>Salt Rejection (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Brackish water</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1990</td>
<td>8,000</td>
<td>98</td>
</tr>
<tr>
<td>1998</td>
<td>10,000</td>
<td>99.2</td>
</tr>
<tr>
<td>2007</td>
<td>11,000</td>
<td>99.8</td>
</tr>
<tr>
<td>Seawater</td>
<td></td>
<td></td>
</tr>
<tr>
<td>1990</td>
<td>4,000</td>
<td>99.4</td>
</tr>
<tr>
<td>1998</td>
<td>5,500</td>
<td>99.5</td>
</tr>
<tr>
<td>2007</td>
<td>7,500–8,000</td>
<td>99.8</td>
</tr>
</tbody>
</table>

of RO seawater desalination is 40%. When the recovery is increased, the salt concentration on the upstream side increases and more pressure should be applied to overcome the osmotic pressure. A special design of both membrane and membrane module is required to apply such high operating pressures. The problem of salt precipitation will arise as well. On the other hand, the recovery of RO brackish water desalination can reach as high as 85%.

Assuming the passages of water and salt through the membrane are independent from each other, the salt transport through the membrane occurs by the concentration difference between the upstream and the downstream sides of the membrane as the driving force. The salt transport through the membrane can be given by

\[ N_B = B(C_f - C_p) \]  \hspace{1cm} (3.5)

where
- \( N_B \) is the salt flux (subscript \( B \) denotes salt)
- \( B \) is the salt transport parameter
- \( C_f \) and \( C_p \) are the salt concentrations at the membrane surface on the feed side and in the permeate, respectively

\( B \) is further given by

\[ B = \frac{D_B K_B}{l} \]  \hspace{1cm} (3.6)

where
- \( D_B \) is the diffusivity of salt through the membrane
- \( K_B \) is the partition coefficient of salt between the solution and the membrane

A more important parameter to express the salt passage through the membrane is salt rejection given by

\[ R = \frac{C_f - C_p}{C_f} \times 100\% \]  \hspace{1cm} (3.7)

where
- \( C_f \) is the salt concentration in the bulk of the solution on the feed side
- \( R = 100\% \) means perfect salt rejection and \( R = 0 \) means no salt rejection by the membrane

It should be noted that \( C_f \) is the salt concentration at the membrane surface on the feed side that was used to calculate \( N_B \), while \( C_f \) is the salt concentration in the bulk of the solution on the feed side of the membrane. \( C_f \) and \( C_f \) may become different due to a phenomenon called concentration polarization. This occurs due to the rejection of the salt and its accumulation at the membrane surface. The concentration polarization is schematically illustrated in Figure 3.6.

By solving simple transport equations in the boundary layer at the membrane surface, the preceding two concentrations can be related to each other by the following equation:

\[ \frac{C_f - C_p}{C_f} = \exp \left( \frac{N_A}{c_A k} \right) \]  \hspace{1cm} (3.8)

where
- \( c_A \) is the molar concentration of water
- \( k \) is the mass transfer coefficient of the salt

To minimize the concentration polarization, that is, to make \( C_p \) as close to \( C_f \) as possible, \( k \) should be increased, which is enabled by increasing the turbulence of the feed solution. All the RO modules include devices to achieve this goal.

### 3.3.2 Multicomponent System

Seawater is a complicated mixture of many components. Hence, it is difficult to predict the membrane performance under different operating conditions based on transport theories. Most of the theories treating the separation of multicomponent electrolytic systems are based on the Debye–Hückel theory, Donnan effect, and Nernst–Planck equation [52,53]. Although they are applicable to the mixture of any number of ions involved in the feed, there are only few works in which
agreement between experimental results and theoretical predictions was tested for mixtures of more than four ions. One of the noted exceptions is the work of Rangarajan et al. [54]. These authors have expanded the transport theory to treat a multicomponent system involving nine ions of primary importance present in seawater. The separations of individual ions were predicted theoretically and the prediction was examined by experiments.

3.4 COMMERCIAL MODULES

The commercial RO modules that have appeared in the literature since 1977 are summarized in Table 3.2 together with the typical operating conditions and performance. Some of them might no longer exist in the market. From the table, it is clear that the majority of high-salt-rejection RO membranes are spiral-wound aromatic PA-TFC membranes with a noted exception of Toyobo that is based on CTA hollow fiber. The reason might be easier cleaning of the spiral-wound module, on one hand, and chlorine resistance of CA, on the other.

3.5 RECENT OPERATIONAL EXAMPLES OF LARGE-SCALE RO MODULES

Table 3.3 summarizes recent operational examples of large-scale RO modules.

The brief outline of each paper in Table 3.3 is as follows:

This study deals with the optimization of membrane filtration performance in the treatment of acid mine drainage (AMD) using two NF membranes (NF99 from Alfa Laval and DK from GE Osmonics) and one RO membrane (RO HR98PP from Alfa Laval) [55]. RO experiments were performed with a laboratory-scale membrane area of 63.6 cm², and pressure, pH, temperature, and flow rate were changed.

In 2007, two of the world’s largest wastewater reclamation plants using similar energy-saving RO membrane technology were commissioned [57]. One plant, Orange County Water District’s (OCWD) Groundwater Replenishment System (GWRS) located in Southern California, was expanding the plant to 265,000 m³/day (70 MGD), while the other, Ulu Pandan located in Singapore, was producing 148,000 m³/day (39 MGD). This paper discusses the evolution of the design of these large-scale wastewater reclamation plants using energy-saving RO membranes (PA-TFC membranes from Koch, Dow, and Hydranautics). Lessons learned from years of pilot, demonstration plants, and full-scale plant experience at both OCWD and UP were presented.

A case study of a natural gas production site covering various technical issues related to selection of an appropriate RO system is presented [59]. As part of the pretreatment selection, two types of ultrafiltration (UF) membrane modules, namely, spiral wound and hollow fiber, with molecular weight cutoff (MWCO) of 8,000 and 50,000 Da, respectively, were tested in parallel with NF membranes of the spiral-wound type with MWCO 200 Da. The NF plant with 50% capacity gave a recovery of 75%, and the RO plant gave a recovery of 60% in comparison to the expected 92%–95%. The long-term tests have indicated that the remainder of the membranes could be installed to achieve full capacity of the plant. This study also demonstrates the importance of the selection of a proper pretreatment setup for the RO system design.

The following three case studies of wastewater desalination are described [60]. (1) Effluent from a poultry and meat processing industry in southern Europe was treated by RO (Filmtec PA-TFC membrane). Pretreatment by UF and/or continuous MF (CMF) was made before NF treatment followed by RO. (2) Secondary municipal and agricultural effluent was treated by Filmtec BW30-365FR2 with BW30-330RO. (3) High-salinity brackish well water was treated by (SWRO) Filmtec SW30-8040. In each case, a cost estimation of the product water was made.

The performance and characteristics of new RO membranes operating at ultralow pressure (ULP) (7–10 bar) are presented [61]. The total cost comparison based on a 4 km³/day plant revealed that TFC ULP provides about 10% savings over the traditional TFC HR (traditional brackish water RO membrane). Generally, the use of TFC ULP contributes to energy saving, but the capital cost increases due to the higher membrane cost.

The study focuses on the behavior of the ULP RO membranes in the full-scale system [62]. The ULP membranes have salt rejections comparable to conventional RO membranes, but the hydraulic characteristics are significantly different. Hence, modifications to conventional membrane system design must be considered to optimize the use of the ULP RO membrane. Several design options are evaluated for function, effectiveness, and cost impact.

The growing success of membranes in municipal and industrial wastewater is related to improved process designs and improved membrane products [63]. Key factors that have been determined to result from successful operation of large-scale plants are discussed in this paper. Factors that play a key role in the use of RO membranes include ultra- or microfiltration pretreatment, low fouling membranes, flux rate, and recovery and control of fouling and scaling. In particular, high flux rates can be used when UF or MF pretreatment is used. These technologies remove most of the suspended particles that would normally cause heavy fouling of lead elements. It was demonstrated by a 2.5-year operation of the 10,000 m³/day Bedok, Singapore, demonstration plant.

The conventional hollow-fiber-type RO element is a single open-ended (SOE) structure, in which the pressure drop of the permeated water in the bore side of the hollow fibers prevents the development of a large-sized (longer) RO element [70]. In this work, a both open-ended (BOE) element was devised to reduce the permeate pressure drop. Medium-sized Toyobo RO module HJ9155 (CTA) was used for the test. It has been confirmed that the permeate flow rate of BOE was greater by about 30% than that of SOE.

Organic foulants obtained from UF (Kolon Membrane, Gwacheon, Korea), and RO (4-stage spiral-wound PA-type TFC-RO membrane system, Woongjin Chemical, Seoul, Korea) membranes of a large-scale municipal water...
# Table 3.2

## Typical Commercial Module Performance at Various Operating Conditions

<table>
<thead>
<tr>
<th>Manufacturer</th>
<th>Membrane Code</th>
<th>Module, Area (m²)</th>
<th>Surface Material</th>
<th>Operating Pressure (MPa)</th>
<th>Temp. (°C)</th>
<th>pH</th>
<th>Feed Conc. (mg/L)</th>
<th>PWP (m³/day)</th>
<th>Salt Rejection (%)</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>Alfa Laval, Sweden</td>
<td>HR98PP</td>
<td>Thin-film-composite (TFC) on polypropylene</td>
<td>16–40 bar</td>
<td>25</td>
<td>2–10</td>
<td>200 NaCl</td>
<td>39 L/m²h at 20 bar</td>
<td>&gt;96</td>
<td>[55]</td>
<td></td>
</tr>
<tr>
<td>Dow Chemical</td>
<td>XFS-4167.08</td>
<td>Aromatic polyamide (PA)</td>
<td>56 atm</td>
<td>35</td>
<td>4–7.5</td>
<td>290–470 TDS</td>
<td>9.4 (2,500 gpd)</td>
<td>98.5</td>
<td>[56]</td>
<td></td>
</tr>
<tr>
<td></td>
<td>BW30-400FR</td>
<td>PA reverse osmosis (RO)</td>
<td>225 psi</td>
<td>6</td>
<td>2,000 NaCl</td>
<td>39.7 m³/h (10,500 gpd)</td>
<td>99.5</td>
<td>[57]</td>
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<td>XLE-440</td>
<td>100 psi</td>
<td>6</td>
<td>500 NaCl</td>
<td>48.1 m³/h (12,700 gpd)</td>
<td>99.0</td>
<td>[57]</td>
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<tr>
<td>DuPont de Nemours</td>
<td>Permasep B-10</td>
<td>Aromatic PA</td>
<td>56 atm</td>
<td>35</td>
<td>5–9</td>
<td>290–470 TDS</td>
<td>5.5 (1,500 gpd)</td>
<td>98.5</td>
<td>[56]</td>
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<tr>
<td></td>
<td>Permasep B-10</td>
<td>Aramid</td>
<td>800 psi</td>
<td>25</td>
<td></td>
<td>30,000 NaCl</td>
<td>1.500 gpd</td>
<td>[58]</td>
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<td></td>
<td>Permasep B-10</td>
<td>Aramid</td>
<td>800 psi</td>
<td>25</td>
<td></td>
<td>30,000 NaCl</td>
<td>5,000 gpd</td>
<td>[58]</td>
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<td>Filmtec Corp.</td>
<td>BW30-4040</td>
<td>TFC</td>
<td>18 bar</td>
<td>30</td>
<td>6.81</td>
<td>33,030 TDS</td>
<td>5.8 m³/h (8,000 gpd)</td>
<td>99.3</td>
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<td></td>
<td>SW30-2540-2</td>
<td>TFC</td>
<td>60 bar</td>
<td>30</td>
<td>6.81</td>
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<td>0.14 L/m²h</td>
<td>[59]</td>
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<td></td>
<td>SW30-2540-2</td>
<td>TFC</td>
<td>62 bar</td>
<td>30</td>
<td>6.81</td>
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<td>0.5 L/m²h</td>
<td>[59]</td>
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<td></td>
<td>SW30-380</td>
<td>Thin-film composite (TFC)</td>
<td>61.0 bar</td>
<td>19.9</td>
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<td></td>
<td>105.8 m³/h</td>
<td>90.2</td>
<td>[61]</td>
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<tr>
<td></td>
<td>SW30-2540-2</td>
<td>PA</td>
<td>100–150 psi</td>
<td>4–11</td>
<td>1,300 TDS</td>
<td>103.3 m³/h</td>
<td>96.2</td>
<td>[61]</td>
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<td></td>
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<tr>
<td></td>
<td>SW30-2540-2</td>
<td>PA-TFC</td>
<td>1.55 (225 psi)</td>
<td>4–11</td>
<td>1,300 TDS</td>
<td>32.2 (8,500 gpd)</td>
<td>99.5</td>
<td>[62]</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>Hydranautics</td>
<td>PA-TFC</td>
<td>1.05 (150 psi)</td>
<td>500 NaCl</td>
<td></td>
<td>32.2 (8,500 gpd)</td>
<td>99.0</td>
<td>[62]</td>
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<td>8040-LSV-CPA2</td>
<td>PA-TFC</td>
<td>1.55 (225 psi)</td>
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<td>1,500 NaCl</td>
<td>32.2 (8,500 gpd)</td>
<td>99.5</td>
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<td></td>
<td>8040-UYH-ESPA</td>
<td>PA-TFC</td>
<td>1.05 (150 psi)</td>
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<td>1,500 NaCl</td>
<td>32.2 (8,500 gpd)</td>
<td>99.0</td>
<td>[62]</td>
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<td></td>
<td>LFC</td>
<td>PA</td>
<td>15.5 bar</td>
<td>25</td>
<td>6.8–7.0</td>
<td>1,500 NaCl</td>
<td>(2,380 gpd)</td>
<td>98.7–99.5</td>
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<td></td>
<td>Hydranautics</td>
<td>PA RO</td>
<td>225 psi</td>
<td>6</td>
<td>1,500 NaCl</td>
<td>41.6 m³/h (11,000 gpd)</td>
<td>99.5</td>
<td>[57]</td>
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<td></td>
<td>LFC3</td>
<td>225 psi</td>
<td>6</td>
<td>1,500 NaCl</td>
<td>36.0 m³/h (9,500 gpd)</td>
<td>99.7</td>
<td>[57]</td>
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<tr>
<td></td>
<td>ESPA2</td>
<td>150 psi</td>
<td>30</td>
<td>6</td>
<td>1,500 NaCl</td>
<td>34.1 m³/h (9,000 gpd)</td>
<td>99.6</td>
<td>[57]</td>
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</table>

(Continued)
### TABLE 3.2 (Continued)
Typical Commercial Module Performance at Various Operating Conditions

<table>
<thead>
<tr>
<th>Manufacturer</th>
<th>Membrane Code</th>
<th>Module, Area (m²)</th>
<th>Surface Material</th>
<th>Operating Pressure (MPa)</th>
<th>Temp. (°C)</th>
<th>pH</th>
<th>Feed Conc. (mg/L)</th>
<th>Test Conditions</th>
<th>Performances</th>
<th>Salt Rejection (%)</th>
<th>References</th>
</tr>
</thead>
<tbody>
<tr>
<td>Nitto Electric Ind. Co./Kobo Steel</td>
<td>NRO-A (first stage)</td>
<td>Tubular, 190 mmφ × 2674 mm L, 1.75 (tube) (12.5 mm ID × 18 pcs)</td>
<td>Cellulose diacetate, FPR (tube)</td>
<td>50 kg/cm²</td>
<td>25</td>
<td>5.5–6.0</td>
<td>35,000 NaCl</td>
<td></td>
<td></td>
<td></td>
<td>[65]</td>
</tr>
<tr>
<td>Nitto Electric Ind. Co.</td>
<td>NRO-A, NRO-B (first stage)</td>
<td>Tubular, 12:4 parallel 2–1 × 3 series</td>
<td>Blend of cellulose diacetate and CTA, FPR (tube)</td>
<td>50 kg/cm²</td>
<td>35</td>
<td></td>
<td>3,800 TDS</td>
<td></td>
<td></td>
<td>2.75</td>
<td>[56]</td>
</tr>
<tr>
<td></td>
<td>NRO-A, NRO-B (second stage)</td>
<td>Tubular, 3 numbers</td>
<td>Blend of cellulose diacetate and CTA</td>
<td>40 kg/cm²</td>
<td>35</td>
<td></td>
<td>250 TDS</td>
<td></td>
<td></td>
<td>1.7</td>
<td>[56]</td>
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<tr>
<td>Toray Co.</td>
<td>SC-5000A (first stage)</td>
<td>Spiral wound, 12: X-mas tree series 6–4–2</td>
<td>CA</td>
<td>56 kg/cm²</td>
<td>35</td>
<td></td>
<td>3,000 TDS</td>
<td></td>
<td></td>
<td>12.5</td>
<td>[56]</td>
</tr>
<tr>
<td></td>
<td>SC-5000B (second stage)</td>
<td>Spiral wound, 4 numbers</td>
<td>CA</td>
<td>40 kg/cm²</td>
<td>35</td>
<td></td>
<td>200 TDS</td>
<td></td>
<td></td>
<td>10</td>
<td>[56]</td>
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<tr>
<td></td>
<td>SC-5100A (first stage)</td>
<td>Spiral wound, 100 mmφ × 1.0 mL, 6.8</td>
<td>CA</td>
<td>56 kg/cm²</td>
<td>25</td>
<td>5.5–6.0</td>
<td>35,000 NaCl</td>
<td></td>
<td></td>
<td>2.0</td>
<td>[65]</td>
</tr>
<tr>
<td></td>
<td>SC-5100B (second stage)</td>
<td>Spiral wound, 100 mmφ × 1.0 mL, 6.8</td>
<td>CA</td>
<td>30 kg/cm²</td>
<td>25</td>
<td>5.5–6.0</td>
<td>2,000 NaCl</td>
<td></td>
<td></td>
<td>3.2</td>
<td>[65]</td>
</tr>
<tr>
<td></td>
<td>PEC-1000</td>
<td>Spiral wound</td>
<td>53.4–59.5 kg/cm²</td>
<td>11.8–23.8</td>
<td>6.4–6.9</td>
<td>18,400–19,800 Cl⁻</td>
<td>706–812</td>
<td>99</td>
<td>[66]</td>
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</tbody>
</table>

(Continued)
### TABLE 3.2 (Continued)

Typical Commercial Module Performance at Various Operating Conditions

<table>
<thead>
<tr>
<th>Manufacturer</th>
<th>Membrane Code</th>
<th>Module, Area (m²)</th>
<th>Surface Material</th>
<th>Test Conditions</th>
<th>Performances</th>
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</thead>
<tbody>
<tr>
<td></td>
<td></td>
<td></td>
<td></td>
<td>Operating Pressure (MPa)</td>
<td>Temp. (°C)</td>
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<tr>
<td>Toyobo Co.</td>
<td>SU-720</td>
<td>Spiral wound, low pressure</td>
<td>PA composite</td>
<td>1.5</td>
<td>25</td>
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<tr>
<td></td>
<td>SUL-G20</td>
<td>Spiral wound, ULP</td>
<td>PA composite</td>
<td>0.75</td>
<td>25</td>
</tr>
<tr>
<td></td>
<td>SUL-H20</td>
<td>Spiral wound, super ULP</td>
<td>PA composite</td>
<td>0.5</td>
<td>25</td>
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<tr>
<td>Hollosep</td>
<td>HR-5350</td>
<td>Hollow fiber, 5 parallel</td>
<td>CA</td>
<td>38 k/cm²</td>
<td>35</td>
</tr>
<tr>
<td></td>
<td>HR-5350S</td>
<td>Hollow fiber, 1 number</td>
<td>CA</td>
<td>70 k/cm²</td>
<td>35</td>
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<tr>
<td>Hollosep</td>
<td>HR 5350</td>
<td>Hollow fiber, 125 mmφ × 1.1 mL, 80</td>
<td>CA</td>
<td>55 k/cm²</td>
<td>25</td>
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<tr>
<td>Hollosep</td>
<td>HR 5350S</td>
<td>Hollow fiber, 125 mmφ × 1.1 mL, 80</td>
<td>CA</td>
<td>22 k/cm²</td>
<td>25</td>
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<tr>
<td>Hollosep</td>
<td>HR 8350</td>
<td>Hollow fiber, 305 mmφ × 1330 mm L</td>
<td>CA</td>
<td>75 k/cm² G</td>
<td>40</td>
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<tr>
<td>Hollosep</td>
<td>HR 8350S</td>
<td>Hollow fiber, 125 mmφ × 1.1 mL, 80</td>
<td>CA</td>
<td>70 k/cm² G</td>
<td>25</td>
</tr>
<tr>
<td>JM-12</td>
<td>Hollow fiber, 5 elements (HR1235ON-5)</td>
<td>CA</td>
<td>70 k/cm² G</td>
<td>25</td>
<td>3.500 NaCl</td>
</tr>
<tr>
<td>Hollosep</td>
<td>Hollow fiber, 8 in. module</td>
<td>CA</td>
<td>54.8–58.3 k/cm²</td>
<td>11.7–28.1</td>
<td>6.15–6.94</td>
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<tr>
<td>Hollosep</td>
<td>Hollow fiber, 12 in. module</td>
<td>CA</td>
<td>53.2–57.4 k/cm²</td>
<td>11.7–28.1</td>
<td>6.15–6.94</td>
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<tr>
<td>Universal Oil Products</td>
<td>ROGA-2B-TFC</td>
<td>Hollow fiber, 870 in. module</td>
<td>CTA TFC</td>
<td>5.5</td>
<td>25</td>
</tr>
<tr>
<td>Voontron Membrane Tech., Beijing</td>
<td>HOR21-8040</td>
<td>Hollow fiber, 33.9 (365 ft²)</td>
<td>Aromatic PA</td>
<td>4.14</td>
<td>25</td>
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<tr>
<td>Woongjin Chemical, Seoul</td>
<td>HOR21-4040</td>
<td>Hollow fiber, 7.9 (85 ft²)</td>
<td>Aromatic PA</td>
<td>4.14</td>
<td>25</td>
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<tr>
<td></td>
<td>HOR-2012</td>
<td>Spiral wound, 0.46 (5.0 ft²)</td>
<td>Aromatic PA</td>
<td>2.07</td>
<td>25</td>
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<td>RE8040-FL</td>
<td>Spiral wound, 37.2</td>
<td>PA-TFC</td>
<td>4.01–11.5 kgs/cm²</td>
<td>25</td>
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</table>

* TDS, total dissolved solid; TSS, total suspended solid; FRP, fiber-reinforced plastic.
TABLE 3.3
Examples of Large-Scale RO Applications

<table>
<thead>
<tr>
<th>Membrane and Module</th>
<th>Plant* (Elementb)</th>
<th>Water Type</th>
<th>Location</th>
<th>Other Remarks</th>
<th>References</th>
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<tbody>
<tr>
<td>Pretreatment and fouling</td>
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<tr>
<td>PA-TFC membrane</td>
<td>30 m³/day</td>
<td>Wastewater generated from natural gas production</td>
<td>Thailand</td>
<td>Selection of pretreatment (pretreatment)</td>
<td>[59]</td>
</tr>
<tr>
<td>PA-TFC</td>
<td>&gt;2, 1.05, and 14 km³/day</td>
<td>Meat processing effluent, municipal and agricultural effluent, agricultural runoff</td>
<td>Southern Europe, South America, Southern Spain</td>
<td>Pretreatment study and cost evaluation (pretreatment)</td>
<td>[60]</td>
</tr>
<tr>
<td>PA-TFC membrane</td>
<td>&gt;10 m³/h</td>
<td>Municipal wastewater</td>
<td>Bedok and Kranji, Singapore</td>
<td>MF and UF pretreatment (pretreatment)</td>
<td>[63]</td>
</tr>
<tr>
<td>PA-TFC</td>
<td>37.2 m²</td>
<td>Municipal wastewater</td>
<td>Gwangju, Korea</td>
<td>Characterization of organic foulant (fouling)</td>
<td>[71]</td>
</tr>
<tr>
<td>PA spiral wound</td>
<td>&gt;36 m³/h RO</td>
<td>River water</td>
<td>Steel industry Bremen, Germany</td>
<td>UF pretreatment of RO feedwater (pretreatment)</td>
<td>[72]</td>
</tr>
<tr>
<td>Filmtec BW 30 LE 440</td>
<td>&gt;1.56 km³/day</td>
<td>Surface water</td>
<td>Geleen (NL)</td>
<td>Pretreatment by UF (pretreatment)</td>
<td>[73]</td>
</tr>
<tr>
<td>Not specified</td>
<td>Various</td>
<td>Various</td>
<td>Various</td>
<td>Review, pretreatment (pretreatment)</td>
<td>[74]</td>
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<td>New membranes and modules</td>
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<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>PA spiral wound</td>
<td>&gt;24,000 m³/day</td>
<td>Brackish water</td>
<td>Hypothetical</td>
<td>ULP (100–150 psi) (membrane)</td>
<td>[61]</td>
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<tr>
<td>Fluid Systems TFC and Hydranautics TFC</td>
<td>&gt;1.9 km³/day</td>
<td>TDS 1000 and 5000 mg/L</td>
<td>Collier County, FL</td>
<td>System design for ULP RO membrane (membrane)</td>
<td>[62]</td>
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<tr>
<td>CTA hollow fiber</td>
<td>&gt;34 m³/day</td>
<td>Seawater</td>
<td>Arabian Gulf</td>
<td>Open-ended hollow fiber (membrane)</td>
<td>[70]</td>
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<td>16 and 8 in. spiral-wound element (Koch membrane)</td>
<td>&gt;180 mgd</td>
<td>Colorado river water</td>
<td>Metropolitan Water District of California</td>
<td>Cost evaluation of 16 and 8 in. RO element (cost)</td>
<td>[75]</td>
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<td>Energy</td>
<td></td>
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<tr>
<td>PA-TFC membrane</td>
<td>&gt;265 and 148 km³/day</td>
<td>Groundwater and wastewater</td>
<td>Orange County District, CA, and Ulu Pandan, Singapore</td>
<td>Use of energy-saving RO membranes (energy)</td>
<td>[57]</td>
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<tr>
<td>(Koch, Dow, and Hydranautics)</td>
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<tr>
<td>PA-TFC membrane</td>
<td>&gt;0.18 L/min (maximum)</td>
<td>Brackish water</td>
<td>Aqaba Gulf, deep well</td>
<td>Photovoltaic powered (energy)</td>
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<td>Review, solar-powered RO (energy)</td>
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<td>Cost evaluation</td>
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<tr>
<td>PA-TFC and zeolite membranes</td>
<td>&gt;50 m³/day</td>
<td>Wastewater produced from various sources</td>
<td>San Juan Basin</td>
<td>Cost evaluation (cost)</td>
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<td>Not specified</td>
<td>Not specified</td>
<td>NF/RO/MD hybrid system cost evaluation (hybrid)</td>
<td>[79]</td>
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<td>Miscellaneous</td>
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<tr>
<td>PA-TFC membrane</td>
<td>Small lab scale</td>
<td>AMD</td>
<td>—</td>
<td>Synthetic AMD (miscellaneous)</td>
<td>[55]</td>
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<td>(Alfa Laval)</td>
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<td></td>
<td>&gt;120 k + 30 km³/day</td>
<td>Seawater</td>
<td>Gaza Strip</td>
<td>Model development based on BBNs (miscellaneous)</td>
<td>[80]</td>
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<tr>
<td>PA composite, Dow BW30FR</td>
<td>1 km³/day</td>
<td>2000 ppm NaCl solution, fouled by BSA solution</td>
<td>—</td>
<td>Removal of the top skin layer for reuse (miscellaneous)</td>
<td>[81]</td>
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</table>

Reclamation plant were rigorously characterized using conventional and advanced characterization analyses (e.g., pyrolysis and mass spectrometry) in order to identify major constituents of the organic foulants and investigate fouling characteristics in a large-scale application of the UF and RO membranes [71]. Hydrophobic fractions comprising carboxylic acids and aldehydes strongly contributed to the fouling formation of the UF membrane, whereas the RO membrane foulants mainly consisted of hydrophilic fractions comprising amides and alcohols, indicating that the membrane characteristics could play an important role.
UF was used as a single-stage process to pretreat surface water as an alternative to conventional multistage treatment processes (e.g., ozonation–precipitation–flocculation–coagulation–chlorination–gravel filtration), in which the use of different chemicals necessitates special safety measures and careful harmonization and control of water chemistry in view of the requirements of downstream RO [72]. By contrast, processes based on membrane technology enable a simply designed plant to be used with several advantages. The pretreated water was supplied to the final stage of RO treatment to produce drinking water. Two UF modules MOLPURE FW50 technology, a four-end module produced by Daicen (cross-flow), and type S-225-X PVCUFC M5 from X-Flow (dead end) were used for comparison. The total operating costs were 0.25–0.35 EURO/m³ and lower than 0.5 EURO/m³ of the conventional pretreatment.

The water demineralization plant in Geleen (NL) for the production of boiler feedwater from surface water with a capacity of 1560 m³/h uses RO and UF for the pretreatment [73]. In this study, a series of pilot scale study was made using different UF membranes and a suitable UF membrane was selected. As for the RO membrane, Filmtec BW 30 LE 440 was chosen primarily due to its high-flux and high-salt-rejection capacity.

This paper reviews the recent representative research that is related to SWRO antifouling strategies and answers the most crucial questions about the design and operating parameters of SWRO and its pretreatment processes [74]. Also the economic evaluation of the SWRO system in regard to antifouling strategies is discussed. The pretreatment technologies to prevent membrane fouling and to extend the lifetime of the RO membrane are commonly grouped into two categories: conventional and nonconventional (membrane). After demonstrating many case studies, the authors concluded that the conventional pretreatment system applied in various SWRO plants in various parts of the world proved to be able to provide the feedwater quality required for the RO lines. However, the system was very difficult to control. On the other hand, membrane pretreatment provides a more stable and reliable system that can tolerate feedwater quality variations.

Three 16 in. diameter and 60 in. long RO-ULP elements from Koch Membrane Systems were evaluated parallel with two commercially available 8 in. diameter elements [75]. The overall specific flux for 16 in. was comparable to that of 8 in. element. Slightly higher fouling was observed for the former element. Cost evaluation revealed that for a 185 million gpd plant, 16 in. element can save as much as 12.4% in total cost. The capital cost saving alone is 27%.

An experimental study was conducted to investigate the potential of the development of water desalination using a photovoltaic powered system in Jordan [76]. A testing rig was built, where an RO (PA-TFC) desalination system is driven by photovoltaic power by directly coupling the photovoltaic powered system to a DC motor, which was coupled to a pump that was capable of providing sufficient torque to run the RO system. Analysis of results shows that a gain of 25% and 15% of electrical power and pure water flow, respectively, could be achieved using the east–west one-axis tracking system compared with fixed flat plate.

The paper aims to provide a comprehensive review of all the indirect solar desalination technologies along with plant-specific technical details [77]. Solar desalination plants are summarized since the year 1978 when solar RO desalination started. The factors affecting PV-RO water cost are capital cost of PV array and battery, inclusion of energy recovery device, type of feedwater, and type of RO unit. A water cost of around $7 m⁻³ for a 44 m³/day is estimated, which is very high. For large-scale plants having a capacity of >1000 m³/day, however, the specific plant cost is in the range of $4500–$6200 m⁻³/day with an estimated water cost of <$2 m⁻³ owing to the reduction of high-efficiency PV module cost.

RO membranes including polymeric (PA-TFC membrane) and molecular sieve zeolite membranes were investigated for ion removal from the water produced at oil field and coal bed methane sites by a cross-flow RO process [78]. Pretreatments including NF and adsorption by active carbon were implemented. The study revealed that (1) most of permeation tests lasted only 3 months due to severe fouling, (2) multistage pretreatment is crucial to extend membrane life, and (3) only NF treatment could extend the membrane life to 6 months.

In a comparative study of an integrated hybrid membrane-based system with an earlier locally designed RO unit, such system comprises NF, RO, and membrane distillation (MD) subsystems [79]. The comparison is essentially based on using the NF technique in pretreatment section, while the MD was contributed to concentrate the two brine streams from both NF and RO. Thus, high recovery rate of product water is aimed for. The proposed system was economically evaluated and compared with the RO unit. It was concluded that 76.2% water recovery was possible with a water production cost of $0.92 m⁻³.

Characterization of uncertainties in the operation and economics of the proposed seawater desalination plant in the Gaza Strip was made by using a Bayesian belief network (BBN) approach [80]. In particular, the model was used to (1) characterize the different uncertainties involved in the RO process, (2) optimize the RO process reliability and cost, and (3) study how uncertainty in unit capital cost, unit operation and maintenance (O&M) cost, and permeate quality was related to different input variables. The minimum specific capital cost was found to be 0.224 ± 0.064 US$/m³, and the minimum O&M cost was found to be 0.59 ± 0.11 US$/m². This unit cost was for a production capacity of 140,000 m³/day.

This study aimed to assess the technical feasibility of removing the dense PA active layer of RO membranes, with the intent to reuse degraded RO into porous low-pressure membranes [81]. The study assessed the ability of three degrading solutions (NaOH, KMnO₄, and NaOCl) to remove the active layer. The most promising results were found using NaOCl. Membranes treated with at least 300,000 ppm of NaOCl presented an increased permeability to 175 ± 4 L/m²h/bar with <4% salt rejection. The fouling behavior of the degraded RO membranes was also compared to commercially available UF membranes, to which they displayed similar
fouling characteristics, and by using LC-OCD, the MWCO of the reused membranes was estimated to be 5–10 kDa.

From this review, it is quite obvious that the most serious concern over the large-scale plant operation is pretreatment and fouling prevention. The membrane pretreatment seems more effective and economical than the conventional pretreatment. The recent membrane module development is focused on ULP RO membranes and the construction of a 16 in. diameter module, both generating economic benefits. But care must be taken for the large-diameter module operation. As energy cost keeps increasing, search for alternative energies is continued. In the arid regions of the Middle East, use of solar power has been attempted for quite some time to construct medium-sized desalination plants. However, the production of water seems quite costly due to the high cost of photovoltaic power. But the price may go down as the size of the plant increases. Last but not least important, almost all of the articles deal with the cost of water production.

3.6 FUTURE DIRECTIONS OF THE RESEARCH AGENDA

Improvement of membrane performance to achieve higher flux, higher salt rejection, and low energy consumption simultaneously is still the subject of greatest interest. As is obvious from this review, enhancement of membrane flux, while maintaining high salt rejection, has been achieved primarily by the major industrial RO membrane manufacturers by decreasing the skin layer of the PA-TFC membrane. On the other hand, many attempts have been made, mostly in the academia, to increase the flux by incorporating nanoscale hydrophilic particles. Another interesting attempt is to replace the conventional support membrane for in situ polymerization with a highly porous material such as nanofibrous membrane [82]. In all of these attempts, the approach is based on the improvement of the PA-TFC membrane due to the latter membrane’s enormous commercial success. But there may be polymeric materials other than PA to form the thin surface layer in situ. Moreover, membranes of overwhelmingly high fluxes may come from totally different approaches, such as the use of recently developed aquaporin or carbon nanotubes [83].

Fouling continues to be of great concern for the practical application of RO membranes, and the development of intrinsically fouling-free membranes is the goal of many researchers [16]. Surface modification is the most popular approach to reduce fouling. In this regard, one of the unconventional approaches of blending SMMs to the host polymer matrix should be further explored [36–39]. Despite the simplicity of its process, that is, the requirement of only a small amount of SMM in one step to modify the membrane surface, the method has not yet fully taken off, since amphiphilic macro-molecules of rather exotic structures have so far been used. The SMMs should be found among the bulk chemicals that are easily accessible to many laboratories with low cost.

As the modern characterization method advances, the details of the membrane morphology are being revealed. The size and size distribution of the water channels can be measured precisely by the modern instrument, which makes the water and solute transport theories based on the black box approach look more obsolete. A new transport model that matches the advancement of the modern characterization method is called for. As well, the precise control of the water channel size and its distribution, as well as the surface roughness, is required for the future membrane design.

Regarding the module development, the module size continues to become larger along with the increase of the plant size, and currently, there seems no limit for this trend. As the module size increases, so also do the sizes of pumps, pipes, etc., and more precise system control will be required.

Hybrid processes are also being considered more seriously than ever [84]. There are hybrid processes, such as pretreatment by UF and NF to provide RO feedwater and RO treatment of boiler water for the distillation process, and those are already in practice. Combining RO with the emerging membrane separation processes such as MD and forward osmosis (FO) will become more important to decrease the amount of effluent from the RO desalination plant. As drinking water production by the desalination of seawater and brackish water increases, increase in the quantity of salty RO effluent will become more of environmental concern. The RO hybrid process with MD and FO may become the answer to solve the problem.

3.7 CONCLUSIONS

Many new polymers have been synthesized and tested for their permeation properties, specifically fouling resistance, chlorine resistance, thermal and mechanical stabilities, etc., aiming at improvement in membrane performance for RO applications. These efforts seem to continue with insightful vision and a strong commitment for the future. However, only a handful of polymers are currently being used as materials for commercial applications, and they are not necessarily the polymers with the best performance properties. This is mainly due to the cost factor that governs the present membrane market. The RO performance of membranes, on the other hand, is known to be primarily ruled by the membrane’s active surface. The surface modification and surface coating will be focused on the new membrane development techniques in future, as only small amount of polymer is required for the surface coating. The knowledge of the effects of these factors on the RO performance will allow for full utilization of the potential of polymers for membrane materials, which will open up new promising avenues for further research and development of the areas.

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