Chapter

# Energy Recovery in Membrane Process

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## Abstract

One way in order to reduction energy consumption and providing the required water in both well-established technologies such as reverse osmosis (RO) and electrodialysis is use of the strengths of two or more processes through hybridization. Other key objectives of hybridization include increasing the capacity of the plant flexibility in operation and meeting the specific requirements for water quality. At this section, has been provided a critical review of hybrid desalination systems, and methods used to optimize such systems with respect to these objectives. For instance, coupling two process like as electrodialysis with RO is very effective in order to overcome the low recovery in RO systems. On the other hand, we can use for two or more processes such as RO with membrane distillation (MD) or zero liquid discharge (ZLD) for treatment of hypersaline feed solutions. At this section, also have been reviewed the applicability of salinity gradient power technologies with desalination systems and we identified the gaps that for effective upscaling and execution and implementation of such hybrid systems need to be addressed.

**Keywords:** energy recovery, desalination, hybrid systems, reverse osmosis, membrane

## 1. Introduction

Sustainable energy is the key solution for addressing major concerns about the future such as climate change, environmental protection, and balanced growth of the economy and society. In many nations at past two decades have witnessed advancement in economic development. However, industrial advancement, deterioration of the environment, energy shortage, the rapid economic growth and increasing demands of growing populations pose a huge threat for future generations [1–3]. For many years, economic development has been the key focus of many policy makers in sustainable development until the inception of the Kyoto protocol agreement in 1997, which includes environmental quality as a crucial variable for sustainable development [3]. According to global energy consumption, expected that electricity demands to be double in the next twenty-five years, so, major opportunities for innovation in energy production, storage, transmission and use of it have begun to open up. In particular, in order to improving the efficiency of the

processes and reducing the global carbon footprint, there is a huge interest in sustainable energy technologies [3, 4].

Development of an approach to sustainable energy that addresses greenhouse gas emission, environmental concerns, availability of resources, social impact and cost is an immense challenge. The key focus for obtaining energy sustainability is the generation of energy with renewable energy sources and replace them slowly with power fossil fuels [5]. There is much research that has worked for developing the membrane sector, which emphasizes the use of renewable energy in membrane technology. Although the efficiency of the process is still a high priority. Recently, membrane technologies, especially, in the water and energy sector, have begun to play a basic role in developing the infrastructure for sustainable energy. Some of the membrane-based approaches that are currently adapted at an industrial scale include desalination by RO, membrane-based bioreactors (MBR) for pure water generation, lithium-ion batteries, and membrane-based fuel cells and $CO_2$ capture [6–8]. Many advantages of membrane technologies like flexibility, feasibility and adaptability have been able to decrease many concerns related to water scarcity and energy demands in recent years. However, with achievement to advancements in membrane-based technologies. we still need to improve affordability and costs.

## 2. Membrane technology and sustainable water generation

In the past decades, following the increase in freshwater demand, various techniques including multiple-effect distillation (MED), vacuum distillation, multi-stage flash distillation (MSF), and other membrane-based technologies, such as reverse membrane distillation (MD), osmosis (RO) and etc., in order to sea water desalination, have been developed. Among these technologies, some of the membrane-based techniques such as RO, MD and forward osmosis (FO), because of some advantages like as lower maintenance and operating costs, lower capital requirements and low energy consumption, are considered as suitable alternatives [3, 9].

### 2.1 Desalination

Desalination is a process which use for producing freshwater from either sea or brackish water, by removing the salt content either by membrane technologies or by a thermal distillation process.

As can be seen from **Table 1**. the membrane technologies, specifically the RO, mainly, because of lower energy requirements, are preferred over the other technologies. In different technologies, the specific energy consumption (SEC) varies widely and depending on the operation and process control as well as the quality of the produced water, this value might have further differed significantly for a particular technology.

### 2.2 Reverse osmosis (RO)

To date, for desalination and stress reduction due to depletion of available water resources, reverse osmosis (RO) is the key technology [1]. In desalination plant such as RO, membrane played a key role which is largely determine the separation performance of the overall plant (**Figure 1**). In several recent studied suggests that in ultra-permeable membranes (UPMs) by increasing the water permeability up to three times than normal could reduce the energy consumption pressure vessels for seawater desalination about 15% and 44%, respectively.

| Technology | Specific energy consumption (kWh/m$^3$) | | |
|---|---|---|---|
| | Electric thermal | Thermal | Total electric equivalent |
| ED | 1–3.5 | — | 1–3.5 |
| EDR | 1–2 | — | 1–2 |
| SWRO | 3–6 | — | 3–6 |
| BWRO | 0.5–3 | — | 0.5–3 |
| MVC | 7–15 | — | 7–15 |
| MD | 1.5–4 | 4–40 | 3–22 |
| FO | 0.2–0.5 | 20–150 | 10–68 |

*ED = electrodialysis; EDR = electrodialysis reversal; BWRO = brackish water reverse osmosis; SWRO = seawater reverse osmosis; MVC = mechanical vapor compression; MD = membrane distillation; MSF = multi-stage flash; MED = multiple effect distillation; MEB = multi-effect boiling; FO = forward osmosis.*

**Table 1.**
*Specific energy consumption (SEC) by different desalination techniques.*

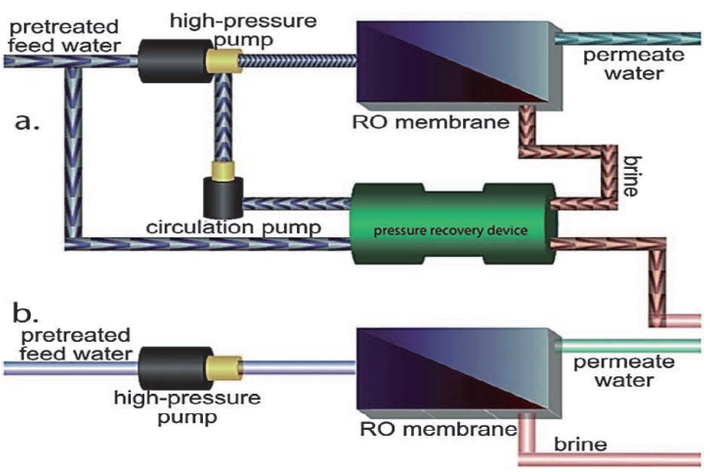


**Figure 1.**
*The RO process diagram with (a) and without (b) pressure recovery for SWRO and BWRO respectively.*

In the context of wastewater reclamation, even greater savings (e.g., 45% less energy input and 63% fewer pressure vessels [10]) can be achieved. Moreover, increasing the properties of the membrane selectivity can cause improvement the quality of the product [11].

Recent studies introduce the promise of developing new membrane materials. These materials can desalinate water while showing far greater permeability than traditional reverse osmosis (RO) membranes. But the question remains whether higher permeability means significant reductions in the cost of desalinated water. Research evaluates the potential of ultra-permeable membranes (UPM) to improve the performance and cost of RO.

### *2.2.1 Ultra-permeable membranes (UPM)*

By modeling the mass transport inside a reverse osmosis pressure vessel (PV), the study assesses how much tripling water permeability lowers energy consumption. And also lowers the number of required pressure vessels for a particular desalination plant. The findings were very interesting, it proved that a tripling (3×)

in permeability permits 44% fewer pressure vessels and 15% less energy for a seawater reverse osmosis plant (SWRO) [10, 12]. This is done Â at a both given capacity and recovery ratio. Moreover, tripling permeability results in 63% fewer pressure vessels or 46% less energy for brackish water reverse osmosis (BWRO). However, it also shows that the energy savings of ultra-permeable membranes (UPM) exhibits a law of diminishing returns due to thermodynamics and concentration polarization at the membrane surface [10].

In terms of reducing energy consumption, the benefits of ultra-permeable membranes (UPM) are limited to approximately 15% in the case of SWRO. It also shows that membranes with 3× higher permeability reduces number of pressure vessels by 44% for seawater reverse osmosis RO plants SWRO. And 63% in brackish water RO plants BWRO. This does not affect the energy consumption or permeate recovery [13].

In order to calculation of systems-level quantities the typical RO process diagram that shown in **Figure 2**, is used. In SWRO systems, for pressurizing the feed using mechanical energy Regenerated force from isobaric brine, pressure recovery devices (PRDs) are used (**Figure 2a**), while at BWRO typically this is not done (**Figure 2b**).

In case of energy consumption, ultra-permeable membranes proved to lower energy consumption of seawater reverse osmosis systems—SWRO—by %15. While on the other hand lowered energy consumption of brackish water reverse osmosis systems—BWRO—by 46%. The research was made at the same permeate flow per pressure vessel as what is typical nowadays. As can be shown in **Figure 2a** by reducing the inlet pressure, lower energy consumption (membrane area, feed flowrate and for a given recovery ratio) would be obtained. In SWRO (the line with purple dye in the figure), the pressure of inlet feed reduces to the outlet of the brine osmotic pressure. This limitation in the membrane, that corresponds to the osmotic pressure of the brine, is independent from membrane performance. As can be seen in the **Figure 2a**, with increasing $A_m$ up to triple from 1 to 3 L ($m^2$ h bar), we can reduce the inlet pressure about 1% and reach from 70 bar to 63 bar. For every 1% reduction in the inlet pressure, the SEC could be reduced up to 1.5%. However, as can be seen in this figure, any further improvements in membrane permeability beyond 3 L ($m^2$ h bar)$^{-1}$, since 63 bar is already within 1% of the osmotic limit for SWRO at the chosen recovery ratio, would have essentially no effect on energy consumption.

As can be shown in **Figure 2a**, in order to achieve 65% recovery in BWRO and with increasing $A_m$, inlet pressure rapidly drops. Due to the limitation of the osmotic

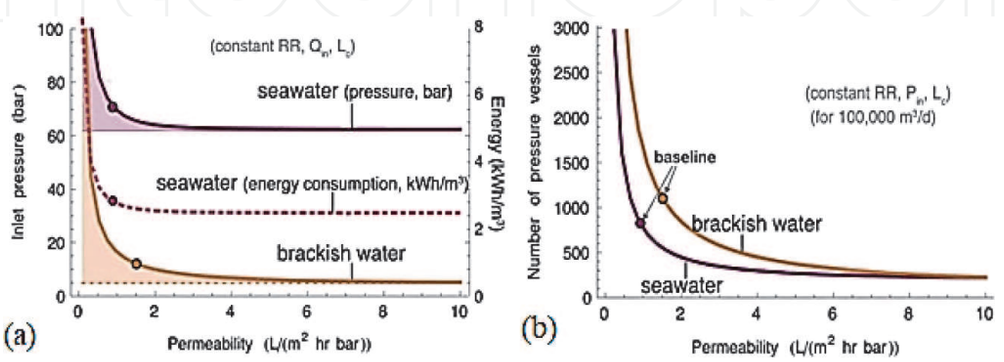


**Figure 2.**
*Investigation of key performance criteria and their effect in membrane permeability for BWRO at 2000 ppm NaCl (orange) and SWRO at 42,000 ppm NaCl (purple). (a) Energy consumption (dashed) and minimum required inlet pressure (solid lines) at fixed feed flowrate and recovery. In BWRO, energy consumption and pressure are linearly related. (b) Number of pressure vessels required for a total capacity of 100,000 $m^3$ $day^{-1}$ at fixed recovery ratio and pressure. Membrane width is held fixed in both subplots.*

for BWRO that is only a fraction of that of SWRO, with the increase in membrane permeability up to triple, it is causing a much greater reduction in inlet pressure, namely down to 6.4 bar from 12 bar in the case of thin-film composite (TFC) membranes (a 46% reduction in pressure and energy consumption). In the membranes with more permeability, with increasing the membrane's water permeability ($A_m$) (L ($m^2$ h bar)$^{-1}$) to over 5 L ($m^2$ h bar)$^{-1}$, the pressure essentially reaches the asymptotic limit. So, for the RO plant in the stage of brackish water, the UPMs could reduce the energy consumption to half. In the BWRO, the number of pressure vessels is lower than the SWRO. On the basis of **Figure 2b**, with a tripling $A_m$, we can reduce the pressure vessels up to 63%, for a given plant capacity, by increasing the feed flowrate per vessel from 139 $m^3$ $day^{-1}$ to 378 $m^3$ $day^{-1}$. Furthermore, increases in feed flowrate have no effect on the energy, since, viscous losses in a BWRO system represent a negligible component of the overall energy consumption [3].

Commercial RO membranes are dominated by TFC polyamide and its derivatives **Figure 3**. These membranes are facing critical challenges such as low selectivity, relatively low water permeability and high fouling tendency [2]. For example, in RO membranes, TFC has a typical water permeability range from ~1–2 L $m^{-2}$ $h^{-1}$ $bar^{-1}$ for SWRO membranes and ~ 2–8 L $m^{-2}$ $h^{-1}$ $bar^{-1}$ for BWRO [10, 14]. So, in synthesizing novel RO membranes, focused on the improvement of separation properties and better antifouling performance that is a key research focus in the field of desalination.

When it comes to capital costs, on the basis of our analysis, we can propose certain qualitative trends. According to Global Water Intelligence, in a typical SWRO plant with capacity of 150,000 $m^3$ $day^{-1}$, the levelized capital cost today is about 0.20 $ per $m^3$ (excluding land) that 20% of this cost is due to piping,, pressure vessels and membranes [15, 16]. So, with using of UPMs membrane, in a surface area similar to conventional membranes but with triple permeability, membranes can be reduced by up to 44%, in this situation the membranes would save on the order of 0.02 $ per $m^3$ in capital costs. The benefits are more significant for BWRO. in BWRO systems with UPMs membrane, saw that reduction of the energy consumption could be up to 46% [8]. Following increase of membrane permeability mass transfer coefficients and also typical cross-flow velocities decrease. With enhancement of membrane permeability, permeate water flux increases routinely [10].

The consequences of producing a product with less working pressure or more permeability can be estimated with confidence. As described above, the energy savings in SWRO with UPMs membrane could be limited to about 15%. At SWRO plants, because of the high salinity of seawater, operation has been optimized in such a way that these plants work with minimum pressure (60–70 bar) in order to extract permeate water from seawater [8, 10]. The difference between pre- and

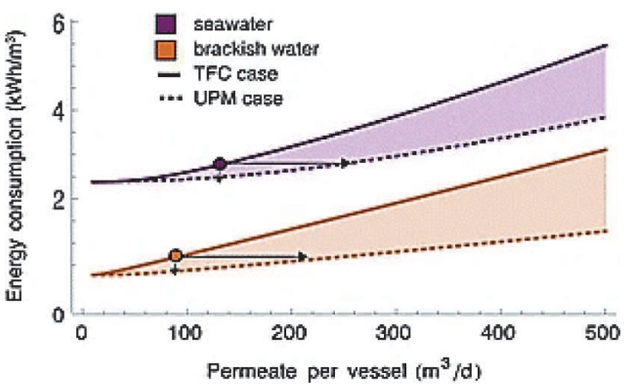


**Figure 3.**
*Ultra-permeable membranes UPM thin film composite TFC for BWRO and SWRO.*

post-treatment is about ~1 kWh $m^{-3}$, in RO stage, a 15% reduction in the energy consumption could only reduce ~10% of the overall cost of the energy in SWRO plants. With the reducing of the total energy consumption in SWRO plants from 3.8 kWh to 3.5 kWh, If the price of electricity is assumed to be 0.10 $ per kWh, could be saved the cost about 0.03 $ per $m^3$ [17, 18].

Wilf [19] evaluated with replacing the RO elements with membranes which have 80% higher permeability, in situation which recovery ratio and feed salinity was 85% and 1500 ppm, respectively, the SEC of BWRO decrease. He found that in two different averages flux (25.5 LMH and 34 LMH) the SEC was decreased (from 0.52 to 0.40 kWh/$m^3$ and from 0.72 to 0.49 kWh/$m^3$, respectively).

Franks et al. [20] evaluated, in BWRO plants, when a membrane element with 34.1 $m^3$/d of permeate flow replace with another elements that has 45.4 $m^3$/d of permeate flow, the SEC decrease. In this study, with decreasing the feed pump pressure 9.8–8.3 bar, the specific energy consumption decreased from 0.41 to 0.35 kWh/$m^3$ (the pump efficiency was 83%, the recovery ratio was 85% and the feed salinity was 1167 ppm (for wastewater). The simulation conditions were shown in **Tables 2** and **3**.

For a BWRO plant, Werber et al. [24] assumed a 85% recovery rate and feed with NaCl concentration about 5844 ppm. They observed, in a single-stage process, with increasing the water permeability in membrane from 4 to 10 LMH/bar, the SEC can be reduced up to 2.2%. On the other hand, in this study observed that in a two-stage RO with membrane permeability of about 4 LMH/bar, the required energy was 22% lower (0.11 kWh/$m^3$) than the single-stage RO, also the SEC decreased by increasing the membrane permeability from 4 to 10 LMH/bar by 12% (0.05 kWh/$m^3$) that compare to a single-stage BWRO was slightly larger. In this study, in SWRO with single stage process and membrane permeability about 2 LMH/bar, the hydraulic pressure was only 7.6% above the brine osmotic pressure (**Figures 4** and **5**). The results of their findings of the relationship between membrane water permeability and the SEC have shown.

Busch et al. [29] assessed the CAPEX and OPEX reductions with higher permeable SWRO elements. They compared the energy use, power cost, water cost by replacing SW30HR-380 with 28.4 $m^3$/d of permeate flow rate and 99.75% of NaCl rejection rate by SW30HR LE-400 with 34.1 $m^3$/d of permeate flow rate and 99.70% of NaCl rejection rate using the test results for each element. Test conditions and calculation assumptions were 32,000 mg/L NaCl of feed concentration, 8% of recovery rate, 55 bar of feed pump pressure, 5 years of operating time, 20% of RO membrane elements replacement rate per year, 90% of pump efficiency, and 0.08 US$/kWh of power cost. The pretreatment, chemical cleaning, and other costs were not considered. They indicated that decreasing membrane area by using higher water permeability RO elements can decrease the water cost by 4.7% from 0.190 to 0.181 US $/$m^3$ with the same energy cost.

For SWRO, the energy cost contributes 40–50% of the total water production cost; therefore, the ratio of the specific membrane cost to the total water production cost is about 1.2–6%. Hence, doubling the membrane water permeability halves the specific membrane cost so that the total water production cost is reduced to 0.6–3%. When the cost of pressure vessels is taken into consideration, the decrease of total water production cost is 0.7–3.5% [12]. But, with increasing the membrane permeability, the feed velocity and the pressure loss increase, as a result, more energy is needed, these could increase the SEC up to 6%.

As can be shown in **Figure 6**, Cohen-Tanugi et al. [10] calculated the total number of pressure vessels needed in a single-stage SWRO and BWRO with 100,000 $m^3$/d permeate and 42,000 ppm and 2000 ppm salinity concentration, respectively.

| Condition | | | | | | | Author (year) | Reference |
|---|---|---|---|---|---|---|---|---|
| **Feed concentration** | **Recovery rate** | **Average system flux/average TMP** | **No. of elements per vessel** | **Salt rejection** | **Pump efficiency** | **ERD efficiency** | | |
| **ppm or mg/L** | **%** | **LMH/bar** | **—** | **%** | **%** | **%** | | |
| 35,000 mg/L TDS | 50 | –/15.5 | N.D. | 99 | 100 | 100 | Zhu et al. (2009) | [21] |
| 42,000 ppm NaCl | 42 | 16/– | 8 | 99.8 | 75 | 97 | Cohen-Tanugi et al. (2014) | [10] |
| 32,000 ppm NaCl | 50 | N.D | N.D. | N.D. | 85 | 95 | Shrivastava et al. (2015) | [22] |
| 30,000 mg/L NaCl ($\pi_F$ = 25.6 bar) | 50 | 15/– | Not considered | 100 | 100 | 100 | McGovern et al. (2016) | [23] |
| 35,000 mg/L NaCl | 50 | 15/– | N.D | *B*-value was used | 100 | 100 | Werber et al. (2016) | [24] |
| 35,000 mg/L NaCl | 50 | 22.9/– | 8 | 100 | 100 | 100 | Mazlan et al. (2016) | [25] |
| 35,000 mg/L NaCl | N.D. | N.D. | N.D | 100 | N.D | N.D | Shi et al. (2017) | [26] |
| 35,000 ppm NaCl | 70 | 15/– | 8 | 100 | 100 | 100 | Wei et al. (2017) | [27] |
| 40,000 ppm | 50 | N.D. | N.D | N.D | 85 | 95 | Karabelas et al. (2018) | [28] |

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**Table 2.**
*Simulation conditions of each reference that includes the relationship between membrane water permeability and SEC for SWRO [12].*

| Condition | | | | | | | Author (year) | Reference |
|---|---|---|---|---|---|---|---|---|
| **Feed concentration** | **Recovery rate** | **Average system flux/average TMP** | **No. of elements per vessel** | **Salt rejection** | **Pump efficiency** | **ERD efficiency** | | |
| **ppm or mg/L** | **%** | **LMH/bar** | **—** | **%** | **%** | **%** | | |
| 3500 mg/L TDS | 50 | –/1.55 | N.D | 99 | 100 | 100 | Zhu et al. (2009) | [21] |
| 2000 ppm NaCl | 65 | 13.2/– | 7 | 99.8 | 75 | 97 | Cohen-Tanugi et al. (2014) | [10] |
| 804 mg/L TDS | 85 | N.D. | N.D | N.D | 85 | 95 | Shrivastava et al. (2015) | [22] |
| 5844 mg/L NaCl | 75 | 15/– | N.D | *B*-value was used | 100 | 100 | Werber et al. (2016) | [24] |
| 1000 mg/L NaCl | N.D. | N.D. | N.D | 100 | N.D | N.D | Shi et al. (2017) | [26] |
| 3000 ppm NaCl | 60–98 | 15 | 8 | 100 | 100 | 100 | Wei et al. (2017) | [27] |
| 2000 ppm | 70 | N.D | N.D | N.D | 85 | 95 | Karabelas et al. (2018) | [28] |



**Table 3.**
*Simulation conditions of each reference that includes the relationship between membrane water permeability and SEC for BWRO [12].*

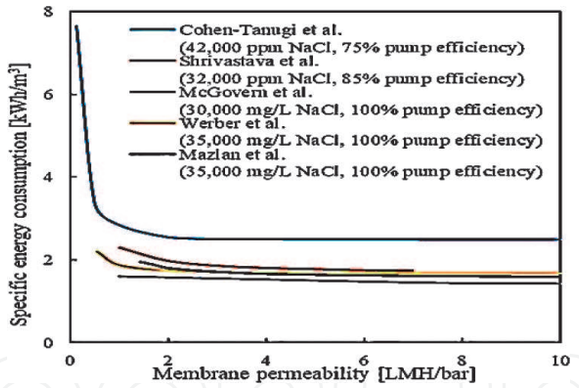


**Figure 4.**
*Calculated specific energy consumption as a function of membrane water permeability for single-stage SWRO from several references at different conditions.*

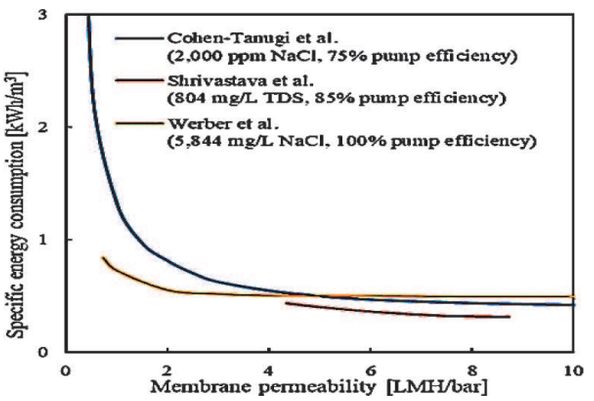


**Figure 5.**
*Calculated SEC as a function of membrane permeability for BWRO from several references.*

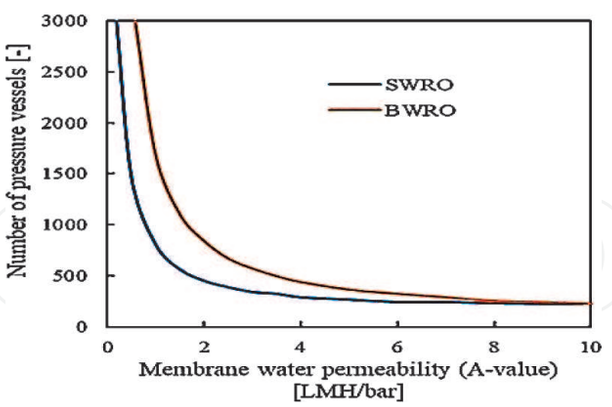


**Figure 6.**
*Calculation of the required number of pressure vessels in order to investigation of function of the membrane water permeability for single-stage BWRO with12 bar feed pump pressure, 2000 ppm feed, 65% recovery rate and single stage SWRO with 42% recovery rate, 42,000 ppm feed and 70 bar feed pump pressure for total capacity of 100,000 $m^3$/d.*

## 3. RO membrane: types, structures and materials

Based on the membrane structure, The RO membrane is consisted of two groups: conventional thin-film composite and thin-film nanocomposite. Based on the thin-film material, conventional RO membrane is classified into two main

groups: cellulose acetate (CA) and aromatic polyamide (PA). The RO membrane on the basis of the membrane configuration can be divided into three main groups: hollow-fiber, flat-sheet (plate-and-frame) and spiral-wound [30, 31].

### 3.1 Conventional thin-film composite membrane structure

The RO membrane which is used widely today are composed a semipermeable thin film (0.2 um), made of either CA or PA, supported by a 0.025- to 0.050-mm microporous layer that in turn is cast on a layer of reinforcing fabric (**Figure 7**). Maintaining and reinforce the membrane structural integrity and durability is the main functions of the two support layers underneath the thin film [31].

In the dense semipermeable polymer film that is made up from a random molecular structure (matrix), there is no any pores. Water molecules are transported through the membrane film by diffusion and travel on a multidimensional curvilinear path within the randomly structured molecular polymer film matrix [12, 31].

### 3.2 Thin-film nanocomposite membrane structure

Thin-film nanocomposite (TFC) consisting from two main structure; inorganic nanoparticles in traditional membrane polymeric film structure (**Figure 8**) and highly structured porous film consisting of a densely packed array of nanotubes (**Figure 9**). In **Figure 8**, part A shows the thin film of a conventional PA membrane that supported by the polysulfone support layer. Part B shows the same type of membrane with embedded nanoparticles.

In nanocomposite membrane the specific water permeability, at comparable salt rejection, is higher than the conventional RO membrane. In addition, the fouling rates in TFC membrane, at the same operation conditions, is lower in comparison to conventional TFC RO membrane. In other words, in case of production of tubular membranes with completely uniform size, theoretically the membrane could produce up to 20 times more water per unit surface area than the common RO membrane commercially available on the market today.

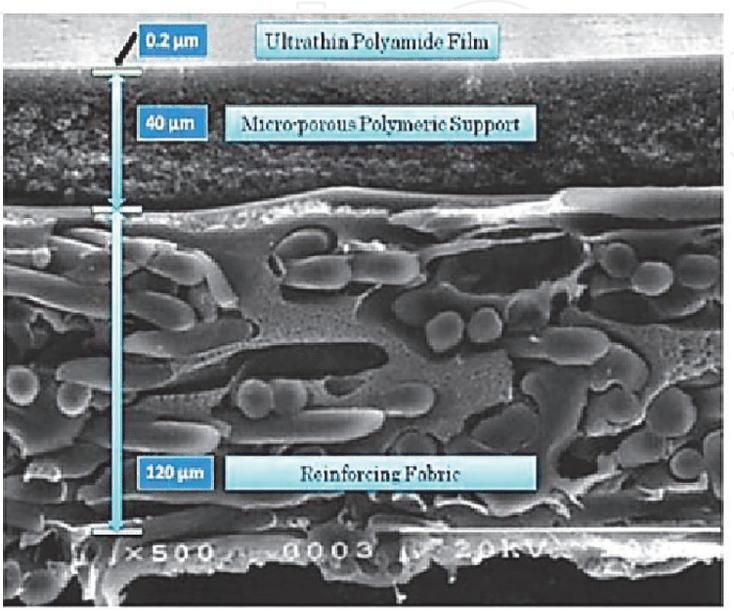


**Figure 7.**
*Structure of a typical reverse osmosis RO membrane with ultrathin PA film.*

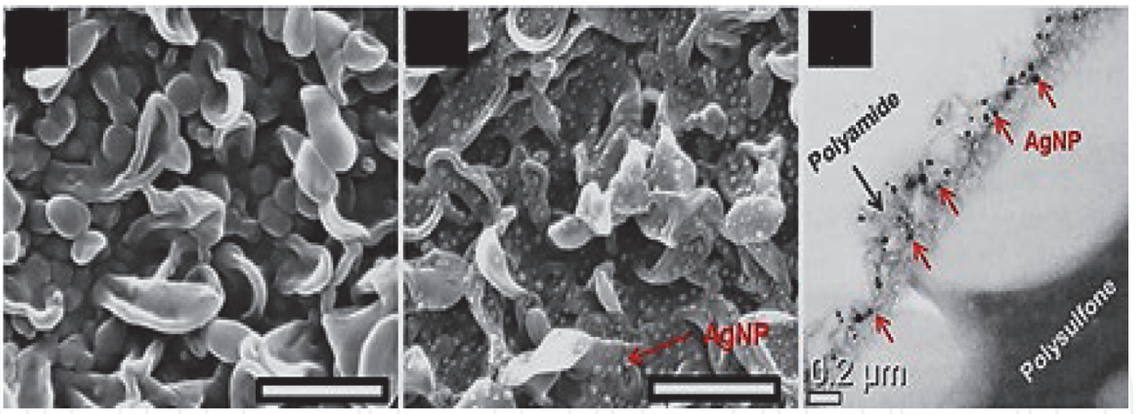


**Figure 8.**
*Polyamide reverse osmosis RO membrane with nanoparticles.*

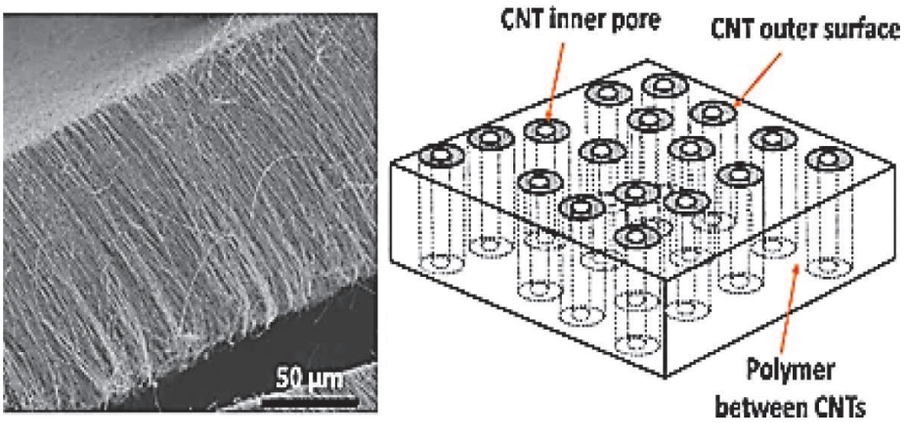


**Figure 9.**
*The RO membrane with carbon nanotubes [32].*

### 3.3 Cellulose acetate CA membrane

For the first time in the late 1950s the thin semipermeable film as the first membrane element from cellulose acetate (CA) polymer was made at the University of California, Los Angeles [33]. Although the CA membrane is similar to the aromatic polyamide (PA), but, because of the existence of the top two layers (the ultrathin film and the microporous polymeric support) in the main structure of the CA that are made of different forms of the same CA polymer, the CA is different from PA [34]. In PA membrane unlike the CA these two layers consist of two completely different polymers, the polyamide and polysulfone form the semipermeable films and microporous supports, respectively. In CA membrane similar to PA membrane, thickness of the film layer is typically about 0.2 μm, but the thickness of the entire membrane in CA membrane is different (about 100 μm) from the PA membrane (about 160 μm) [35].

One of the important advantages of CA membrane is its surface very little charge, which is considered practically uncharged, while in PA membrane, because of negative charge in the surface of the membrane, with use of cationic polymers for water pretreatment, the potential for fouling increases dramatically. Furthermore, due to the smoother surface in CA membrane than the PA membrane, the CA membrane less clogged [34].

Some disadvantages of the CA membrane are; low operation temperatures 35°C (95°F) and narrow pH working range (4–6). Operation outside of this pH range can

cause hydrolysis of the membrane, also, exposure to temperatures above 40°C (104°F) causes membrane compaction and failure [33]. Due to these limitations, the pH in feed water interring to the CA membrane has to be reduced and maintain between 5 and 5.5, which, in order to normal plant operation, the use of acid increases. in addition, the requires reverse osmosis RO permeate adjustment by addition of a base (typically sodium hydroxide) to achieve adequate boron rejection [36].

Since CA membrane has a higher density than PA membrane, it creates a higher head loss and has to be operated at higher feed pressures, which results in increase in energy consumption. Despite their disadvantages, due to their high tolerance to oxidants (chlorine, peroxide, etc.) than the PA membrane, CA membrane is used in municipal applications for ultrapure water production in pharmaceutical and semiconductor industries and for saline waters with very high fouling potential (mainly in the Middle East and Japan).

### 3.4 Aromatic polyamide membrane

The aromatic polyamide (PA) membrane widely used in RO membrane structure and production of potable and industrial water at today. The thin polyamide film of this type of semipermeable membrane is formed on the surface of the microporous polysulfone support layer. For production of PA membrane uses the interfacial polymerization of monomers containing polyamine and then immersion of it in the solvent containing a reactant to form a highly cross-linked thin film. Because of some properties such as lower working pressure, lower salt passage than CA membrane and higher productivity (specific flux), the PA membranes have wider application at today [37, 38].

By changing pH, the surface charge of PA and CA membrane is also changes. For example, CA membrane has a neutral charge while, PA membrane in pH greater than 5 has a negative charge, and for this reason, co-ion repulsion amplified and therefore salt rejection is higher than CA membrane. However, when pH is lower

| Parameter | Polyamide membrane PA | Cellulose acetate CA membrane |
|---|---|---|
| Salt rejection | High (>99.5%) | Lower (up to 95%) |
| Feed pressure | Lower (by 30–50%) | High |
| Surface charge | Negative (limits use of cationic pretreatment coagulants) | Neutral (no limitations on pretreatment coagulants) |
| Chlorine tolerance | Poor (up to 1000 mg/L-hours); feed dechlorination needed | Good; continuous feed of 1–2 mg/L of chlorine is acceptable |
| Maximum temperature of source water | High (40–45°C; 104–113°F) | Relatively low (30–35°C; 86–95°F) |
| Cleaning frequency | High (weeks to months) | Lower (months to years) |
| Pretreatment requirements | High (SDI < 4) | Lower (SDI < 5) |
| Salt, silica, and organics removal | High | Relatively low |
| Biogrowth on membrane surface | May cause performance problems | Limited; not a cause of performance problems |
| pH tolerance | High (2–12) | Limited (4–6) |

**Table 4.**
*Comparison between polyamide PA membrane and cellulose acetate CA membrane [37].*

than 4, the charge of the PA membrane changes to positive and rejection reduces significantly to lower than the CA membrane [38]. One another of the most important advantage of the PA membrane is much wider operation pH range (2–12). This allows easier maintenance and cleaning. Furthermore, the PA membrane has resistant to biodegradation and have a longer useful life (5–7 years) compare to usually membrane (3–5 years). From Aromatic polyamide membrane is used in order to production of membrane elements for nanofiltration, seawater desalination and brackish water [33, 37].

### 3.5 Comparison between PA and CA membrane

For PA membrane, the chlorine is and other strong oxidants the biggest threat and can destroying the membrane structure and consequently reduce the salt rejection performance of the membrane. In order to biofouling control in nanofiltration and RO membranes, Oxidants are widely used, so, before separation, the feed water to PA membrane has to be dechlorinated. In **Table 4**, the key parameters of polyamide and cellulose acetate RO membrane has been shown.

## 4. Recent development of novel membranes for desalination

In commercial RO membranes, almost the majority of materials that are used are dominated by thin-film composite (TFC) polyamide and its derivatives. At these membranes, we are faced with critical challenges like relatively low water permeability, high fouling tendency and low selectivity [39]. For example, in commercial TFC RO membranes the typical water permeability for seawater reverse osmosis (SWRO) and brackish water reverse osmosis (BWRO) is range from $\sim$1–2 L $m^{-2}$ $h^{-1}$ $bar^{-1}$ and $\sim$2–8 L $m^{-2}$ $h^{-1}$ $bar^{-1}$, respectively [40]. One of the fields in desalination that is been focus on it, is synthesizing novel membranes with better antifouling performance and improved separation properties.

Much of the exciting progresses are fueled by the recent emergence of promising novel materials for desalination. Among them, the most notable examples include aquaporin (AQP) proteins [11, 41, 42] and some carbon-based materials such as carbon nanotubes (CNTs) [43] and graphene-based materials [44]. At the moment, in RO membranes, the old asymmetric cellulose acetate largely replaced with TFC polyamide membranes [45, 46]. New TFC polyamide membranes compared to the former membranes, have been shown better performance in water permeability and salt rejection (e.g., in SWRO rejection of NaCl is >99.9%), pH tolerance (1–11) and wider operating temperature range (0–45°C) [11].

### 4.1 Novel materials and methods for synthesizing desalination membranes

#### *4.1.1 Carbon-based materials*

Because of exceptional water transport properties of Carbon based materials (CBMs), e.g., nanoporous graphene (NPG) [47, 48], carbon nanotubes (CNTs) [49, 50], and graphene oxide (GO) [11, 51] have been raised hopes of improvement in the membrane processes (**Tables 5** and **6**). In these materials, the characteristic of water channel dimensions as well as chemical modifications (e.g., the presence of carboxyl, amine and other groups) determines the rejection properties [11, 56, 75]. The characteristic of the channel dimensions in NPG and CNTs are sorted by their respective pore sizes [51]. In CNTs and NPG, the channel sizes determined by their synthesis conditions, but, in GO the characteristic channel size is highly dependent

| | Polyamide | AqpZ | CNT | NPG | Graphene oxide (GO) |
|---|---|---|---|---|---|
| Material transport mechanism | Cross-linked polymer solution-diffusion | Natural protein for charge repulsion and size-exclusion | Material with 1-D carbon size-exclusion (enhanced by charge repulsion) | Material with 2-D carbon size-exclusion (enhanced by charge repulsion) | material with 2-D carbon size-exclusion (enhanced by charge repulsion) |
| Characteristic channel size (Å) | Irregular pores in a random network, characteristic pore diameter of ∼4–5.8a Å based on positron annihilation lifetime spectroscopy [52], possibly heterogeneous pore distribution for some membranes [53] | Well-defined hour-glass-shaped channel [54], pore size of ∼3 Å [55] | Well-defined cylindrical pores (e.g., ∼13–20 Å [56]) | Nano-sized pores across 1-atomthick graphene layer, possibly with non-uniform pore sizes (e.g., obtained from plasma etching, ∼5–10 Å [52]) | Channels formed by adjacent GO layers, channel size depending on the degree of oxidation or solution environment [57] |
| Separation properties | ∼1–2 L $m^{-2}$ $h^{-1}$ bar-1 for SWRO and ∼ 2–8 L $m^{-2}$ $h^{-1}$ $bar^{-1}$ for BWRO; ∼ > 99% NaCl rejection (obtained from cross-flow filtration tests) [40] | ∼600 L $m^{-2}$ $h^{-1}$ $bar^{-1}$; nearly 100% NaCl rejection (obtained from stopped-flow measurements of AQP-containing vesicles) [41, 58] | Gas permeability is >10 times higher than the predictions of the Knudsen diffusion model; experimental water permeability is >1000 times higher than the calculated results from continuum hydrodynamics (obtained from measuring the water flux of an aqueous suspension of gold nanoparticles; CNTs pore density ≤ 2.5 × $10^{11}$ $cm^{-2}$ and length of ∼3 μm) [56] | ∼3.6 × 106 L $m^{-2}$ $h^{-1}$; nearly 100% KCl rejection at 40°C for a 5-μm-diameter sample (obtained from gravity-driven test in an oven) [47] | Water permeability is at least $10^{10}$ times faster than that of helium (obtained from weight-loss measurements by a 1-μm-thick GO membrane) [44]; water permeability and rejection are sensitive to the interlayer spacing |
| Antifouling properties | Prone to fouling [36] | Not reported in literature | Antimicrobial [56] (and improved hydrophilicity for functionalized CNTs [59]) | Not reported in literature | Antiadhesion (due to hydrophilicity) and antimicrobial [60] |
| Electrical conductance | No | No | Yes | Yes | No |



**Table 5.**
*Material properties of polyamide, AqpZ, CNT, NPG and graphene oxide [11].*

| Type | Classification | $P_w$ ($L\ m^{-2}\ h^{-1}\ bar^{-1}$) | Rejection (%) | Testing conditions and membrane area ($cm^2$) | Results | Ref. |
|---|---|---|---|---|---|---|
| PRL AqpZ DOPC[a] | NF | 3.6 | $R_{NaCl}$ = 20% | 1 mM NaCl @1 bar Area: 28.3 | DOTAP coated NF270, with both decreased water flux and $R_{NaCl}$ compared to virgin membranes | [61] |
| AqpZ-ABA[b] | NF | 34.2 | $R_{NaCl}$ = 32.9% | 200 ppm NaCl @5 bar Area: 0.071 | Silanized CA substrate, high $P_w$ with low $R_{NaCl}$, the amount of AqpZ has huge impact on membrane performance | [62] |
| AqpZ-ABA | NF | 16.1 | $R_{NaCl}$ = 45.1% | 200 ppm NaCl @5 bar Area: 0.2 | Gold coated porous alumina substrate cross-linked with disulfide: high $P_w$ with less defects | [63] |
| AqpZ-DOPC/ DOTAP[c] | NF | 5.5 | $R_{NaCl}$ = 75%<br>$R_{MgCl2}$ = 97% | 500 ppm NaCl @4 bar Area: 19.56 | AQP containing lipid bilayers deposited on PSS/PEI/ PAN substrate | [64] |
| AqpZ-ABA | FO | $J_v$[d] = 16.4 L $m^{-2}$ $h^{-1}$ | $R_{NaCl}$ = 98.8% | 0.3 M sucrose as DS, 200 ppm NaCl as FS[e] Area: 0.096 | Gold and cysteamine coated polycarbonate with UV cross-linking | [65] |
| AqpZ-DOPC/ DOTAP | FO/NF | $J_v$ = 23.1 L $m^{-2}$ $h^{-1}$ NF:6.31 | FO: Js = 3.1 g $m^{-2}$ $h^{-1}$<br>NF: $R_{MgCl2}$ = 90% | 2 M $MgCl_2$ as DS, DI water as FS 2000 ppm $MgCl_2$ @ 4 bar Area: 36 | AqpZ-DOPC/DOTAP coated on PDA modified porous polysulfone substrate via amidation reaction to form covalent bonds. | [66] |
| TFN AqpZ-DOPC | RO | 4 | $R_{NaCl}$ = 97% @ 5 bar | 10 mM NaCl @5 bar Area: > 200 | AqpZ containing vesicles incorporated in PA layer serving as protection layer via IP. Large membrane area can be obtained | [67] |
| AqpZ-DOPC | RO | 8 | $R_{NaCl}$ = 97.5% | 500 ppm NaCl @5 bar Area: 34.2 | Vesicles embedded in PA rejection layer with superior water flux | [68] |
| AqpZ-DOPC | RO | 4.1 | $R_{NaCl}$ = 97.2% | 10 Mm NaCl @10 bar Area: 42 | Vesicles embedded in PA rejection layer for long term stability test | [69] |
| AqpZ-POPC/ POPG/cholesterol[g] | NF | ~6 | $R_{MgCl2}$ = 96% | 200 ppm $MgCl_2$ @ 4 bar Area: 0.785 | Vesicles embedded in PSS/PAA LBL[f]. Membranes with AqpZ showed Pw ↑ 60% with $MgCl_2$ rejection↑ compared to the control | [70] |
| AqpZ-DOPC | NF | 36.6 | $R_{MgCl2}$ = 95% | 100 ppm $MgCl_2$ @ 1 bar Area: 28.3 | PDA coated vesicles incorporated in cross-linked PEI matrix | [71] |

| Type | Classification | $P_w$ ($L\ m^{-2}\ h^{-1}\ bar^{-1}$) | Rejection (%) | Testing conditions and membrane area ($cm^2$) | Results | Ref. |
|---|---|---|---|---|---|---|
| AqpZ-ABA | NF/FO | NF: 22.9<br>$J_v$ = 5.6 L $m^{-2}$ $h^{-1}$ | $R_{NaCl}$ = 61%<br>$R_{MgCl2}$ = 75%<br>FO: $R_{NaCl}$ = 50.7% | 200 ppm salt @5 bar 0.3 M sucrose as DS and 200 ppm NaCl as FS | AqpZ-vesicle loaded membrane cross-linked by UV | [72] |
| AqpZ-ABA | FO | $J_v$ = 43.5 L $m^{-2}$ $h^{-1}$ | $J_s$ = 8.9 g $m^{-2}$ $h^{-1}$ | 0.5 M NaCl as DS, DI water as FS<br>Area: 0.196 | Pressure assisted sorption, further coated with cysteamine and cross-linked by polydopaminehistidine. The control membrane has FO water flux of 8.6 L $m^{-2}$ $h^{-1}$ and Js = 6.6 g $m^{-2}$ $h^{-1}$ | [73] |
| AqpZ-POPC/ POPG/Cholesterol | FO | $J_v$ = 21.8 L $m^{-2}$ $h^{-1}$ | Js = 2.4 g $m^{-2}$ $h^{-1}$ | 0.3 M sucrose as DS and 200 ppm $MgCl_2$ as FS<br>Area: 0.785 | Magnetic-assisted AQPs embedded membranes | [74] |

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[a]*DOPC: 1,2-dioleoyl-sn-glycero-3-phosphocholine.*
[b]*ABA: methacrylate end functionalized poly(2-methyloxazolineb-dimethylsiloxane-b-2-methyloxazoline) PMOXA(1000)-b-PDMS(4000)-PMOXA(1000) triblock.*
[c]*DOTAP: 1,2-dioleoyl-3-trimethylammonium-propane.*
[d]*$J_v$: FO water flux; $J_s$: FO solute flux.*
[e]*DS: draw solution; FS: feed solution.*
[f]*LBL: layer by layer deposition of polyacrylic acid (PAA) and polystyrene sulfonate (PSS).*
[g]*POPC: 1-palmitoyl-2-oleoyl-sn-glycero-3-phosphocholine; POPG: 1-palmitoyl-2-oleoyl-sn-glycero-3-phospho-(19-rac-glycerol).*

**Table 6.**
*Summary of RO, NF and FO performance of biomimetic membranes [11].*

on solution environment and its degree of oxidation [11]. In this section, we have summarized the detailed materials properties of NPG, CNT and GO [76–78].

## 5. Hybrid technologies: the future of energy efficient desalination

Desalination processes traditionally rely on mechanically driven membrane processes such as reverse osmosis (RO) or thermal distillation such as multi-effect distillation (MED) and multi-stage flash (MSF). In the use of membrane technologies, the principle is based on the use of technology with easy operation, limited use of chemicals, compactness, low energy consumption and the development of enhanced membrane materials [79]. Some emerging desalination technologies like forward osmosis (FO) and freeze desalination (FD), despite the serious challenges in the road to commercialization, have also recently garnered interest.

In a desalination plant, roughly 20–30% of the overall cost in water production is related to the energy [22, 29]. There is growing interest in combining the benefits of two or more systems, to meet specific water quality goals and/or reduce energy consumption. Using hybridization in desalination technologies is often in order to one or more objectives such as increasing water recovery rate, eliminating the need for a second pass or reducing brine salinity. Hybrid systems have been considered as economically superior alternatives to standalone systems due to their ability to reduce energy consumption and therefore cost of desalinated water through improved recovery rate and/or water quality [80].

### 5.1 Current status and energy consumption in desalination systems

#### *5.1.1 Multi-stage flash (MSF)*

The basis of working multi-stage flash distillation (MSF) is distills sea water by flashing a portion of the water into steam in multiple stages of what are essentially countercurrent heat exchangers [81]. In order to occur the flashing, the pressure in each stage must be lower than the vapor pressure of the heated liquid. by passing the cold feed from each stage, it be heated that is further heated in the brine heater. At the time of brine flows return, because of higher temperature than the boiling point in brine, in the normal pressure, a fraction of the brine boils to the steam. After this stage, the steam is starting to condensation on the external surface of heat exchanger tubes [82]. At the moment, two more well-known configurations of the MSF are the once-through MSF (MSF-OT) and brine mixing MSF (BM-MSF) [80].

At this moment, about 23% of all desalinated water in the world is produced by MSF plants, but due to the high energy consumption, their use is declining [83].

In the practical scale, for commercial MSF systems, a value of 8 to 12 $kg_{distillate}/kg_{steam}$ are typically reported [84]. Some parameters like as corrosion and pipe fouling, scale formation and etc. reduce the energy efficiency of MSF systems. In MSF plants the amount of energy that consume is between 23 and 27 $kWh/m^3$ [80, 85]. El-Naser [86] reported that in MSF plants the energy consumption is average 12–24 $kWh/m^3$.

#### *5.1.2 Multi-effect distillation (MED)*

One of the oldest industrial desalination processes that are used today is Multi-effect distillation (MED) [87, 88]. The MED evaporator consists of cells, called effects, decreasing pressure and temperature from first to last, with temperature typically between 65 and 90°C [89]. Each effect consists of evaporator tube bundles

on which seawater is sprayed. Heating steam or hot water through the tubes is supplied in the first effect and it transfers energy to the seawater in each effect, causing partial evaporation [90]. In each effect, the low pressure and temperatures affect the boiling point of water and with decreases of its, water becomes evaporate [88]. By using a heat exchanger and condensing the steam, clean distillate water is produced. This product water is pumped into a storage tank while the brine is pumped back into the sea.

For the production of water in a MED plant with a capacity range between 5000 and 50,000 $m^3$/day, we require thermal energy between 145 and 230 MJ/$m^3$, which will be equal to 12.2–19.1 kWh/$m^3$ of electrical energy. Furthermore, for pumps consumption will have been needed 2–2.5 kWh/$m^3$ of additional electrical energy [91]. Vapor flow and feed configurations are two major parameters that can effect on energy consumption in the MED process.

### *5.1.3 Electrodialysis (ED)*

Electrodialysis (ED) is an electro membrane process in which with use of an electric field ionic and non-ionic components are removed [29]. In these kinds of processes, Anions and cations migrate towards the positive and negative electrode, respectively, and so the separation process happens. As can be show in the **Figure 10**, an ED system consists of alternately arranged anion exchange membranes (AEM) and cation exchange membranes (CEM).

The energy consumption in ED strongly depends to the salt concentration in feed solution. The rate of salt removal is proportional to the electric current [80, 92]. In order to efficient separation of ions from feed solution with high concentration, would require a high potential difference, thus, the use of ED process for seawater desalination, due to high concentration of ions in seawater and the need for high energy consumption, it is not affordable. This process is suitable for solutions with low-concentration of TDS (<5000 mg/L) such as brackish water [93]. Other parts that consume energy is the pumping unit and electrodes. On the basis of recent study, about 1–3% of the total energy consumption is related to these sections [92, 94].

Theoretically, in ED, for producing water with TDS about 800 mg/L the requirement of energy is 3.3 kWh/$m^3$ and 26 kWh/$m^3$ for desalination of brackish water and seawater, respectively [95]. On average, 0.7 kWh for each 1000 mg/L

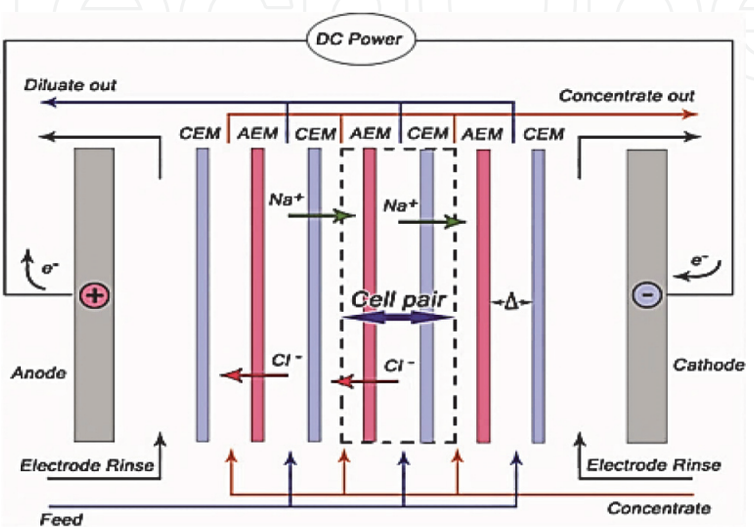


**Figure 10.**
*Schematic of electrodialysis desalination.*

TDS removed, 0.5–1.1 kWh/$m^3$ for pumping, and roughly 5% accounts for energy losses in a brackish water ED desalination system [96]. In a study that was reviewed by Sajtar and Bagley, they found in order to removal of TDS up to 2000 mg/L in feed stream, the energy consumption is ranges from 0.1 to 1 kWh/$m^3$ [92, 94]. Although ED is typically applied as a room temperature process, introducing a temperature gradient or increasing the temperature of the system can cause energy reductions [94]. Benneker et al. [97] found that the energy required for ED can be reduced by 9% if the temperature of one of the feed streams is increased by 20°C. Increasing the temperature increases ion mobility, reduces electrical resistance of the solution and decreases solution viscosity.

On the basis of the water salinity, the consumption of the electrical energy by an ED system can be about 0.5–10 kWh/$m^3$ [98]. For example, to lower TDS from 1500 ppm to 500 ppm, an ED unit would consume ~1.5 kWh/$m^3$. Due to high energy consumption in ED systems, in order to management and reduced the energy consumption, Recently, multi-stage electrodialysis systems have been investigated. Chehayeb et al. [99] found that by using a two-stage system for brackish water desalination the energy consumption can be reduced up to 29%, that, this can reduce the fixed costs. The application of ED remains limited by the high cost of ion exchange membranes and electrodes, and the electrically-driven degradation of polymeric membranes [100].

### *5.1.4 Membrane distillation (MD)*

Membrane distillation is one kind of separation process which in it, a porous membrane with hydrophobic properties is in contact with aqueous heated feed solution on one side. In MD process, the membranes that was use it works like this, that inhibit from the passage of the liquid water, but on the contrary allowing permeability for free water molecules and thus, for water vapor. These membranes are made of hydrophobic synthetic material (e.g. PTFE, PVDF or PP) and offer pores with a standard diameter between 0.1 and 0.5 μm ($3.9 \times 10^{-6}$ and $1.97 \times 10^{-5}$ in) [80, 101].

Due to the high amount of energy consumption and as a result the high cost of water production, MD has not still achieved widespread commercial implementation in desalination. There are four basic MD configurations included [102, 103];

- direct contact membrane distillation (DCMD).
- vacuum membrane distillation (VMD).
- air-gap membrane distillation (AGMD).
- sweeping gas membrane distillation (SGMD).

In several studies it has been reported that both AGMD and VMD have greater thermal energy efficiency compared to other configurations, which makes them more popular choices for companies seeking to commercialize MD processes. In **Table 7**, the SEC values for several selected MD systems have been reported [102, 116–118].

### *5.1.5 Forward osmosis*

One kind of osmotic process is called forward osmosis (FO) that, in this process, like RO, in order to the separation of water from dissolved solutes, uses a semi-permeable membrane. This process for creating the driving force for separation uses the osmotic pressure gradient, such that a “draw” solution of high

| Configuration | Membrane characteristics | Operating conditions | | Feed type | SEC (kWh/m$^3$) | Plant capacity (m$^3$/h) | Refs. |
|---|---|---|---|---|---|---|---|
| | | $T_f$ (°C) | $T_p$ (°C) | | | | |
| DCMD | Spiral wound PTFE (SEP GmbH), pore size 0.2 μ, porosity 80% | 35–80 | 5–30 | Radioactive solution | 6000–1000 | 0.05 | [104] |
| AGMD | PTFE, pore size 0.2 μ | 60–85 | — | Seawater | 140–200 | 0.2–20 | [105] |
| AGMD | | 313–343 | — | Brackish water | 30.8 | | [106] |
| AGMD | PTFE, pore size 0.2 μ, porosity 80% | — | — | Seawater | 200–300 | 3.46–19 | [107] |
| DCMD in hybrid systems | PP models from Microdyn Nadir, Pore size 0.2 μ, porosity 73% | — | — | Seawater | 1.6–27.5 | 931 (overall) | [108] |
| DCMD | Commercial membranes from membrane with pore size 0.2 μ and thickness 91 μ | 39.8–59 | 13.4–14.4 | Distilled water | 3550–4580 | — | [109] |
| VMD | PP, thickness 35 μ, pore size 0.1 μ | 15–22 | — | Underground water | 8100.8–9089.5 | 2.67–6.94 | [110] |
| AGMD | LDPE, thickness 76 μ, pore size 0.3 mμ, porosity 85%, $A_m$ 7.4 m$^2$ | 50–70 | — | Tap water, synthetic seawater | ~65 to ~127 | — | [111] |
| VMD | Flat sheet PP, thickness 400 μm, Pore size 0.1 μ, porosity 70%, $A_m$ 5 m$^2$ | 80 | — | Distilled water | 130 | — | [112] |
| DCMD | PVDF hollow fiber, thickness 240 μm | 80 | 30 | Simulated reverse osmosis brine | ~130–1700 | — | [113] |
| DCMD | PTFE with PP support, mean pore size 0.5 ± 0.08, porosity 91 ± 0.5, active layer thickness 46 ± 1 μm, $A_m$ 0.67 m$^2$ | 60 | 18–21 | Wastewater | 1500 | 3.85 | [114] |
| DCMD | Several commercial membranes with different characteristics | 85 | 20 | Seawater | 697–10,457 | — | [115] |

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**Table 7.**
*Specific energy consumption (SEC) of selected MD systems [80].*

concentration is used to induce a net flow of water through the membrane into the draw solution, thus effectively separating the feed water from its solutes [80, 119]. As a result, separation in FO requires little or no hydraulic pressure as a concentrated draw solution (DS) with a greater osmotic pressure draws in water molecules from the feed solution through a membrane [120].

FO is widely promoted as a low-energy desalination technique. For the determination of the energy consumption in these kinds of plants, a DS recovery step is

used. During the osmosis step, in order to overcome dropping the pressure in the feed channel, at 50% water recovery, a low-pressure pump is needed, and the energy consumed is equal to ~0.10–0.11 kWh/$m^3$ [25, 121]. For the osmosis step the values of 0.2–0.55 kWh/$m^3$ have also been reported [122]. Moon and Lee suggest, in a FO desalination plant, for solute regeneration, the energy consumption range is from 3 to 8 kWh/$m^3$ [123].

### 5.2 Hybrid desalination technologies

In a hybrid desalination system in order to reduce costs or enhance performance in compared to individual components, uses from integration of two or more desalination systems. Due to the high cost of investing in hybrid systems, one of the important parts of these kinds of processes is the optimization of hybrid configurations [80].

#### *5.2.1 Electrodialysis: reverse osmosis hybrid systems (ED-RO)*

Increasing recovery in RO systems requires multiple stages and thus significantly increased capital and operation costs [124]. In the electrical desalination systems such as ED compare to the RO membranes, we cannot achieve to high salt rejection alone [125], and this is very important in energy consumption, however, one of the advantages of ED systems, is operation at higher recovery rate, but and low SEC, by scale formation this process eventually limited [80]. The concept of (ED-RO) hybrid system at first in 1981 by Schmoldt et al. was studied [126]. They proposed the use of ED as a second stage to control permeate quality. However, one of the disadvantages of this system was high energy consumption up to 7.94 kWh/$m^3$ for SWRO system with a concentration of 45,000 ppm, that was due to some problems such as lack of high-flux and high-selectivity membranes in this process [80, 126]. But, in their studies they showed in a desalination plant with capacity of 1000 $m^3$/day and feed concentration of lower than 4000 mg/L, the investment cost for ED can be lower than RO. They noted that with the development of the high flux membranes and with high salt-rejection, not only the cost of the RO system could be reduced, hence, with incoming feed with lower TDS concentration, the energy consumption of the ED unit also reduce [126].

In a another study, by Turek et al. [127], in order to assessment SEC and recovery rates, in four different configurations (single-stage standalone RO, NF-SWRO, hybrid ED-RO and NF-SWRO-ED system) for seawater desalination plants were been compared. As can be seen in **Table 8**, the highest recovery (81.1%) was achieved for SWRO-ED, but, at this recovery rate, the SEC was 7.77 kWh/$m^3$, after that the NFSWRO-ED system had more recovery rate (69.0%) at lower SEC (6.90 kWh/$m^3$). Although SEC in the SWRO system was much less (2.76 kWh/$m^3$), but on the other hand, this single-stage RO system operated at a recovery rate of only 43% [80].

In **Table 9**, a comparison of selected ED-RO studies has been presented.

#### *5.2.2 Reverse osmosis: membrane distillation hybrid systems*

Several advantages of MD system like as operation at high recovery, high separation efficiency and Low capital cost, has made it alternative candidate for hybrid separation technologies [132, 133]. Over the last few years, a few studies on the hybridization of MD and RO in order to treatment of the concentrate stream from the RO process have been done. For example, in a study by Choi et al., economic feasibility of a RO-MD system for desalination of seawater was assessed. In this study, they found that a RO-MD hybrid system or a MD stand-alone system only when the flux and recovery are greater than that for RO, and or the thermal energy

| System | Energy consumption [kWh $m^{-3}$] | Water recovery [%] |
|---|---|---|
| SWRO | 2.76 | 42.6 |
| SWRO-ED | 7.77 | 81.1 |
| NF-SWRO | 3.93 | 41.2 |
| NF-SWRO-ED | 6.90 | 69.0 |

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**Table 8.**
*The SEC and water recovery for SWRO-ED, SWRO, NF-SWRO and NF-SWRO-ED systems [80].*

| Feed type | Hybridization | Feed TDS (mg/L) | Product TDS (mg/L) | Recovery rate | SEC ($kWh/m^3$) | Refs. |
|---|---|---|---|---|---|---|
| Brackish water | ED as pretreatment to lower RO feed salinity | 2000–4000 | 50–120 | RO alone: 10–20% ED-RO: 50–60% | RO alone: 7.8 ED-RO: 8–10 | [128] |
| Wastewater | ED of RO concentrate | 2550–3550 | — | RO alone: 75% ED-RO: 95% | — | [129] |
| Brackish water | ED of RO concentrate; ED product water blended with RO permeate to produce water | 3000 | 300 Hybrid preferred over ED alone only when product TDS requirement is strict | 50% | — | [130] |
| Hypersaline brine | Counterflow ED with RO | 120,000 | — | Performance at high recoveries is limited by concentration differences | — | [131] |

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**Table 9.**
*Key parameters from selected ED-RO hybridization studies [80].*

that has been supplied for MD, had relatively low cost, can compete with RO system [134]. Although, MD is able to achieve a high water recovery rate of 85%, However, the Energy consumption for RO-MD hybrid systems is still unclear and should be further investigated [80].

### *5.2.3 Forward osmosis (FO)-RO*

**Table 10** shows the summary of hybrid FO-RO system for seawater desalination [135].

### *5.2.4 Nanofiltration (NF)-RO*

Using of MF, UF membrane although can be effective for the pretreatment of a SWRO system, but some important parameter such as NOMs, organic matters and dissolved organic matters cannot be fully removed. Since in MF and UF divalent metal ions do not remove, so, the potential of the Scaling cannot be reduced. As we know, in SWRO desalination facility, about 44% of water production costs are related to energy consumption, which is closely related to the salinity of seawater. Hence, in order to pretreatment and effectively reduction of overall salinity (reduce

| System | System Detail | FO | | RO | Effect | Ref. |
|---|---|---|---|---|---|---|
| | | **Membrane** | **Draw solution** | | | |
| FO-RO | Glucose draw solution (DS) is diluted by seawater at FO and diluted glucose solution is subjected to RO to recover water | — | Glucose | Low pressure reverse osmosis (LPRO) | Low osmotic pressure of glucose, high internal concentration polarization (ICP) | [136] |
| FO-RO | Secondary waste water is supplied to FO to dilute Red Sea water, which is then subjected to RO | CTA | Red Sea water | LPRO | Energy requirement 50% of SWRO (1.5 $kWh/m^3$) | [137] |
| FO-RO-FO | Secondary wastewater is supplied to FO to dilute seawater, which is then subjected to RO to obtain product water. RO brine goes to second FO to be diluted before discharge. | CTA | | SW30 2540 Dow Filmtec | Wide range of organic compounds can be removed by FO | [138] |
| Pressure assisted FO (PAFO)-RO | Wastewater supplied to FO to dilute seawater, which is then subjected to RO | | | | Simulation pressure assisted FO (PAFO) at 6 bar further reduces the water production cost. System operation is stabilized | [139] |

**Table 10.**
*Summary of hybrid FO-RO system for seawater desalination [135].*

divalent cations) in SWRO system, nanofiltration (NF) can be used [140–142]. In **Table 11**, the summary of the NF-RO hybrid systems is shown. From the view point of the energy consumption, addition of NF pretreatment will increase the energy consumption due to the added pumping energy. However, due to the reduction of salinity in the influent feed solution of RO, the energy consumption decrease [135].

### *5.2.5 Pressure-retarded osmosis (PRO)-RO*

Pressure-retarded osmosis (PRO) is a device to generate power using osmosis.

There are two advantages of coupling SWRO and PRO; (1) enhancement of the power generation in PRO due to the higher osmotic pressure of concentrated brine than seawater, (2) dilution of the concentrated brine before discharging to the ocean In order to combination of RO and PRO there are many different ways, but they can be classified in two groups. First one is transferring the high pressure of DS to the RO feed by using of pressure exchanger and other is generation of electricity with high-pressure DS that spins the turbine. So, with these changes, the specific energy required for water production is reduced (**Figures 11–13**) [159].

There are a number of simulation studies for the RO-PRO hybrid system but only few experimental works have been done using either a small lab-scale equipment or a large demonstration plant, as summarized in **Table 12**, which was made based on the work of Kim et al. [135, 159].

| Plant or organization | Pretreatment system | Effect | Refs. |
|---|---|---|---|
| Saline water conversion corporation (SWCC) | Dual and fine sand media filtration (DFSMF)-NF (DFSMF)-NF for RO-multiflush distillation (MFD) | Reduction of total hardness 93%, and TDS 57.7% by NF, MFD operable at distillation temperature of 120°C | [143] |
| | (DFSMF)-NF | Production of SWRO increased >60% with 30% cost reduction | [144] |
| Umm Lujj, Saudi Arabia | (DFSMF)-NF | Demonstration plant construction based on the above work | [145] |
| | NF | Removal of colloidal matters and inorganic scale matters was possible | [146] |
| | UF-NF | 96.3% TOC was removed with 0.06–0.36 mg/L TOC in the filtrate. Gradual membrane fouling was observed | [34] |
| | NF for RO-MD | Water production cost of 0.92 $/m3 with recovery factor of 76.2% | [147, 148] |
| | NF-RO-Membrane Crystallization (MCr) NF for RO-MD | It was possible to remove hardness, turbidity, microorganisms, and to reduce chemical and energy consumption. Water production cost was reduced 30% | [149, 150] |
| Desalination household scale plant (Luna Water 100 GPD) | NF, RO, and NF-RO | Hybrid was the best with rejections of salinity 78.65, TDS 76.52, EC 76.42, Cl 63.95, and Na 70.91% | [151] |
| Treatment of mine impaired water | Fertilizer drawn FO (FDFONF) is compared with MF-RO and UF-RO | Energy consumption for FDFO-NF was 1.08 kWh/$m^3$, which is 13.6% less energy than an MF-RO and 21% less than UF-RO | [152] |

**Table 11.**
*Summary of NF-RO hybrid system [135].*

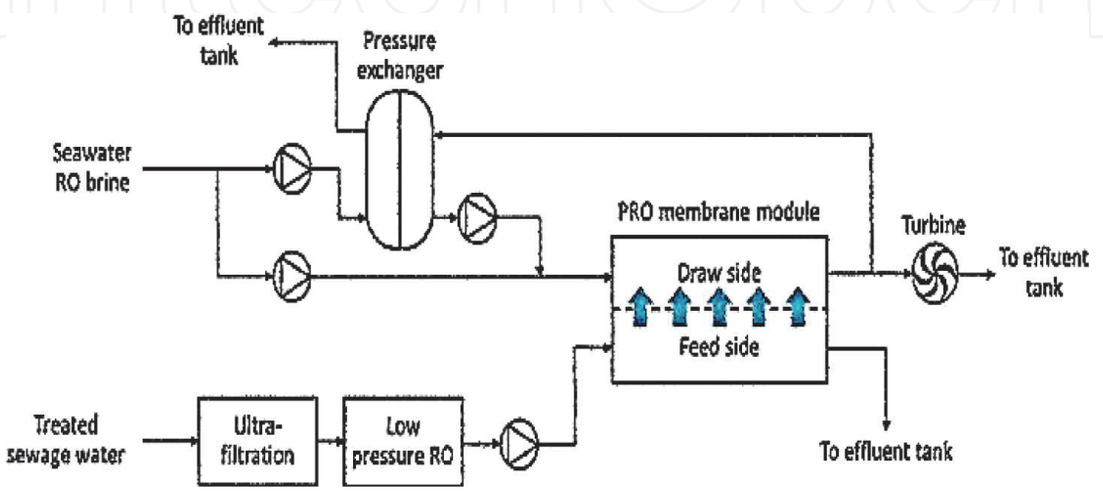


**Figure 11.**
*RO-PRO system in Japan [153].*

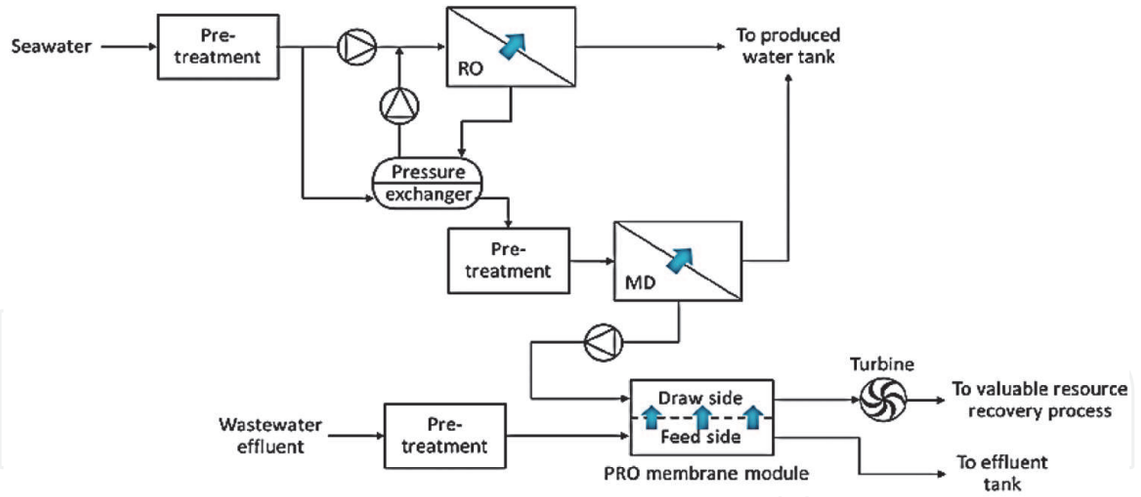


**Figure 12.**
*PRO-MD-RO system in Korea [154].*

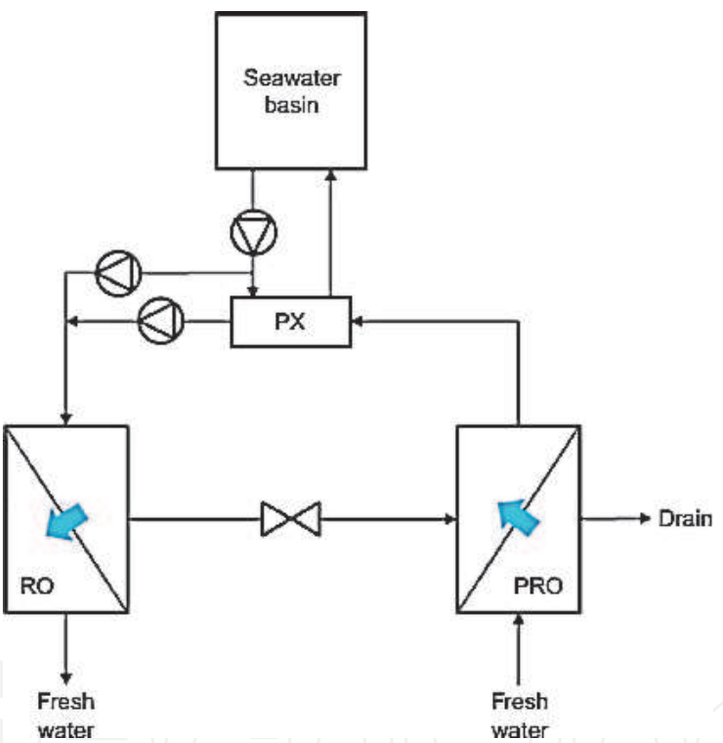


**Figure 13.**
*RO-PRO system of Achilli et al. [80].*

## 6. Conclusion

Considering that consumption of the Energy in hybrid systems, especially for FO-MD, RO-MD and FD-MD processes, due to different operating conditions in many studies are still unclear, we need more research to expand their use in the desalination industry. Research efforts should be directed towards design improvement and evaluation of energy consumption.

Elimination of the restrictions on the use of salinity gradient power technologies and directing them towards commercialization would render hybrid desalination systems more economically and also could use the salinity gradient power as an energy recovery system on their own or with other ERDs in desalination systems as could be used as. In addition to the development of low-cost high power density

| System | System detail | PRO | | RO | Effect | Refs. |
|---|---|---|---|---|---|---|
| | | Membrane | Draw Solution | | | |
| RO-PRO | RO brine goes to DS side and pretreated wastewater goes to feed side of PRO | CTA hollow fiber (Toyobo) | RO brine | | 7.7 W/m$^2$ was obtained at 2.5 MPa | [153] |
| RO-MD-PRO | RO brine goes to MD to be further concentrated. MD brine goes to the DS side and pretreated wastewater goes to the feed side of PRO | | | | RO and MD water production capacity of 1000m$^3$/day and 400 m$^3$/day, respectively, was achieved with power density of 5 W/m$^2$ | [154] |
| RO-PRO | RO brine goes to DS, filtrated tap water goes to the feed side of PRO High pressure of DS is transferred to seawater inlet | 4040 PRO module (Oasys Water) | RO brine | SW30–2540 (Dow Film Tec) | Power density of 1.1–2.3 W/m$^2$ was obtained | [155] |
| RO-PRO | Same as above | CTA membrane (HTI) | RO brine | SW30–4040 (Dow Film Tec) | Simulation based on the experimental data obtained from RO and PRO subsystem. Net specific power consumption for water production is 1.2 kWh/m$^3$ at 50% RO recovery, 40% less than RO standalone | [156] |
| RO-PRO | | | | | Economic evaluation of RO-PRO hybrid system using model equations | [157] |
| RO-PRO | | 10-in hollow Fiber module | RO brine | Toray low pressure RO | 13.5 W/m$^2$ membrane power density. On top of 20% energy reduction by low-pressure RO membrane and RED further 10% energy saving was possible | [158] |

**Table 12.**
*Some experimental results of PRO-RO hybrid system [135].*

membranes and systems for reverse electrodialysis and pressure retarded osmosis, the implementation and testing of pilot plants would speed up their transition and make them more commercially viable for industrial scale operation with other desalination processes [80].

## Acknowledgements

I thank the Desalination publication (Ahmed FE, Hashaikeh R, Hilal N. Hybrid technologies; 2020, Desalination) and (Yang Z, Ma XH, Tang CY; 2018, Desalination) to cultivate the idea of gathering information this book chapter.

*DOI: http://dx.doi.org/10.5772/intechopen.101778*



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