

Contents

1. Introduction	2318
2. History of desalination	2320
3. Reverse osmosis: basic principles	2322
4. Reverse osmosis desalination: feed waters	2325
4.1. Seawater RO	2325
4.2. Brackish water RO	2326
5. Feed water contaminants	2326
5.1. Seawater RO	2326
5.2. Brackish water RO	2326
6. Membrane fouling	2327
6.1. Seawater RO	2328
6.2. Brackish water RO	2328
7. Membrane cleaning	2329
8. RO pretreatment for seawater and brackish water	2329
8.1. Conventional pretreatment	2329
8.2. Membrane pretreatment	2330
9. Reverse osmosis system design	2331
9.1. Typical operational parameter ranges	2331
9.2. Seawater RO system design	2331
9.3. Alternative seawater RO plant design	2334
9.4. Brackish water RO system design	2334
9.5. Alternative brackish water RO plant design	2335
10. RO permeate post-treatment	2336
10.1. Seawater RO	2336
10.2. Brackish water RO	2337
11. RO concentrate disposal	2337
11.1. Seawater RO	2337
11.2. Brackish water RO	2337
12. Alternative energy sources	2338
12.1. Seawater RO	2338
12.2. Brackish Water RO	2339
13. Costs	2339
13.1. Seawater RO	2340
13.2. Brackish water RO	2341
14. Technological challenges and the future of RO	2342
15. Conclusions	2342
Acknowledgements	2342
References	2343

1. Introduction

The U.S. Geological Survey (Gleick, 1996) found that 96.5% of Earth's water is located in seas and oceans and 1.7% of Earth's water is located in the ice caps. Approximately 0.8% is considered to be fresh water. The remaining percentage is made up of brackish water, slightly salty water found as surface water in estuaries and as groundwater in salty aquifers. Water shortages have plagued many communities, and humans have long searched for a solution to Earth's meager fresh water supplies. Thus, desalination is not a new concept; the idea of turning salt water into fresh water has been developed and used for centuries.

Today, the production of potable water has become a worldwide concern; for many communities, projected population growth and demand exceed conventional available water resources. Over 1 billion people are without clean drinking water and approximately 2.3 billion people (41% of the world population) live in regions with water shortages (Service, 2006). For most, solutions such as water conservation and water transfer or dam construction are not sufficient

methods to cope with increasing demand and, in many cases, decreasing supply. Traditional fresh water resources such as lakes, rivers, and groundwater are overused or misused; as a result, these resources are either diminishing or becoming saline. As countries continue to develop and cities expand, few new water resources are available to support daily fresh water needs. As a result, solutions such as water reuse and salt water desalination have emerged as the keys to sustaining future generations across the globe.

Both water reuse and desalination have been incorporated successfully to provide additional fresh water production for communities using conventional water treatment and fresh water resources (Nicot et al., 2007; Reahl, 2004; Sanz et al., 2007; Sauvet-Goichon, 2007; U.S. EPA, 2004). Water reuse has been used to provide water for uses such as irrigation, power plant cooling water, industrial process water, and groundwater recharge and has been accepted as a method for indirect drinking water production (Focazio et al., 2008; Fono et al., 2006; Sedlak et al., 2000). Desalination has become an important source of drinking water production, with thermal desalination processes developing over the past 60 years and

Nomenclature	
C	concentration
CAPS	compact accelerated precipitation softening
CF	concentration factor
D	water diffusivity
DOC	dissolved organic carbon
D_{AB}	binary diffusion coefficient
D_s	salt diffusivity
ED	electrodialysis
EDR	electrodialysis reversal
IAP	ion activity product
K_s	salt partition coefficient
$K_{sp,x}$	solubility product constant
l	membrane thickness
MED	multi-effect distillation
MF	microfiltration
MSF	multi-stage flash distillation
MWCO	molecular weight cut off
N	mass flux
NF	nanofiltration
p	hydrostatic pressure
dp/dx	pressure gradient, x-direction
Q	volumetric flow rate
R	ideal gas constant
RO	reverse osmosis
R_s	salt rejection
R_t	total flow resistance
R_w	recovery
S	water solubility
SDI	silt density index
SI_x	saturation index
t	time
T	temperature
TDS	total dissolved solids (mg/L)
TFC	thin film composite
UF	ultrafiltration
V	partial molar volume
VC	vapor compression
ZLD	zero liquid discharge
κ	permeability
ρ_A	mass density
μ	viscosity
π	osmotic pressure
L	permeability coefficient
ΔP	transmembrane pressure difference

membrane processes developing over the past 40 years (Gleick, 2006).

Desalination is a general term for the process of removing salt from water to produce fresh water. Fresh water is defined as containing less than 1000 mg/L of salts or total dissolved solids (TDS) (Sandia, 2003). Above 1000 mg/L, properties such as taste, color, corrosion propensity, and odor can be adversely affected. Many countries have adopted national drinking water standards for specific contaminants, as well as for TDS, but the standard limits vary from country to country or from region to region within the same country. For example, the World Health Organization (WHO, 1970) has a drinking water taste threshold of 250 mg/L, and the U.S. Environmental Protection Agency (EPA) has secondary (non-enforceable) standards of 250 mg/L chloride and 500 mg/L TDS (U.S. EPA, 2002). Each U.S. state can set a primary, enforceable standard. The state of Utah currently has a TDS limit of 2000 mg/L (Utah Rule R309–200, 2006), while California has a standard of 1000 mg/L TDS (California Code of Regulations, 2007), and Florida has a standard of 500 mg/L TDS (F.A.C., 2007). The WHO and the Gulf Drinking Water standards recommend a drinking water standard of 1000 mg/L TDS for drinking water (Fritzmann et al., 2007). Australia has a drinking water standard of 1000 mg/L TDS (Australian Drinking Water Guidelines, 2004). The European Union does not have a drinking water standard for TDS, although standards for other drinking water contaminants have been established (WHO, 1970). In comparison to governmental standards, most desalination facilities are designed to achieve a TDS of 500 mg/L or less (Gaid and Treal, 2007; Petry et al., 2007; Sanz et al., 2007; Xu et al., 2007). Desalinated water used for other purposes, such as crop irrigation, may have a higher TDS concentration; irrigation water standards often include

concentration limits for TDS, chloride, sodium, and boron. Depending on the type of crop, the chloride standard can range from 350 mg/L to more than 2000 mg/L (Fipps, 2003).

The feed water salinity for desalination facilities ranges from approximately 1000 mg/L TDS to 60,000 mg/L TDS, although feed waters are typically labeled as one of two types: seawater or brackish water. Although most seawater sources contain 30,000–45,000 mg/L TDS, seawater reverse osmosis membranes are used to treat waters within the TDS range 10,000 – 60,000 mg/L. Brackish water reverse osmosis membranes are used to treat water sources (often groundwater sources) within a range of 1000–10,000 mg/L TDS (Mickley, 2001). The feed water type can dictate several design choices for a treatment plant, including desalination method, pretreatment steps, waste disposal method, and product recovery (the fraction of influent water that becomes product).

Desalination processes fall into two main categories, thermal processes or membrane processes. Thermal desalination (distillation) has been used for hundreds of years to produce fresh water, but large-scale municipal drinking water distillation plants began to operate during the 1950s (Gleick, 2006). Countries in the Middle East pioneered the design and implementation of seawater thermal desalination, first using a process called multi-effect distillation (MED) and later using a process called multi-stage flash (MSF) distillation (Van der Bruggen and Vandecasteele, 2002). Today, the Middle East collectively holds 50% of the world's desalination capacity (Henthorne, 2003) and primarily uses MSF technology. While thermal desalination has remained the primary technology of choice in the Middle East, membrane processes have rapidly developed since the 1960s (Loeb and Sourirajan, 1963) and now surpass thermal processes in new plant installations. Outside of the Middle East, new RO desalination installations have

been steadily increasing; in 2001, 51% of new installed desalination capacity used RO desalination, and in 2003, RO desalination accounted for 75% of new production capacity (Wolfe, 2005). Countries in the Middle East continue to use thermal desalination due to easily accessible fossil fuel resources and the poor water quality of the local feed water. Water bodies such as the Persian Gulf and the Gulf of Oman have extremely high salinities, high temperatures, and high-fouling potential for membrane systems. At high salinities and high recoveries (55,000 mg/L TDS and above 35% recovery), the pressure required for membrane desalination can be greater than the maximum allowable pressure of membrane modules, and thermal desalination must be used (Kim et al., 2007; Mandil et al., 1998). High feed water temperatures and foulants can also cause problems in membrane desalination that can be avoided by using thermal desalination.

Reverse osmosis (RO), nanofiltration (NF), and electrodialysis (ED) are the three membrane processes available for desalination. ED membranes operate under an electric current that causes ions to move through parallel membranes and are typically only used for brackish water desalination (Reahl, 2004). NF membranes are a newer technology developed in the mid-1980s (Singh, 1997) and have been tested on a range of salt concentrations (Hilal et al., 2005; Tanninen et al., 2006; Wang et al., 2005, 2006). Research has shown that NF, as a singular process, cannot reduce seawater salinity to drinking water standards, but NF has been used successfully to treat mildly brackish feed water (Bohdziewicz et al., 1999; Lhassani et al., 2001; M'nif et al., 2007). Coupled with RO, NF can be used to treat seawater (Hamed, 2005; Hassan et al., 1998; Hilal et al., 2005). In particular, NF membranes are used to remove divalent ions, such as calcium and magnesium that contribute to water hardness, as well as dissolved organic material (Choi et al., 2001; Gorenflo et al., 2002; Wilf, 2003).

RO membranes, however, are able to reject monovalent ions, such as sodium and chloride. Today, seawater RO membranes have salt rejections greater than 99% (Bates and Cuozzo, 2000; Brehant et al., 2003; Reverter et al., 2001); some membranes, when operated under standard test conditions (32,000 mg/L NaCl, 5.5 MPa, 25 °C, pH 8, 8% recovery), can achieve as high as 99.7–99.8% salt rejection (Hydranautics, 2007; Reverber and Gorenflo, 2007). RO membrane technology has developed for both brackish and seawater applications. Brackish water RO membranes typically have higher product water (permeate) flux, lower salt rejection, and require lower operating pressures (due to the lower osmotic pressures of less saline waters), while seawater RO membranes require maximum salt rejection. Membranes designed for higher salt rejection, have lower permeate fluxes, due to the trade-off between membrane selectivity (salt rejection) and membrane permeability (permeate flux). In addition, seawater RO membranes must operate at higher pressures to compensate for the higher osmotic pressure of seawater.

2. History of desalination

In the modern world, desalination first began to be developed for commercial use aboard ships. Distillation, the process of

using a heat source to separate water from salt, was used to provide drinking water to ocean-bound ships to avoid the possibility of depleting onboard fresh water supplies (Seigal and Zelonis, 1995). Thermal desalination enabled ships to travel farther for longer periods of time because it was no longer necessary to transport all the fresh water required for the voyage. In the 17th century, Japanese sailors used a simple distillation technique where water was boiled in pots, and bamboo tubes were used to collect the evaporated water (Desalination in History, 2005). Eventually distillation units were developed to provide makeup water for steam ship boilers; countries began to develop advanced distillation technology in the late 18th century, including investigations into chemical addition. Some of the first attempts at commercial desalination plants include those installed in Tigne, Malta in 1881 and in Jeddah, Saudi Arabia in 1907 (Desalination in History, 2005). Although Jeddah's first desalination plant, nicknamed 'al Kindasah' (the local pronunciation of "condenser"), did not produce much water (Saudi Water and Power Forum, 2007), the effort paved the way for modern facilities first installed in Jeddah in the 1970s. Today, Kindasa Water Services operate two seawater RO facilities in Jeddah, with a total production capacity of 40,500 m³/day (Kindasa Water Services, 2007; Pearce et al., 2004).

The first countries to use desalination on a large scale for municipal drinking water production were in the Middle East. Seawater distillation plants were first developed in the 1950s, and in the 1960s, the first industrial desalination plant opened in Kuwait. Membranes then began to enter the desalination market, and the first successful RO plants used brackish water as the feed (Amjad, 1993) in the late 1960s. In the following decade, membrane material improvements increased product permeability, and RO membranes were then applied to seawater desalination (Van der Bruggen and Vandecasteele, 2002). Over the past 40 years, dramatic improvements in RO membrane technology elevated RO to be the primary choice for new desalination facilities. The worldwide desalination capacity (distillation and membrane processes combined) is shown in Fig. 1 as a function of time over the past 60 years.

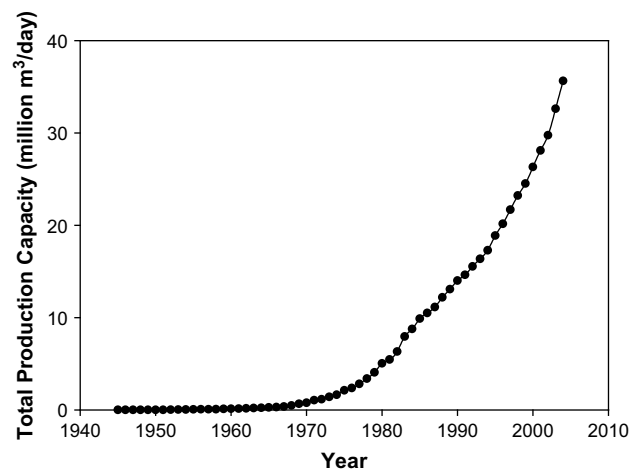


Fig. 1 – Total worldwide installed desalination capacity since 1945, including plants that are operating, built but not operating, and built but shut down (Gleick, 2006).

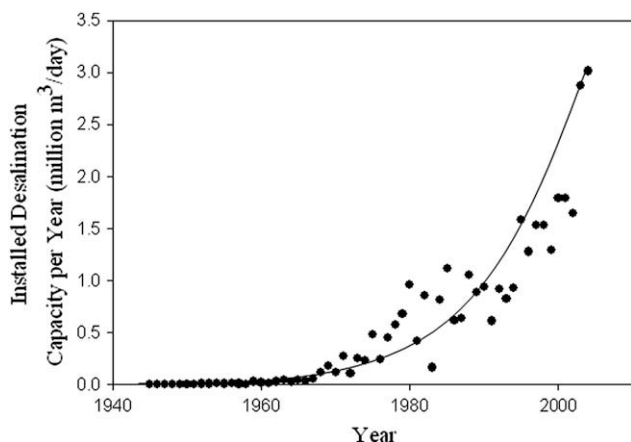


Fig. 2 – New installed desalination capacity each year worldwide from 1945 to 2004 (Gleick, 2006).

From the late 1960s and 1970s, developments in both distillation and membrane technology led to exponential growth in world desalination capacity.

Today, over 15,000 desalination plants are in operation worldwide, and approximately 50% of those are RO plants. The Middle East holds approximately 50% of the world’s production capacity (and 2.9% of the world’s population) and has forged ahead as the leader in large-scale seawater desalination. In 2005, Israel opened the world’s largest seawater RO desalination plant, with a production capacity of 330,000 m³/day, or 100 million m³/yr (Sauvet-Goichon, 2007). The United Arab Emirates (UAE) opened its Fujairah desalination plant in 2005; the plant combines MSF and RO technology to produce 454,000 m³/day of fresh water (Sanz et al., 2007). The yearly worldwide new desalination production capacity (distillation and membrane processes) is shown in Fig. 2. The annual increase in new installations corresponds to technology advances in the late 1960s and 1970s; large jumps in new capacity during recent years signify a new trend in seawater desalination plants with productions of 100,000 m³/day or more.

Saudia Arabia is currently the world leader with approximately 26% of global production capacity, and the United States ranks second, with 17% of the world’s desalination production (Gleick, 2006; Miller, 2003; Wangnick/GWI, 2005; Wolff, 2006). In addition, six of the 11 countries with the greatest desalination production capacity are located in the

Middle East (Miller, 2003). However, statistics on production capacity only touch the surface of desalination use. For example, in Saudia Arabia, thermal desalination is the typical process choice, and most plants are coastal seawater desalination plants. In the United States, 69% of plants use reverse osmosis and only 7% of desalination plants use seawater (Shoaiba Desalination Plant, 2003; Wolff, 2006). In addition, the U.S. represents 4.5% of the world’s population, while Saudi Arabia represents just 0.4%. While only 20% of the total number of desalination plants worldwide use thermal technologies, 50% of the desalination production capacity can be attributed to thermal processes (Frenkel, 2000). Such differences illustrate the wide applicability of desalination to countries with very different resources and water needs.

The distribution of desalination production capacity for different separation technologies is shown in Fig. 3 for the entire world, the United States, and Saudia Arabia. Membrane processes include reverse osmosis (RO), electrodialysis (ED), and nanofiltration (NF), and distillation processes include vapor compression (VC), multi-stage flash (MSF), and multiple effect distillation (MED). The statistics for the world (Fig. 3a) show membrane and distillation processes equally sharing production capacity, with RO dominating the membrane processes and MSF dominating distillation. However, the statistics change dramatically when the number of plants is considered; RO membrane plants represent 80% of the number of desalination plants worldwide, with thermal processes representing just 20% (Frenkel, 2000). In addition, in Saudi Arabia (Fig. 3c), more than 86% of production is achieved using MSF technology, while in the United States, RO is the dominant desalination technology, with membrane processes (RO + NF) representing 84% of the country’s desalination capacity.

Many other countries have begun to utilize desalination for drinking water production, but no other region of the world has implemented desalination on as large a scale as the Middle East. Spain has been using desalination since 1964 (Graber, 2006) to provide drinking water in the Canary Islands, the Balearic Islands, and along the southern and eastern coasts (Reverter et al., 2001; Rybar et al., 2005). Spain and Italy hold the majority of the European desalination capacity, with each country holding 2.6% of world production capacity; 69% of desalination plants in Spain use RO technology, while only 20% of plants in Italy use RO (Miller, 2003). Japan holds 3.7% of global production (Miller, 2003) and has been using seawater

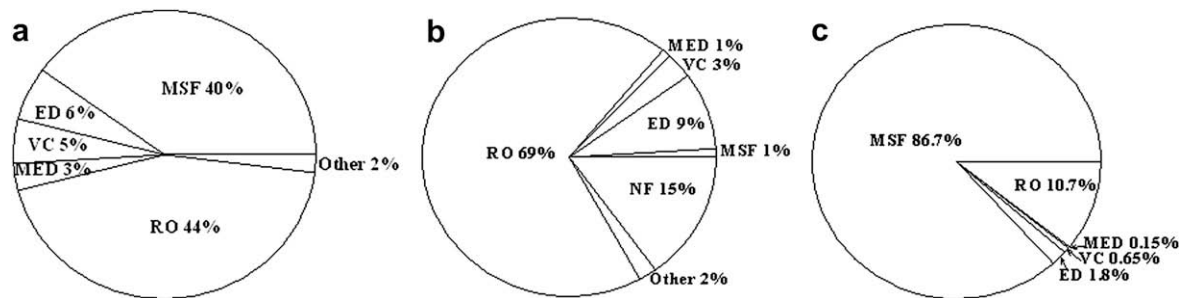


Fig. 3 – Distribution of desalination production capacity by process technology for (a) the world, (b) the United States, and (c) the Middle East (countries include Saudi Arabia, Kuwait, United Arab Emirates, Qatar, Bahrain, and Oman) (Murakami, 1995; Wolff, 2006; Zhou and Tol, 2005).

RO technology since 1974 (Magara et al., 2000). China has seen an explosion in population, along with modern development and industrialization, with little control over or protection of water resources. As a result, the current water transfer strategy (from Southern water resources to water-poor Northern China) is no longer sufficient (Zhou and Tol, 2004), and the country is beginning to investigate desalination technologies (Xu et al., 2007; Zhou and Tol, 2004). In particular, a feasibility study is currently underway for a thermal desalination (MED) plant coupled with a nuclear power plant in the Yantai region (Uranium Information Centre, 2007). Countries in North Africa and the Middle East, such as Algeria, Tunisia, and Jordan, have limited fresh water resources and have investigated using both brackish and seawater desalination (Afonso et al., 2004; Bouchekima et al., 2001; M'nif et al., 2007; Mandil and Bushnak, 2002; Walha et al., 2007). The world's largest brackish water RO desalination plant was finished in 2006 in Wadi Ma'in in Jordan, operating at 129,000 m³/day, with a maximum capacity of over 150,000 m³/day (Mohsen, 2007). Algeria plans to increase its number of plants from 10 to 43 by the year 2019, with a production goal of 2 million m³/day; in 2007, the largest RO desalination plant in Africa started production (200,000 m³/day) in Algeria's capital city, Algiers (Mooij, 2007). Countries in South America, such as Chile, have recently implemented large desalination plants (Petry et al., 2007), and Australia has been battling a water crisis with new RO installations from Melbourne to the Gold Coast (Degremont, 2005; Veolia, 2006; Veolia, 2007). England will construct its first desalination plant in East London, using the Thames Estuary as the brackish water source; production of drinking water is planned to start in 2009 (Thames Water Desalination Plant, 2007).

Although membrane and distillation processes equally share current desalination production capacity worldwide, RO has emerged as the leader in future desalination installations. RO will be the key to increasing water supplies for drinking water production throughout the world. Although wealthy Middle Eastern countries have been able to afford distillation processes, RO technology can now produce fresh water (from seawater) at one-half to one-third of the cost of distillation (Miller, 2003). Brackish water desalination is even less expensive than seawater desalination.

3. Reverse osmosis: basic principles

RO membranes do not have distinct pores that traverse the membrane and lie at one extreme of commercially available membranes. The polymer material of RO membranes forms a layered, web-like structure, and water must follow a tortuous pathway through the membrane to reach the permeate side. RO membranes can reject the smallest contaminants, monovalent ions, while other membranes, including nanofiltration (NF), ultrafiltration (UF), and microfiltration (MF), are designed to remove materials of increasing size, as indicated in Fig. 4. UF and NF membranes are also categorized by the molecular weight cut off (MWCO) of the membrane, or the molecular weight where the membrane will retain 90% of the solute in solution. The general MWCO ranges for UF and NF are 2000–500,000 Da and 250–2000 Da,

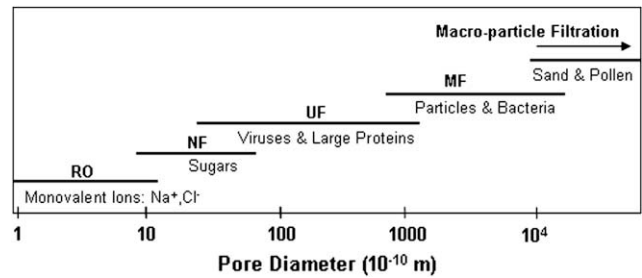


Fig. 4 – Range of nominal pore diameters for commercially available membranes (Perry and Green, 1997).

respectively. MF is usually characterized by a nominal pore size (0.05 μm–10 μm) or by the membrane's rejection (90% rejection of a specific size in μm).

Membranes can be used in either dead-end or crossflow filtration. RO membranes are typically operated in crossflow mode and are most commonly available as spiral wound modules, where the membrane sheets are wound around an inner tube that collects the permeate (Baker, 2004). Most membranes allow filtration through pore flow, where the fluid is forced through the membrane by a positive hydrostatic pressure. The fluid flow depends upon the membrane porosity, the fraction of membrane volume that is void space and can contain liquid, and tortuosity, the distance a molecule must travel through the membrane divided by the thickness of the membrane. Fluid flux through membranes also occurs due to diffusion. The general relationship that describes transport due to pore flow and diffusion can be expressed as follows (Bird et al., 2002):

$$N_{Ax} = \frac{\rho_A \kappa}{\mu} \frac{dp}{dx} - D_{AB} \frac{d\rho_A}{dx} \quad (1)$$

where N_{Ax} is the mass flux of A in the x-direction (perpendicular to the membrane surface), ρ_A is the mass density of A, κ is the permeability, μ is the viscosity, dp/dx is the pressure gradient in the x-direction, and D_{AB} is the binary diffusion coefficient for the diffusion of A in B (the membrane). For MF and UF membranes, the diffusion term is negligible compared to the convection term. Solvent transport through NF membranes occurs through a combination of convective flow and diffusion (Bowen and Welfoot, 2002; Otero et al., 2008), while recent studies show that solute transport through NF membranes is primarily controlled by diffusion (Kedem and Freger, 2008).

Transport through RO membranes, however, is controlled by diffusion, and no open channels exist for pore flow; the RO transport mechanism has been termed solution-diffusion (Lonsdale et al., 1965; Merten, 1963; Paul, 1972, 2004; Wijmans and Baker, 1995). In the solution-diffusion model, water transport across an RO membrane occurs in three separate steps: absorption onto the membrane surface, diffusion through the thickness of the membrane, and desorption from the permeate surface of the membrane. Once a water molecule has absorbed onto the membrane surface, the water concentration gradient (of the water-membrane system) across the membrane causes the water molecules to diffuse down the concentration gradient to the permeate side of the

membrane. The water molecule then desorbs from the membrane and becomes part of the bulk permeate. A complete development and explanation of the solution-diffusion model for transport through RO membranes can be found elsewhere (Lonsdale et al., 1965; Paul, 2004). An RO membrane is operated by achieving a hydrostatic pressure greater than the osmotic pressure of the solution. The positive difference in pressure creates a chemical potential difference (concentration gradient) across the membrane that drives the liquid through the membrane against the natural direction of osmosis (the movement of water molecules from an area of high concentration to an area of low concentration), while the salts are retained and concentrated on the influent surface of the membrane. Some salt passage through the membrane does occur; salt passage for the same membrane increases with salt concentration and temperature. Mass transport through RO membranes can be described as follows:

$$N_A = L(\Delta p - \Delta\pi) \quad (2)$$

where N_A is liquid (water) flux through the membrane, L is the permeability coefficient, Δp is the transmembrane pressure difference, and $\Delta\pi$ is the osmotic pressure difference between the influent and the product water (permeate). The osmotic pressure, π , depends on the solution concentration and the solution temperature. For a thermodynamically ideal solution, the relationship is described as follows:

$$\pi = CRT \quad (3)$$

where C is the ion concentration (molar units), R is the ideal gas constant, and T is the operating temperature.

The permeability coefficient, L , depends on characteristics of the membrane and is described by (Wijmans and Baker, 1995):

$$L = \frac{DSV}{RTl} \quad (4)$$

where D is the water diffusivity, S is the water solubility, V is the water partial molar volume, R is the ideal gas constant, T is the operating temperature, and l is the membrane thickness. This definition of L is based on the solution-diffusion model of water transport across a RO membrane (Bird et al., 2002).

The osmotic pressure of seawater is typically 2300–2600 kPa and can be as high as 3500 kPa (Perry and Green, 1997; Sagle and Freeman, 2004). Osmotic pressures of brackish water are much smaller than those of seawater; for a concentration range of 2000–5000 mg/L, the osmotic pressure ranges from 100 to 300 kPa (Sagle and Freeman, 2004). The osmotic pressure, π , in the concentrate is related to the recovery, R_w , by (Perry and Green, 1997):

$$\pi_{\text{concentrate}} = \left(\frac{1}{1 - R_w} \right) \quad (5)$$

To overcome the osmotic pressure, feed pressures in seawater applications range from 6000 to 8000 kPa, whereas those in brackish water are 600–3000 kPa.

Recovery is an important indicator of RO performance. The recovery of a membrane or an overall RO system is given by:

$$R_w = \frac{Q_p}{Q_f} \quad (6)$$

where Q_p is the permeate volumetric flow rate and Q_f is the feed volumetric flow rate (Rahardianto et al., 2007). Reverse osmosis recovery varies from 35% to 85%, depending on feed water composition, feed water salinity, pretreatment, concentrate disposal options, and optimum energy design configuration. Slight changes in recovery can significantly affect the overall cost of the RO system, as well as the extent of typical limiting factors, such as osmotic pressure, fouling propensity, and mineral scaling (Morenski, 1992; Wilf and Klinko, 2001).

An increase in recovery requires an increase in feed pressure and an increase in permeate flux; increased membrane area may also be necessary to optimize the higher recovery. When permeate flux increases, permeate salinity decreases due to a dilution increase (Wilf and Klinko, 2001). However, operating an RO module at a higher permeate flux often results in flux decline, and operating an RO module at higher recovery without an increase in flux causes an increase in salt passage (Wilf and Klinko, 2001). During RO operation, concentration polarization occurs at the surface of the membrane where dissolved ions accumulate in a thin layer of the feed water; concentration polarization is the ratio of the salt concentration at the membrane surface and the salt concentration in the bulk solution (Kim and Hoek, 2005; Song and Elimelech, 1995). At any recovery, concentration polarization causes greater salt permeation through the membrane than what would be expected based on the bulk solution salinity. When a membrane module is operated at higher recovery, the concentrate or reject stream becomes more concentrated, thus increasing the concentration at the membrane surface. As the salinity increases at the membrane surface, the local osmotic pressure increases as well. Consequently, the overall pressure difference between the hydrostatic pressure and the osmotic pressure decreases, decreasing permeate flow, and the increase in salinity at the membrane surface increases salt transport through the membrane. In addition, phenomena such as salt precipitation and fouling can increase due to the higher local salinity. Models are available to calculate the actual salt concentration at the RO membrane surface (Bacchin et al., 2002; Gekas and Hallström, 1987; Kim and Hoek, 2005; Sutzkoever et al., 2000; Zydney, 1997) and predict precipitation.

RO membrane performance can also be measured by salt flux through the membrane, but it is more often measured by salt rejection. Salt flux is a function of salt concentration, and salt transport occurs from a region of high salt concentration to a region of low salt concentration. Salt flux is described by (Baker, 2004):

$$N_s = B(C_{\text{feed}} - C_{\text{permeate}}) \quad (7)$$

where N_s is the salt flux across the membrane, B is a constant (similar to L in the water flux equation) that depends on membrane characteristics, C_{feed} is the ion concentration in the feed solution, and C_{permeate} is the ion concentration in the permeate. B is described by:

$$B = \frac{D_s K_s}{l} \quad (8)$$

where D_s is the salt diffusivity through the membrane, K_s is the salt partition coefficient between the solution and membrane phases, and l is the membrane thickness.

Membrane salt rejection is a measure of overall membrane system performance, and membrane manufacturers typically state a specific salt rejection for each commercial membrane available. Salt rejection through an RO membrane (crossflow operation) is nominally given by:

$$R_s = \left(1 - \frac{C_{\text{permeate}}}{C_{\text{feed}}}\right) \times 100\% \quad (9)$$

However, RO membranes are typically packed in a spiral wound element, where several membranes are wound around a central tube and separated by spacers. In a spiral wound element, the feed becomes increasingly concentrated from the beginning to the end of the tube and the salt rejection is described by:

$$R_s = \left(1 - \frac{C_{\text{permeate}}}{\frac{C_{\text{feed}} + C_{\text{concentrate}}}{2}}\right) \times 100\% \quad (10)$$

where $C_{\text{concentrate}}$ is the ion concentration in the concentrate (Bartels et al., 2005). When membranes are tested using dead-end operation, Eq. (9) becomes:

$$R_s = \left(1 - \frac{C_{\text{permeate}}}{C_{\text{concentrate}}}\right) \times 100\% \quad (11)$$

RO membranes achieve NaCl rejections of 98–99.8% (Bartels et al., 2005), while NF membranes exhibit rejection values greater than 90% for multivalent ions and between 60 and 70% for monovalent ions (Choi et al., 2001; Hilal et al., 2004). NF salt rejection, particularly for monovalent ions, is highly dependent on the total dissolved solid (TDS) concentration and the presence of other ions (Hilal et al., 2005; Lhassani et al., 2001; Wang et al., 2005).

While membrane manufacturers offer high salt rejection membranes for RO plants, the membranes do not retain the initial salt rejection throughout the membrane's lifetime (up to 7 years with effective pretreatment). Normal membrane aging causes the salt passage (salt passage % = 100 – R_s) to increase approximately 10% per year (Wilf and Klinko, 2001), and other factors, such as temperature, salinity, target recovery, and cleaning methods, can also affect salt passage.

As temperature increases, both the water and the salt permeability increase. Changes in temperature can have a negative or positive overall effect on the RO system. For example, at temperatures below 30 °C, a feed water temperature increase allows the system to operate at a lower feed pressure (or at the same feed pressure and a higher permeate flux) (Wilf and Bartels, 2004). However, the effect of temperature increase on feed pressure decrease is nonlinear. Further increases in temperature increase the osmotic pressure (thus increasing the required feed pressure) and can adversely affect the power consumption if a second RO stage is required to meet permeate quality standards. Typically, RO plants are operated at constant permeate flux, and the permeate salinity varies proportionally with temperature. Even if a plant is operated at constant pressure (increasing permeate flux with increasing temperature), the permeate salinity will increase with temperature due to a greater increase in salt permeability than water permeability.

Increases in feed water salinity increase membrane salt passage. The salt passage is affected by both the TDS

concentration and the composition of bivalent ions in solution, due to interactions between the ions and the membrane surface. RO membranes have an overall negative surface charge and repel negatively charged ions or molecules (Zhao et al., 2005). As negative ions are repelled, more cations than anions are present near the membrane surface; this phenomenon creates an electric potential known as the Donnan potential (Bartels et al., 2005; Tanninen et al., 2006). The Donnan potential helps repel ions from the membrane, but an increase in salinity or divalent ions decreases the Donnan potential effect on membrane salt rejection. The magnitude of the change in salt rejection on specific membranes can vary greatly depending on water composition and membrane charge strength. Bartels et al. found an increase in salt passage (for several different brackish water RO membranes) from approximately 0.4% to between 1.2% and 4% for a salinity increase of 1000 mg/L NaCl–10,000 mg/L NaCl.

The membrane surface charge will increase with increasing pH, with a resulting increase in salt rejection (Yoon et al., 2005). While all RO membranes have an isoelectric point where the overall membrane charge is zero, polar groups within the polyamide membrane material provide local dipoles that allow salt rejection. However, most commercial RO membranes have isoelectric points at relatively low pH values (3–4) when tested with a standard electrolyte solution (0.01 M NaCl or KCl) (Childress and Deshmukh, 1998; Deshmukh and Childress, 2001; Elimelech et al., 1994; Liu et al., 2008) and have a negative surface charge at typical RO operating pH (5–7). Other water components, such as divalent ions and dissolved organic matter (humic acids) can change the isoelectric point (Childress and Deshmukh, 1998; Elimelech et al., 1994).

The salt rejection, R_s , and the recovery, R_w , can be used to calculate the concentration factor (CF) of the concentrate stream (Le Gouellec and Elimelech, 2002; Rahardianto et al., 2006; Shih et al., 2005):

$$CF = \left(\frac{1}{1 - R_w}\right) [1 - R_w(1 - R_s)] \quad (12)$$

CF can also be calculated as the ratio of the concentrate TDS concentration to the feed TDS concentration (C_c/C_f) (Rahardianto et al., 2007). The effect of increasing recovery on concentration factor is shown in Fig. 5, for an assumed salt rejection of 99%. CF increases exponentially as recovery increases; small changes at high recovery can greatly increase the TDS concentration in the concentrate. In particular, a significant difference exists between the CF range for seawater RO compared to that for brackish water RO. When recovery is increased from 35% to 60% (seawater RO), the concentration factor increases slightly from 1.5 to 2.5. However, when recovery is increased from 70% to 90% (brackish water RO), the concentration factor increases more dramatically from 3.3 to 9.9. This difference in concentration factor increase illustrates a key problem found primarily in brackish water RO systems: precipitation of sparingly soluble salts (CaCO_3 , CaSO_4 , BaSO_4 , SrSO_4 , silicates). CF is a useful indication of the overall concentrate salinity, but it does not enable direct comparison of specific ions or salts (particularly salts that form problematic precipitates) between different concentrates.

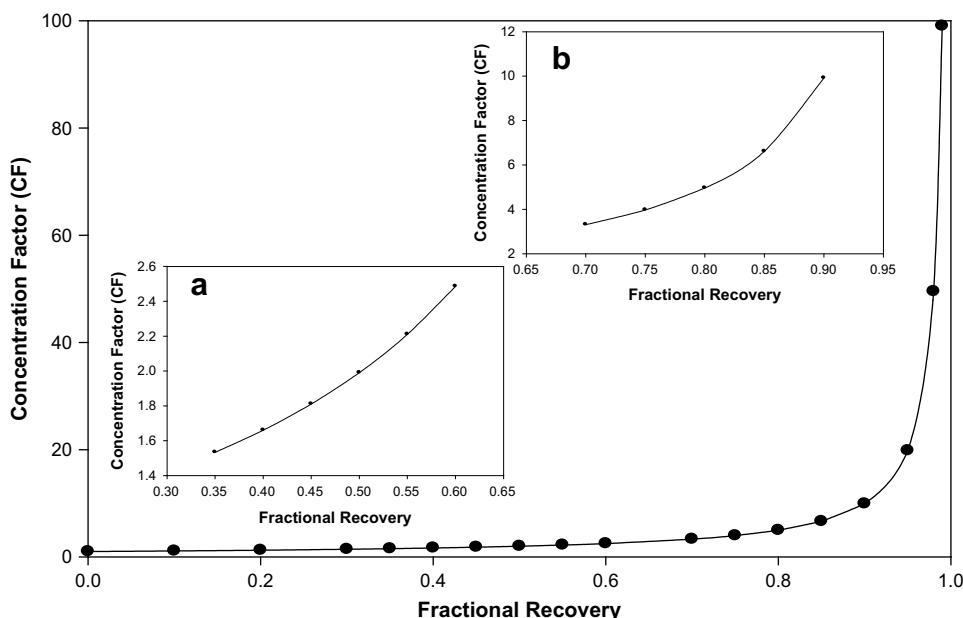


Fig. 5 – The effect of increasing recovery on concentrate concentration factor ($R_s = 99\%$). Insets: (a) the typical range of recoveries for seawater RO operation and (b) the typical range of recoveries for brackish water operation.

Reverse osmosis desalination plants typically operate using 1–4 passes (the permeate of one is the feed to the next RO in series) or stages (the concentrate is the feed to the next RO in series) (Afonso et al., 2004; Atwater et al., 1995; Petry et al., 2007; Reverberi and Gorenflo, 2007; Rybar et al., 2005; Sanz et al., 2007; Sauvet-Goichon, 2007; Singh, 1997). Each pass or stage contains multiple pressure vessels (between 100 and 200 modules in large systems) operating in parallel, with each pressure vessel containing 6–8 membrane elements connected in series (Wilf, 2003). The parallel system of pressure vessels is often referred to as an RO train (Reverberi and Gorenflo, 2007; Sanz et al., 2007; Wilf, 2003; Zidouri, 2000). Recovery and concentration factor can be used to describe each pass or stage of an RO process, as well as the combined RO system (including all passes or stages).

4. Reverse osmosis desalination: feed waters

To illustrate key differences between brackish water and seawater, a comparison of water data is shown in Table 1 (DBHYDRO, 2001; Blavoux et al., 2004; Gaid and Treall, 2007; Jurenka and Chapman-Wilbert, 1996). The seawater source is surface water from the Mediterranean Sea, and both of the brackish water sources are groundwaters. Data for boron concentration in groundwater are limited because boron is a relatively new regulated contaminant, and concentrations are typically low for groundwater. Data from groundwater in California (Boegli and Thullen, 1996) show a boron concentration range of 0.3–0.6 mg/L, which is much lower than that of surface seawater and well below most standards (1.0 mg/L or less) today (Glueckstern and Priel, 2003). The data show obvious differences between the two types of water. Brackish water has a much lower TDS concentration, but the ratio

between TDS and magnesium concentration is approximately the same for all three waters shown. However, the ratios of calcium/TDS, carbonate/TDS, and sulfate/TDS are significantly higher in brackish water than in seawater.

4.1. Seawater RO

Seawater RO plants have two options for feed water source: seawater wells (beach wells) or surface water (open seawater intake). Some of the first plants built in the Caribbean Sea during the 1970s and the 1980s used open seawater intakes and had severe fouling problems on the RO membranes, even with chemical pretreatment (Winters, 1997). In later years, the plants began to use beach wells and achieved improved RO

Table 1 – Comparison of seawater and brackish water sources.

Component	Mediterranean Seawater – Toulon, France (mg/L)	Brackish water – Port Hueneme, USA CA (mg/L)	Brackish water – Martin County, USA FL (mg/L)
Ca ²⁺	440–670	175	179
Mg ²⁺	1400–1550	58	132
Ba ²⁺	0.010	<0.10	0.06
Sr ²⁺	5–7.5	–	26.4
Boron	4.9–5.3	–	–
Na ⁺	12,000	170	905
Cl [–]	21,000–23,000	72	1867
SO ₄ ^{2–}	2,400–2,670	670	384
HCO ₃ [–]	120–142	260	146
TDS	38,000–40,000	1320	3664
DOC	<2	–	1.4

membrane performance (Winters, 1997). The key difference between water sources from open seawater intakes and beach wells is the concentration of organic and particulate material in the water. Similar to fresh water resources (groundwater versus lakes and rivers), water obtained from wells is naturally filtered through porous media (sand, clay, stone, etc.), and much of the organic material typically present in surface waters is removed. Today, as larger and larger RO plants are designed, beach wells cannot always provide enough water, and open seawater intakes are the only feed source option (Bonnelye et al., 2004; Brehant et al., 2003).

Typical seawater concentrations around the world can range from less than 35,000 mg/L to greater than 45,000 mg/L. A summary of several feed water sources and associated TDS concentrations is shown in Table 2 (Gaid and Treall, 2007; Ravizky and Nadav, 2007; Wilf and Klinko, 2001; Zidouri, 2000).

4.2. Brackish water RO

Brackish water sources are often groundwaters; these groundwaters can be naturally saline aquifers or groundwater that has become brackish due to seawater intrusion or anthropogenic influences (e.g., overuse and irrigation). Surface brackish waters are less common but may occur naturally or through anthropogenic activities. Brackish waters can have a wide range of TDS (1000–10,000 mg/L) and are typically characterized by low organic carbon content and low particulate or colloidal contaminants. Some brackish water components, such as boron and silica, have concentrations that can vary widely from source to source; an important factor in brackish water RO system optimization is accurate characterization of the specific feed water.

5. Feed water contaminants

5.1. Seawater RO

Seawater sources often have particulate and colloidal contaminants, as well as hydrocarbons from oil contamination and biological contaminants (algal blooms and other microorganisms).

One of the most difficult seawater components to remove is boron, an inorganic molecule shown to cause adverse reproductive and developmental effects, as well as plant and crop damage (Desotelle, 2001; Magara et al., 1998; Nadav et al.,

2005). In general, ions are rejected better by RO membranes than respective neutral counterparts. Boron naturally exists as boric acid ($B(OH)_3$) in aqueous solution and is typically found in seawater within the concentration range of 4.5–6.0 mg/L (Gaid and Treall, 2007; Glueckstern and Priel, 2003; Magara et al., 2000). Due to a relatively high pK_a , boron ($pK_a = 9.2$ for fresh water, 8.5 for seawater), has limited ion dissociation at neutral or low pH values. In addition, drinking water standards for boron have become increasingly stringent (Glueckstern and Priel, 2003, 2007; Sauvet-Goichon, 2007). Boron ionization and, thus, boron rejection can be increased by increasing the pH of the feed water, but increasing the pH can cause salt precipitation and subsequent membrane scaling (deposition of salt precipitates on the RO membrane). Therefore, boron removal often requires multiple RO stages with different pH values, where the first stage (at lower pH) achieves salt removal and a second stage (at higher pH) achieves boron removal (Glueckstern and Priel, 2003; Sauvet-Goichon, 2007; Wilf and Bartels, 2004).

While a standard seawater RO membrane will reject up to 99.7% of sodium (Na^+) and chloride (Cl^-), operating at neutral pH, the membrane will only reject approximately 75–80% of boron (Glueckstern and Priel, 2003; Magara et al., 2000; Wilf and Bartels, 2004). For a boron concentration of 4.5 mg/L with a rejection of 80% and a recovery of 45%, the boron concentration in the permeate would be 2.0 mg/L. This permeate concentration is at least double the concentration of the minimum boron concentration for many drinking water standards (0.3–1.0 mg/L) (Glueckstern and Priel, 2003; Magara et al., 1998, 2000; Sauvet-Goichon, 2007). Boron rejection increases as pH increases, and can reach 98–99% at pH 11 (Glueckstern and Priel, 2003; Magara et al., 1998); however, even a second pass RO unit treating first pass permeate cannot operate much above pH 10, due to salt precipitation (Glueckstern and Priel, 2003).

5.2. Brackish water RO

Boron can also be a contaminant in brackish water RO systems. As a consequence of the lower general salt rejection of brackish water RO membranes (compared to seawater RO), boron is typically rejected at 65–80% at manufacturer test conditions (25 °C, 15% recovery, pressure of 1030 kPa, 1500 mg/L NaCl feed solution, pH 6.5–7.0) (Glueckstern and Priel, 2003; Hydranautics, 2002). However, in real systems, boron rejection can be as low as 15–20% (Glueckstern and Priel, 2007). Due to this low boron removal by brackish water membranes, the higher overall recovery of brackish water RO systems, and the presence of scaling ions in the permeate, a second RO pass at high pH (>10) is not feasible. Therefore, brackish water RO systems utilize another boron removal strategy: boron-specific ion exchange (Glueckstern and Priel, 2007; Jacob, 2007).

Other contaminants exist in certain water resources, due to either natural occurrence or human pollution. Contaminants such as radionuclides and fluoride naturally exist in some brackish groundwater resources. Human-impacted water sources also have artificially increased levels of nitrates (fertilizers), pesticides (agricultural land use), arsenic (mining operations), and endocrine disruptors (pharmaceuticals in wastewater) (Mickley, 2001). During RO membrane filtration,

Table 2 – Total dissolved salt concentrations for selected salt water bodies around the world.

Water body	TDS concentration (mg/L)
Tampa Bay	18,000–31,000
Pacific Ocean	34,000
Mediterranean Sea	38,000–40,500
Atlantic Ocean	38,500–40,000
Red Sea	41,000–42,000
Gulf of Oman	40,000–48,000
Persian Gulf	42,000–45,000
Dead Sea	275,000

these contaminants are retained in the concentrate, and the concentrate must be treated before disposal. Specific water contaminants can also dictate the type of concentrate disposal used; deep well injection is used for some brackish water RO plants in Florida that have radionuclides in the concentrate.

6. Membrane fouling

Two fouling mechanisms are generally observed for membrane processes: surface fouling and fouling in pores. However, RO membranes do not have distinguishable pores and are considered to be essentially non-porous. Thus, the main fouling mechanism for RO membranes is surface fouling. Surface fouling can occur from a variety of contaminants, including suspended particulate matter (inorganic or organic), dissolved organic matter, dissolved solids, and biogenic material (Amiri and Samiei, 2007). In addition, fouling can develop unevenly through a membrane module or element and can occur between the membrane sheets of a module, where spacers are located to create space for the concentrate stream (Tran et al., 2007). Overall, seawater RO plants, particularly those treating water from an open water intake, are primarily fouled by organic and particulate material, while brackish water RO plants are fouled by dissolved inorganic salts and precipitation. However, both types of RO can experience both general groups of contaminants. In addition, the types of problematic foulants are site-specific, particularly for brackish water RO, and can depend on pretreatment processes.

The capacity of a water to foul RO membranes is often described using the silt density index, or SDI. The SDI of a water is determined from the fouling rate of a 0.45 μm filter at a pressure of 207 kPa (30 psi) and is described in the ASTM standard method D4189 (ASTM, 2007). The equation used to calculate SDI is as follows:

$$\text{SDI} = \frac{100\% \times (1 - t_1/t_2)}{t} \quad (13)$$

where t is the total elapsed flow time, and t_1 and t_2 are the times (in seconds) required to filter 500 mL of water initially and after t minutes, respectively (ASTM, 2007; Wilf and Bartels, 2006). The experiment is setup as a dead-end filtration with continuous flow under pressure, and the membrane is perpendicular to the permeate flow. The total time t is chosen (the standard is 15 min), and the sample flows through the filter during the entire 15 min. As water flows through the membrane, foulants will continuously accumulate on and foul the membrane; therefore, t_1 is expected to be smaller than t_2 . Both conventional and membrane pretreatment lower the SDI of feed water, but each pretreatment choice may have negative and positive aspects (technologically and financially). An SDI of 3 or less (Bonnelye et al., 2004; Reverter et al., 2001; Rybar et al., 2005) is preferred for RO influent. However, many plants tolerate SDI values between 4 and 5, which is often the achievable range through conventional pretreatment (Bonnelye et al., 2004; Bu-Rashid and Czolkoss, 2007;

Chua et al., 2003; Isaias, 2001; Morenski, 1992; Petry et al., 2007).

Kremen and Tanner (1998) showed the relationship between SDI and water fouling propensity by relating the SDI to a total flow resistance. The total flow resistance (R_t) is the combination of two resistances, the resistance of the filter (R_f) and the resistance of the foulant (on the filter) (R_F). The theoretical relationship between SDI and R_t , shown in Fig. 6, displays an exponential relationship between increasing SDI and increasing foulant resistance (or increasing foulant accumulation on the membrane) (Kremen and Tanner, 1998). This relationship indicates far greater fouling resistance between SDI values of 4 and 5 than between SDI values of 1 and 4. Therefore, ideally, a pretreatment scheme that can lower the SDI to below 2 (membranes) will provide a water with a lower fouling propensity than a pretreatment scheme that provides an SDI of 3–5 (media filtration).

An index similar to the SDI, the modified fouling index (MFI), has been developed to better correlate membrane fouling, flux decline, and particle concentration. The original MFI method used a 0.45 μm microfiltration membrane in dead-end filtration and provided a linear correlation between the index and the particle concentration (Schippers et al., 1981; Schippers and Verdouw, 1980). However, the MFI did not always accurately predict the fouling observed in membrane systems, due to the number of small particles that pass through the 0.45 μm membrane. More recently, a modified MFI, the MFI-UF, has been developed (Boerlage et al., 1998, 2002); the MFI-UF uses ultrafiltration membranes to retain a larger portion of the small particles that can pass through microfiltration membranes but will foul an RO membrane. The MFI-UF has subsequently been used to analyze pretreatment performance and RO membrane fouling potential during plant operation (Boerlage et al., 2003).

Turbidity, a measure of the light scatter by particles in solution, is also often reported as a measure of pretreatment efficiency. Measured in NTU (nephelometric turbidity units), turbidity is recommended to be less than 0.2 NTU for successful RO treatment (Wilf and Bartels, 2006). Raw water can have turbidities between 0.1 and several hundred NTU;

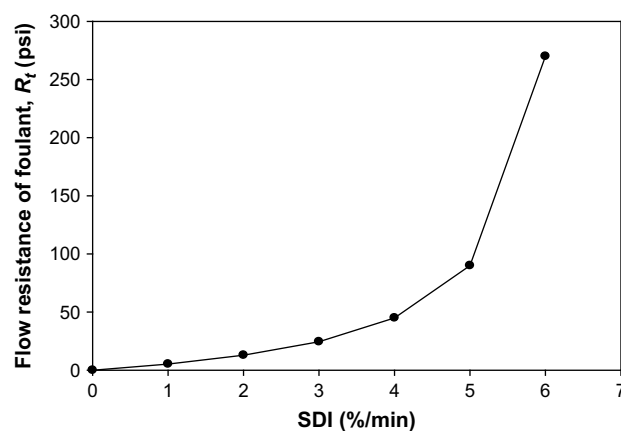


Fig. 6 – The effect of increasing SDI on the permeate flow resistance of water foulant, indicating a higher fouling propensity (Kremen and Tanner, 1998).

most RO plants are reported to have raw water turbidities of below 45 NTU (Bonnelye et al., 2004; Petry et al., 2007; Xu et al., 2007).

Both SDI and turbidity have limitations in predicting the quality and fouling ability of a RO feed water. The SDI test uses a dead-end filtration cell, whereas most commercial RO membrane modules operate in crossflow mode. In addition, the membrane (0.45 μm) used for SDI does not retain contaminants such as biological polymers (Fritzmann et al., 2007); biofouling, often a critical concern for RO operation, cannot be predicted by SDI. In addition, SDI values do not correlate linearly with colloidal or suspended matter, two important foulant groups. Research has shown varying fouling problems that do not necessarily correlate to the SDI value of the feed water (Park et al., 2006; Yiantsios et al., 2005); therefore, SDI can be used as one indicator of a water's fouling potential but should not be relied upon as the sole indicator for fouling (Fritzmann et al., 2007). No direct relationship between SDI and turbidity is possible, although low SDI values often correspond to low turbidity (Wilf and Bartels, 2006).

6.1. Seawater RO

Calcium carbonate is usually the sole problematic precipitate in seawater RO, and often the lower seawater RO recoveries (limited by osmotic pressure) prevent any precipitation problems. Therefore, precipitation is not likely to occur in seawater RO applications (Magara et al., 2000; Reverter et al., 2001), and fouling during seawater RO is primarily caused by particulate matter, organic compounds, and biological growth.

Membrane fouling is caused by the deposition of organic and inorganic water contaminants and can occur in several layers. Suspended and colloidal particles foul a membrane by coagulating together and forming a cake-like layer on the membrane surface, while dissolved organics will interact directly with the membrane surface and with each other to cause fouling (Tran et al., 2007). Colloidal particles are often composed of clay, organics, and metal inorganics, such as aluminum and iron silicates (Amjad et al., 1996). Biological fouling occurs when microbial cells accumulate and attach to surfaces (membranes and spacers), forming biofilms. As membrane fouling occurs, basic membrane functions deteriorate, including salt passage through the membrane, permeate flow, and pressure drop across the membrane (Morenski, 1992). To reverse this fouling, chemical cleaning (acid or base) is used, and operational downtime is often required (Amjad et al., 1996).

6.2. Brackish water RO

The critical fouling problem in brackish water RO systems is salt precipitation and membrane scaling. The higher relative concentrations of calcium, carbonate, and sulfate, combined with the higher recoveries possible for brackish water, cause calcium sulfate and carbonate precipitates to be typical concerns in brackish water RO. An important factor in the membrane fouling potential of dissolved inorganics is concentration polarization. While calcium carbonate is often the primary precipitate of concern, many other salts can be

problematic in brackish water RO. Some groundwater sulfate concentrations, such as those found in southern California (450–670 mg/L) (Leitz and Boegli, 2001), can reach high levels due to farming and fertilization. Calcium sulfate precipitation and membrane scaling have been extensively studied (Hasson et al., 2001; Lee et al., 2003; Öner et al., 1998; Rahardianto et al., 2006; Shih et al., 2004); barium sulfate, strontium sulfate, and silicates, have low solubilities and can become limiting factors in brackish water RO recovery (Rahardianto et al., 2007). However, barium and strontium precipitates tend to be less important because the cations are present in low concentrations, as compared to calcium.

The process of membrane scaling occurs in several stages (Darton, 2000). The first stage of homogeneous precipitation occurs when ions of opposite charge associate and begin to cluster together in large groups (>1000 atoms). In the second stage, the ion clusters begin to form nuclei, characterized by more orderly association and aligning of ions. The third, and final, stage is the growth of salt crystals on the formed nuclei (seed crystals). While the first two stages are reversible, the third stage is irreversible and will continue to occur until the ion concentrations decrease to reach the solubility limit. Heterogeneous precipitation may also occur, where nuclei or ion clusters precipitate associate with suspended or colloidal particles in solution (Boffardi, 1997). In addition, metals such as magnesium, barium, and strontium often coprecipitate when salts such as calcium carbonate precipitate.

Chemicals called antiscalants are used in brackish water RO systems to prevent precipitation. Antiscalants prevent precipitation by disrupting one or more aspects of the crystallization stages. In particular, antiscalants are able to be used at relatively low concentrations (<10 mg/L), where the ion concentrations are stoichiometrically much higher. Antiscalants are effective in increasing the ion concentration threshold required for clustering, as well as disrupting the nuclei ordering and crystal structure. Some antiscalants also will adsorb onto crystal surfaces and repel other ions in solution or fully chelate with dissolved ions. Of all of the possible actions between antiscalants and ions, only the chelation mechanism requires equimolar amounts of ion and antiscalant.

Antiscalants were originally developed in the 1800s for use in boilers and cooling water (Darton, 2000); today, the chemicals have been adapted for use in RO systems. Antiscalants are organophosphonate-, polyphosphate- or polymer-type compounds that are added to the feed water before the feed enters the RO modules. Antiscalants do not completely prevent precipitation at high ion concentrations, and as the salt concentration increases, precipitation will eventually occur.

Antiscalants themselves can become foulants if used at excessive concentrations (Rahardianto et al., 2007); typical antiscalant concentrations in the RO feed do not exceed 35 mg/L and are often less than 10 mg/L (Boffardi, 1997; Hassan et al., 1998; Hasson et al., 2001; Rahardianto et al., 2006; Shih et al., 2004; Vrouwenvelder et al., 2000). Some antiscalants have additional limitations: polyacrylic acid antiscalants will foul membranes in the presence of high iron concentrations, and hexametaphosphate (SHMP) will

eventually hydrolyze in the presence of air, producing inorganic phosphate, possibly leading to calcium phosphate precipitation (Hydranautics, 2003). High concentrations of antiscalants in feed tanks or dosing systems can promote precipitation and biological growth, and the placement of the antiscalant dosing system is critical to avoid unwanted reactions with other chemical additions (Malekar, 2005). Research has shown that biological growth varies with antiscalant type, and some antiscalants can increase biological growth up to 4–10 times the normal rate (Vrouwenvelder et al., 2000).

In addition, chemicals such as alum (potassium aluminum sulfate), ferric chloride, and lime (calcium hydroxide) are used in pretreatment coagulation and can carry through the system and cause RO membrane fouling. Other water components, such as silica, can cause membrane fouling in association with added pretreatment chemicals; aluminum silicates will precipitate during RO operation.

Tran et al. (2007) conducted a study of a spiral wound RO membrane after it had been used for one year in a desalination plant (surface brackish water: 900 mg/L TDS), using advanced analytical and microscopic techniques to determine the composition of the foulant cake deposited on the membrane surface. The results showed that foulants initially deposited on the membrane surface as a thin, amorphous layer (<1 μm thick) containing particulate matter. Layered on top was another amorphous layer ($\sim 3 \mu\text{m}$ thick) containing mostly extracellular polymeric material from organisms and some aluminum silicate. A third and final layer formed in areas where the first two layers were thicker ($\sim 10 \mu\text{m}$ total) and consisted solely of aluminum silicate crystals; in this case, the second amorphous layer contained no aluminum silicate. Further analysis showed high concentrations of calcium, chloride, aluminum, and phosphorus, indicating sparingly soluble salt precipitation (CaCO_3), hindered diffusion or entrapment of dissolved ions, and the presence of pretreatment chemicals (aluminum coagulant and phosphate antiscalant), respectively. The amorphous matrices contained high levels of carbon, oxygen, phosphorus, and aluminum, indicating organic and biologic material and silicate crystals.

7. Membrane cleaning

A combination of acidic and/or basic (alkaline) chemicals is used to clean RO membranes. Common acidic solutions (pH ~ 2) include hydrochloric acid, phosphoric acid, sodium hydrosulfate ($\text{Na}_2\text{S}_2\text{O}_4$) and sulfamic acid ($\text{NH}_2\text{SO}_3\text{H}$), while alkaline (pH ~ 12) chemicals include sodium lauryl sulfate, sodium hydroxide, sodium ethylenediamine tetraacetic acid (Na_4EDTA), and proprietary cleaners (e.g., Permaclean 33) (Bonné et al., 2000; Fritzmán et al., 2007; Reverberi and Gorenflo, 2007). Most cleaning solutions are made from stock chemical solutions to a final concentration of 0.03–2.0% (wt). Membrane cleaning helps restore permeate flux and thus decrease salt passage; Reverberi et al. decreased RO membrane salt passage from 1.9% to 1.2% with alkaline cleaning (0.025% (wt) sodium lauryl sulfate/NaOH, pH 12.5).

8. RO pretreatment for seawater and brackish water

The primary goal of any RO pretreatment system (for seawater or brackish water) is to lower the fouling propensity of the water in the RO membrane system. Surface water resources (seawater and brackish water) typically have a greater propensity for membrane fouling and require more extensive pretreatment systems than groundwater resources (Morenski, 1992). In general, seawater RO tends to use surface water sources, while brackish water RO often uses groundwater sources.

8.1. Conventional pretreatment

Conventional pretreatment typically consists of acid addition, coagulant/flocculant addition, disinfection, media filtration, and cartridge filtration. The first chemical additions, including acid, coagulant, and flocculant, prepare the feed water for granular media filtration (Isaias, 2001; Sauvet-Goichon, 2007). Acid treatment reduces the pH of the feed water (typical pH range 5–7), which increases the solubility of calcium carbonate, the key potential precipitate in many feed waters. The most common acid used to lower feed water pH is sulfuric acid (H_2SO_4) (Bonnelye et al., 2004). However, hydrochloric acid (HCl) is used when sulfuric acid addition has the potential to cause sulfate precipitates (Hydranautics, 2003).

Aqueous particulate and colloidal matter are typically negatively charged, and thus stay separated because like charges repel one another. The role of coagulants is to effectively neutralize like charges and allow the suspended solids to group together in flocs (large groups of loosely bound suspended particles). Therefore, coagulants are typically small, positively charged molecules. Inorganic coagulants are commonly iron or aluminum salts such as ferric chloride or aluminum sulfate, while organic coagulants are typically cationic, low molecular weight (<500,000 Da) polymers (i.e., dimethyldiallylammonium chloride or polyamines) (Sweetwater Technologies, 2006). Aluminum is not as commonly used in pretreatment coagulation prior to membrane filtration due to potential damage to the membrane system. Typical dosing for an inorganic coagulant (5–30 mg/L) is larger than the dose required for a polymer coagulant (0.2–1.0 mg/L) (Wilf and Bartels, 2006). If the feed water is a moderately poor quality water and does not require flocculation and sedimentation, inline coagulation can be used just prior to media filtration (Morenski, 1992). The primary objective of inline coagulation is to change the surface chemistry of the suspended particles so that they attach well to the media filter. It is also possible to use both types of coagulants together to take advantage of the different characteristics of each coagulant.

When a feed water has a high SDI (greater than 10) (Bonnelye et al., 2004), flocculation is often used with coagulation before media filtration. The process of flocculation and sedimentation is a well-known method of particle removal in water treatment (Morenski, 1992). Flocculants are often high molecular weight ($>1 \times 10^7$ Da), anionic polymers.

Granular media filtration includes materials such as sand, anthracite, pumice, gravel, and garnet (Bonnelye et al., 2004;

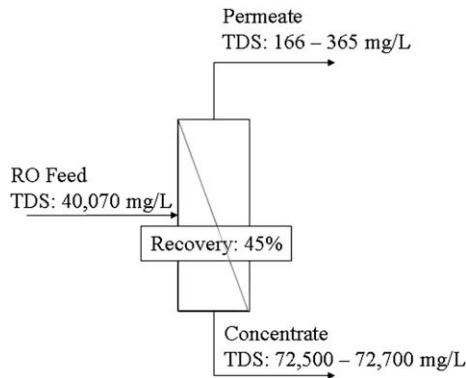


Fig. 7 – Process flow diagram for the one-stage seawater RO plant in Eni Gela, Sicily (Reverberi and Gorenflo, 2007).

a three-year period, the recovery ranged from 46% to 40% (Reverberi and Gorenflo, 2007). Typically, the choice between one or more RO passes depends on several factors including energy cost, feed water characteristics (temperature, composition, and TDS concentration), desired recovery, and product water standards.

The primary energy use in an RO system is the power required to pump the feed water and is directly related to the feed pressure and flow rate (Wilf and Klinko, 2001). The high salt concentrations found in seawater require elevated hydrostatic pressures (up to 7000 kPa); the higher the salt concentration, the greater the pressure and pumping power needed to produce a desired permeate flux. As mentioned previously, the required hydrostatic pressure must be greater than the osmotic pressure on the feed (concentrate) side of the membrane. As the recovery of a RO unit increases, the osmotic pressure increases on the feed side of the membrane, thus increasing the feed pressure required. However, as the recovery increases, the feed flow required decreases (for a specific product flux) (Wilf and Klinko, 2001), and for lower recoveries (35–50%), the overall energy requirement decreases with increasing recovery. Thus, a minimum energy requirement exists, typically at a recovery between 50 and 55% (Wilf and Klinko, 2001), which varies with feed salinity; a case study by Wilf and Klinko (2001) is shown in Fig. 8. If an RO system consists of more than one pass, the energy requirement will be greater due to the increased pressure drop across the subsequent passes.

Energy recovery devices have been developed to help recover some of the energy typically lost from the pumps and membrane system. The primary objective is to recover much of the energy held in the pressurized RO concentrate stream. Before continuing to disposal or treatment, the concentrate is sent through an energy recovery device, and the recovered energy is used to partially power the pumps. Energy recovery devices fall into two general classes (Wang et al., 2004). Class I devices use hydraulic power to cause a positive displacement within the recovery device, and the energy is transferred in one step from hydraulic energy to hydraulic energy. Class II devices use the hydraulic energy of the RO concentrate in a two-step process that converts the energy first to centrifugal mechanical energy and then back to hydraulic energy. While some small- or medium-size plants use Class II devices, such

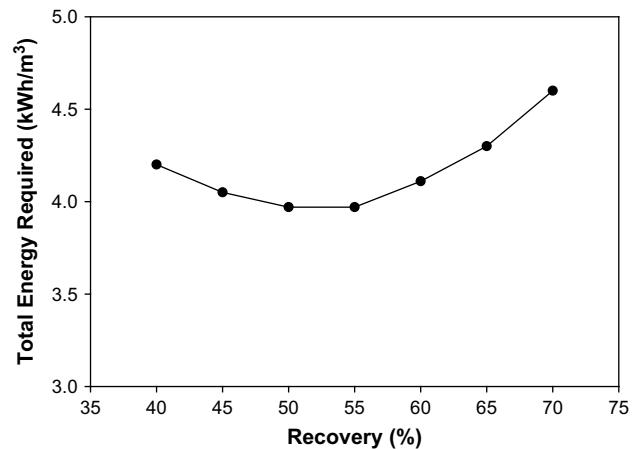


Fig. 8 – Total energy required per volume of permeate produced as a function of RO system recovery (Wilf and Klinko, 2001). (Influent TDS: 34,000 mg/L).

as the Pelton wheel or Pump Engineering's TurboCharger (Harris, 1999), most plants today use Class I devices, such as the DWEER (DWEER Technology, Ltd.), PX Exchanger (ERI), or PES (Siemag's Pressure Exchanger System) commercial products (Andrews and Laker, 2001; Geisler et al., 2001; MacHarg, 2001). The Class I energy recovery devices can provide a net energy transfer efficiency from the concentrate stream to the feed stream of more than 95%.

When an energy recovery device is used to transfer energy back to the RO feed stream, the feed can bypass the main high-pressure feed pump. Instead, a booster pump is used to account for pressure losses in the RO membrane modules, piping, and energy recovery device. The size of the high-pressure pump can then be reduced and used primarily to pump the part of the feed replacing the RO permeate. This type of system design greatly reduces the overall energy consumption and provides separate pumping systems for the permeate and the concentrate (MacHarg, 2001). Booster pumps can also be used between RO passes to recover the pressure lost from the first pass; repressurizing the RO feed allows greater recoveries in the second pass.

Seawater RO recovery is maximized by using only one pass of membranes. However, if a plant is designed to have more than one pass, the overall recovery is typically only a few percent less than the recovery in the first pass (Sanz et al., 2007; Sauvet-Goichon, 2007). Recovery for a two-pass seawater RO system decreases slightly because the feed to the second pass is the permeate from the first pass. A second pass is used to further improve permeate water quality and is operated at high recovery (85–90%).

Increasing recovery is limited in seawater RO systems by the resulting increase in osmotic pressure. Recalling that the hydrostatic pressure is limited to approximately 7000 kPa (Petry et al., 2007; Reverberi and Gorenflo, 2007; Reverter et al., 2001; Rybar et al., 2005), the increase in osmotic pressure limits realistic seawater RO recovery to 55–60%. Wilf and Klinko (2001) evaluated the effect of increasing recovery on feed pressure and concentrate osmotic pressure for several feed water concentrations; the results for a feed TDS

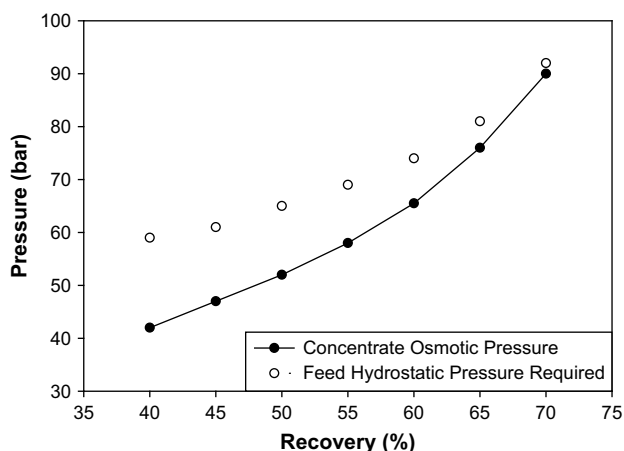


Fig. 9 – The effect of recovery on hydrostatic pressure and concentrate osmotic pressure (Wilf and Klinko, 2001). (TDS: 34,000 mg/L).

concentration of 34,000 mg/L and a permeate flux of 13.5 L/m²-h are shown in Fig. 9. Above 60% recovery, the required feed pressure increases well above 7000 kPa, and a higher feed salt concentration would slide the two curves higher along the ordinate. From 2001 to today, improvements in membrane materials and permeability have allowed recent plants (selected summary shown in Table 5) to operate at lower pressures, while maintaining recoveries above 40%. The seawater RO plants at El Coloso, Chile, and Eilat, Israel, both have higher feed TDS concentrations than the water used for calculations in Fig. 9; yet, both plants operate at 50% recovery and lower pressures than would be predicted by Wilf and Klinko’s calculations.

Feed water characteristics play a large part in the RO system design. For a one-pass RO unit, with a salt rejection of 99.7% and a recovery of 35%, the TDS concentration in the permeate would range from 300 to 400 mg/L. Most RO plants are designed to produce fresh water with less than 500 mg/L TDS, and many plants have target TDS concentrations much lower than 500 mg/L (Gaid and Trear, 2007; Reverberi and Gorenflo, 2007; Sanz et al., 2007; Sauvet-Goichon, 2007). If the product water target concentration for TDS is much below 300–400 mg/L, at least two passes are necessary to achieve the target value. In addition, individual components of a feed water can affect RO system design.

Feed water temperature can vary from 12 °C to 35 °C (Gaid and Trear, 2007; Pearce et al., 2003; Sanz et al., 2007). As

mentioned previously, an increase in feed water temperature will increase water and salt permeability through the membrane. However, most RO plants maintain a constant permeate flux; thus, an increase in temperature will cause an increase in permeate salinity. An increase of 1 °C can increase the salt permeability by 3–5% (Wilf and Klinko, 2001). If high feed water temperatures are expected, multiple passes may be needed to achieve an adequately low product water TDS concentration.

Various design options are available for a multi-pass seawater RO system. The simplest design consists of a two-pass system, where the first pass is a high-pressure seawater RO membrane array and the second pass is a low-pressure brackish water RO “polishing” step. The first RO pass typically operates at a 35–45% recovery, while the second pass will operate at recoveries close to 90%. The seawater RO plant in Fujairah, Saudi Arabia, shown in Fig. 10, operates with a similar setup, except the second RO pass is split into two brackish water RO passes in series (Sanz et al., 2007); the concentrate from the second two passes of the Fujairah plant is recycled back to the RO feed stream, and the overall recovery of the RO system is 41%. The feed water contains 38,000–38,500 mg/L TDS, and the resulting permeate TDS (75–120 mg/L) is lower than that shown for the Eni Gela plant in Fig. 7. Typically, a multi-pass RO system will produce permeate with a lower TDS concentration with only a small loss in recovery.

Another type of two-pass seawater RO system takes advantage of the variation in permeate TDS along a membrane element. As feed water passes through a membrane element, permeate passes through the membranes, and the remaining water, or concentrate, becomes more and more concentrated. As the feed/concentrate TDS increases, the salt passage through the membrane increases, thus increasing the permeate salinity. Therefore, the permeate produced at the beginning (feed entry) of a membrane element has a lower salinity than the permeate produced at the end (concentrate exit) of a membrane element. The alternative two-pass design takes a portion of the higher-salinity permeate as the feed to the second pass, while the lower-salinity permeate is collected directly as product water (Wilf and Bartels, 2004). The RO plant built in Tampa Bay, Florida, USA, takes advantage of this alternate two-pass design (Wilf and Bartels, 2004); the overall power consumption of the RO system is lower because only a portion of the first pass permeate must be pumped to the second pass (Wilf and Bartels, 2004).

More complex multi-pass seawater RO systems exist, such as the system used in Ashkelon, Israel. The Ashkelon plant

Table 5 – Summary of selected single-stage seawater RO desalination plants.

Plant Location	Feed TDS (mg/L)	Recovery (%)	Permeate flux (L/m ² -h)	Feed pressure (kPa)	Reference
El Coloso, Chile	36,500	50	–	5800	Petry et al. (2007)
Las Palmas, Spain	38,000	40–45	11.0–13.5	5500–7000	Reverteri et al. (2001)
Eni Gela, Sicily	40,070	40–46	13.3	6200–6400	Reverberi and Gorenflo (2007)
Gran Canaria, Spain	38,000	42	13.0	5700	Rybar et al. (2005)
Eilat, Israel	41,000	50	13	6100–6300	Glueckstern et al. (2001)

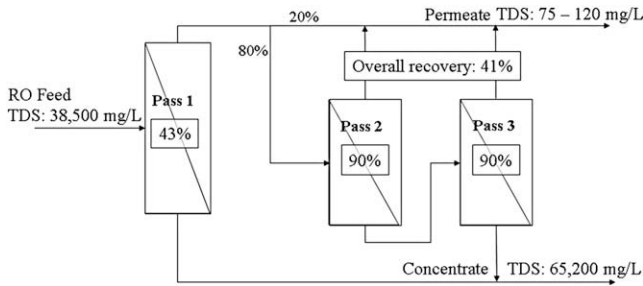


Fig. 10 – Process flow diagram of the seawater RO plant in Fujairah, Saudi Arabia (Sanz et al., 2007). Each box represents one RO pass, and the recovery (in %) is shown for each pass.

uses four RO passes in series to treat seawater from an open water intake on the Mediterranean Sea (40,700 mg/L TDS) (Sauvet-Goichon, 2007). The permeate must be produced with less than 0.4 mg/L boron and 20 mg/L chloride (Sauvet-Goichon, 2007); thus, the series of passes, with changes in pH, was necessary to obtain the required permeate water quality. A process flow diagram of the RO system at Ashkelon is shown in Fig. 11. The first pass has a recovery of 45% and is operated at neutral pH. Permeate from the feed end is collected as product, while permeate from the concentrate end is collected and sent to the second pass. The second pass operates at 85% recovery and high pH to achieve greater boron removal. The concentrate from the second pass continues to the third stage, also operated at 85%, but at low pH. The objective of the third pass is to achieve higher recovery without salt precipitation. However, the boron removal in the third pass is minimal at low pH, and a fourth pass (high pH, 90% recovery) treats the third pass permeate for boron removal. Overall, the recovery is approximately 44%, and the plant uses 25,600 seawater RO membranes and 15,100 brackish water RO membranes (Sauvet-Goichon, 2007).

9.3. Alternative seawater RO plant design

A recent design trend in large seawater RO plants is the “hybrid” plant: a desalination plant that combines the use of

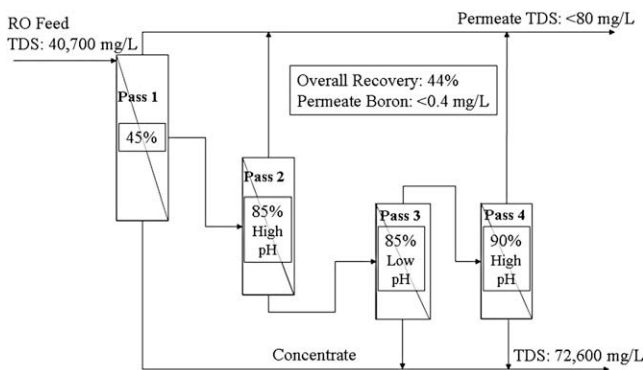


Fig. 11 – Process flow diagram of the seawater RO plant in Ashkelon, Israel (Sauvet-Goichon, 2007). Each box represents one RO pass, and the recovery (in %) is shown for each pass.

several technologies together in the same plant. Plants such as the 454,000 m³/day Fujairah plant in the United Arab Emirates (UAE) use a combination of distillation (MSF) and RO to produce fresh water. In a hybrid plant, the MSF and RO systems operate independently and in parallel (Hamed, 2005). The Fujairah plant uses 5 MSF units connected to an adjacent power plant and a three-pass RO treatment train (Sanz et al., 2007); the overall water production is split 62.5%/37.5%, MSF/RO, with the RO plant producing 170,500 m³/day fresh water. The MSF units receive steam from the power plant, and both the RO and MSF plants have an open water feed intake from the Gulf of Oman (Indian Ocean) (Sanz et al., 2007).

The key advantage of this type of hybrid desalination plant is the fresh water production flexibility of the plant due to the RO component. MSF plants are relatively flexible (changes of 25% production capacity possible), while RO plants can be adjusted by as little as 5% production increments (Wolfe, 2005) because of the modular RO design. The RO component of the Fujairah plant consists of a seawater RO stage (18 trains with a recovery of 43%) and two subsequent brackish water RO stages (8 trains with a recovery of 90%); each stage is split into two separate piping and pumping systems (Sanz et al., 2007) to further increase flexibility. The RO permeate from all three units is blended with the MSF product before distribution; during the first two years of operation, the blend had a TDS range of 75–120 mg/L (Sanz et al., 2007). In addition, the RO component also helps to reduce the overall product water cost. The hybrid plant design is particularly useful in countries such as the UAE, where water demand remains relatively constant throughout the year, but electricity demand varies greatly (Almulla et al., 2005; Hamed, 2005).

Hybrid plants were first used when existing desalination plants needed to be expanded. Saudi Arabia first used hybrid plants in the early 1990s when three existing MSF plants needed additional water production ability (Hamed, 2005); seawater RO plants were built on adjacent property, and the product waters from the MSF and the RO units were blended before distribution. Each MSF plant was already combined with a power plant; thus, the concept of a triple-hybrid power and desalination plant was born. More integrated hybrid systems, where the RO and MSF units are used in series, have been studied as well (Cardona et al., 2002; El-Sayed et al., 1998; Quteishat et al., 2003; Turek, 2002). Possible uses of integrated hybrid systems include blending the RO concentrate with the MSF recycle, using a portion of the heated MSF seawater as RO feed water, and using nanofiltration as a pretreatment to allow increased MSF and RO recoveries (Hamed, 2005). However, the simple hybrid design of MSF and RO units operating in parallel already allows significant advantages in energy and capital cost optimization for the existing MSF plants and balancing of water and power demands (Almulla et al., 2005; Hamed, 2005); the parallel design remains the current hybrid system design of choice (Hamed, 2005).

9.4. Brackish water RO system design

Brackish water RO plants tend to be smaller in production capacity than seawater RO plants, but a greater number of brackish water RO plants (48% of the total number of plants) are in operation worldwide than seawater RO plants (25%)

(Wangnick, 2002). The remaining desalination plants (28%) consist of other feed waters, including rivers, wastewater, and pure water. The feed water for brackish water systems is often groundwater, and groundwater has much lower flow rates into wells than is possible to achieve with an open seawater intake. Many plants produce between 500 and 10,000 m³/day of permeate product (Allam et al., 2002; Jaber and Ahmed, 2004). The production range of brackish water plants in the U.S. is 200–76,000 m³/day, and the range of feed water TDS is 520–8500 mg/L (Mickley, 2001).

The basic system design for brackish water RO is critically different from seawater RO; in brackish water RO, the feed to a second stage is the concentrate from the first stage, whereas in seawater RO, the feed to a second pass is the permeate from the first pass. This key design difference results from the lower TDS concentrations found in brackish water and allows brackish water RO systems to achieve much higher system recoveries. The high TDS of a seawater feed can result in a single-pass RO permeate with a TDS greater than 1000 mg/L; brackish water RO plants do not typically have this problem. Brackish water RO systems often consist of two stages, and each stage has a recovery of 50–60%, achieving an overall system recovery of 70–85%. A third stage may be used in some cases to further increase the recovery or to achieve recalcitrant contaminant removal (e.g., boron) by changing the chemical conditions. Some brackish water RO systems also use NF membranes for salt removal. NF membranes can be used in series following the RO system to treat the RO concentrate and increase system recovery; the RO and NF permeates are blended together as product water (M'nif et al., 2007). Apart from the number of stages, another important decision in brackish water plant design is the method of concentrate disposal.

9.5. Alternative brackish water RO plant design

Recent innovations in brackish water RO plant design have stemmed from a combined need for inland desalination and reduced concentrate production (or increased product water recovery). New regulatory constraints, increasing RO plant design size, and environmental awareness have also influenced interest in alternative concentrate management (Mickley, 2004). The key limiting factor to widespread use of inland desalination is the exorbitant cost of concentrate disposal. The ideal solution would be to further increase brackish water RO recovery, but membrane scaling limits RO systems. High recoveries (95–99%) typically seen in fresh water treatment plants cannot be achieved by RO plants commercially available today.

The United States (U.S.) has been a forerunner in the use of brackish water desalination, although most of the country's RO plants are located near coastal areas, where concentrate disposal is not a large part of the plant costs. The need for alternate concentration management strategies has increased due to the growing size and number of brackish water RO plants, as well as more stringent disposal regulations (Mickley, 2004). In addition, many inland regions of the U.S. are slowly losing (or have already lost) fresh water resources and will need to begin to use brackish water as a drinking water resource. Other countries, such as Egypt, Tunisia, and Jordan,

located in arid climates, but lacking the abundance of energy resources and financing of some other countries, have also started to use brackish water RO (Afonso et al., 2004; Allam et al., 2002; Khalil, 2001; Walha et al., 2007). For these countries, RO is more financially affordable, and using brackish water further reduces the desalination cost.

To increase the brackish water recovery beyond current limitations imposed by the core RO membrane system, various technologies and treatments have been proposed. The general objective of all the options is to treat the concentrate stream or pretreat the RO feed stream to recover more of the water normally considered as waste in the concentrate.

Gilron et al. (2003) compared conventional media filtration pretreatment to pretreatment through compact accelerated precipitation softening (CAPS) for brackish water RO. During CAPS pretreatment, feed water and base (NaOH) are fed to an agitated tank containing 1–3% calcium carbonate solids. The tank also contains submerged filters (8–13 μm pores); the presence of calcium carbonate solids causes a cake to form on the filters. Calcium carbonate precipitation thus occurs both in solution on existing solid particles and at the preformed filter cake. The CAPS method produced RO feed water at a flux of 2000 L/m²-h and removed 92–96% of the calcium from the raw feed water. As a result, the RO unit was able to operate at 80% recovery without acid addition (feed pH = 7.8) and at 88% recovery with acid addition (pH = 7.6). In comparison, media filtration with no pre-RO acid addition achieved a recovery of 72%, and with acid addition, the recovery was 88%. While the specific CAPS design has not been applied to full-scale desalination plants, the general concept of concentrate treatment through salt precipitation is a promising method and a current research focus in brackish water RO.

Another version of accelerated precipitation has been extensively studied and developed by researchers in California (Gabelich et al., 2007; Rahardianto et al., 2007; Williams et al., 2002). The researchers used an interstage precipitation process between two bench-scale brackish water RO units to increase the water's normal 90% recovery to 98% overall recovery. The precipitation process consisted of using either calcium carbonate (calcite) or calcium sulfate (gypsum) seeding, along with pH control, to remove sparingly soluble salts. The precipitation step was carried out in a conical, stirred crystallizer, and the resulting precipitate was allowed to settle, filtered (through fritted glass and a 0.2 μm cartridge filter), and subsequently treated in a second RO module (Rahardianto et al., 2007). While gypsum seeding achieved a calcium removal of only 30%, calcite seeding achieved 92–93% calcium removal in less than 30 min. However, tests without calcite seeding, using only pH elevation, also achieved 92–93% calcium removal, but required greater than 30 min agitation time. Overall, experiments using both synthetic and real water sources (with antiscalant dosing) resulted in 98% overall recovery and no measurable membrane scaling in the second RO module.

Some research has focused on novel combination of existing technologies to increase overall system recovery. Almulla et al. (2002) investigated three different strategies to increase overall recovery from 70–75% to 90–95%, including seawater RO membrane treatment of brackish water RO concentrate, UF treatment of multi-media filtration

backwash, and crystallizer-UF treatment of brackish water RO concentrate. The seawater RO membranes were operated at a feed pressure of 2700 kPa and a recovery of 40%, increasing the overall recovery from 73% to 83% (Almulla et al., 2002). During this brackish water RO concentrate treatment, precipitation of calcium carbonate, magnesium sulfate, and silica dioxide limited the recovery. When UF was used to treat media filter backwash, at least 92% of the backwash water was recovered, and the overall system recovery increased to 85.4%. Finally, the researchers showed, through jar tests, that crystallization could remove significant portions of the silica (83%), calcium (92%), and magnesium (92%) (Almulla et al., 2002); the overall system recovery due to crystallizer-UF concentrate treatment was estimated at 95%.

Another option to traditional concentrate disposal is the treatment of the concentrate for specific salt recovery (Mickleley, 2004). Using the specific makeup of the concentrate, individual salts can be removed in series using pH changes and salt precipitation. The process has been developed by companies such as Geo-Processors Pty Limited (Australia), and has been used successfully to treat concentrate in pilot and commercial applications. In Eilat, Israel, the concentrate from a seawater RO plant is used to produce salt (Ravizky and Nadav, 2007). The feed to the RO plant is a blend of seawater and concentrate from a local brackish water RO plant. In this way, the brackish water RO plant avoids seawater pollution and discharge costs by ultimately using the concentrate to produce a consumer product.

For brackish water resources with elevated boron concentrations, the conventional two-stage RO design is expanded to include additional RO stages or an ion exchange system to achieve boron removal. Although grouped within the general treatment process of ion exchange, the resin used to remove boron will complex with ionized or neutral boron (boric acid) molecules (Jacob, 2007). Therefore, the resin is very specific and can be used at the same neutral or acidic pH designed for the subsequent RO system. No other ions in solution are affected by the boron-specific resin, and a typical ion exchange column will reduce the boron concentration to less than 0.05 mg/L. In comparison, the boron concentration after a first pass seawater RO membrane is between 1 and 2 mg/L, and a second pass at high pH can reduce the boron concentration to below 0.5 mg/L (Jacob, 2007; Sauvet-Goichon, 2007). The boron-specific ion exchange column is typically placed after the RO system (Glueckstern and Priel, 2007; Jacob, 2007); if there are several RO stages, which is common in brackish water RO, boron ion exchange may be used to treat the permeate from all or selected RO stages. The world's first brackish water RO plant using boron ion exchange started operation (in Israel) in May 2006 (Jacob, 2007); as more and more countries follow Israel's lead in adopting strict boron standards for drinking water, the ion exchange technology is likely to be increasingly applied. A RO-ion exchange system proposed by Glueckstern and Priel (2007) is shown in Fig. 12.

Another treatment option for boron removal in brackish water RO systems is specially designed RO membranes for removal at neutral or acidic pH or the use of seawater RO membranes for the second and/or third stages of a brackish water RO system (Glueckstern and Priel, 2007). However, the first membrane option requires more energy, and the second

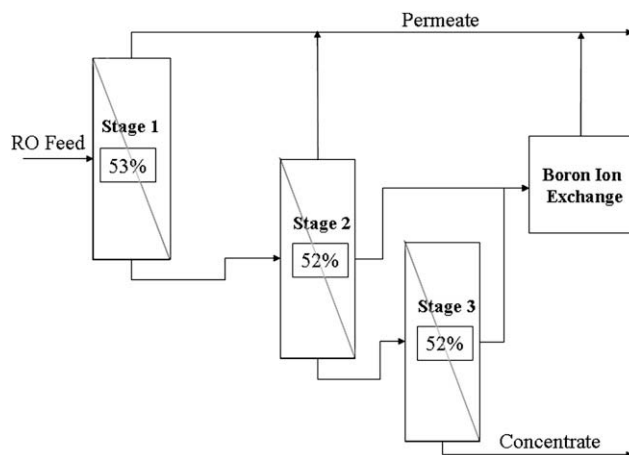


Fig. 12 – A possible configuration for a combined system of brackish water RO and boron ion exchange (Glueckstern and Priel, 2007).

option requires pH elevation (Glueckstern and Priel, 2007); the appropriate technology must be chosen after evaluation of the specific plant design. A design for boron removal in a brackish water RO system using seawater RO membranes in subsequent stages is shown in Fig. 13.

10. RO permeate post-treatment

10.1. Seawater RO

The permeate from seawater RO is often treated before distribution. Depending on the permeate TDS, the permeate may be blended with another water to either increase or decrease the salinity (Sanz et al., 2007; Zidouri, 2000). Lime ($\text{Ca}(\text{OH})_2$) or limestone contactors may be added to increase the hardness, alkalinity and pH, as well as prevent the water from causing calcium to leach from pipes in the distribution system (Khawaji et al., 2007). Hardness is necessary to achieve

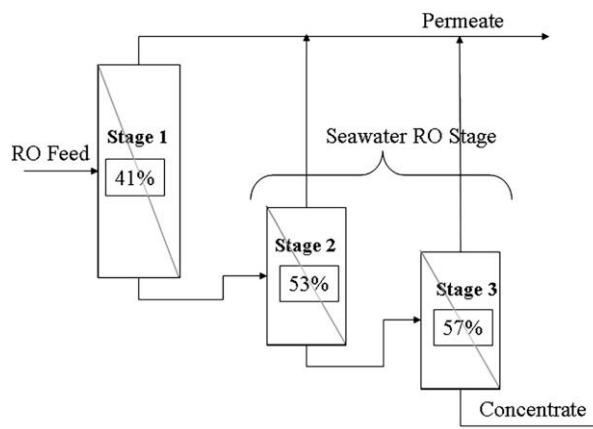


Fig. 13 – Boron removal in a brackish water RO system using seawater RO membranes in the second and third stages.

the typical taste of drinking water and prevent corrosion, while alkalinity is the primary buffering system of all natural waters. Alkalinity (carbonate and bicarbonate) helps to stabilize the water pH during distribution and use. If additional hardness or alkalinity is not needed, another base, such as caustic soda (NaOH), may be used for pH adjustment. The permeate is also disinfected, typically through the use of chlorine, sodium hypochlorite, or chloramines. Permeate treatment may also include fluoride replacement (El-Ramly and Peterson, 1999).

10.2. Brackish water RO

Similar to seawater RO, brackish water RO permeate has extremely low levels of calcium and bicarbonate (alkalinity), and these two components must be replenished before distribution. In brackish water RO systems, a portion of the RO feed water is often blended with the RO permeate to replace hardness and alkalinity, and caustic soda is used for pH adjustment. If feed water blending is not possible, typically due to lower feed water quality, the addition of lime or the use of limestone contactors is often used (Walker et al., 2007; Watson et al., 1995). As with seawater RO permeate, brackish water RO permeate is also disinfected before distribution.

11. RO concentrate disposal

11.1. Seawater RO

For seawater RO plants, the disposal method is usually discharge back into the same body of water; the primary concerns are only the pumping system and length of piping needed to reach the chosen discharge point underwater (Mooij, 2007; Ravizky and Nadav, 2007). The feed water intake and the concentrate discharge are positioned in separate locations, and the concentrate is diluted into the large seawater body without influencing the feed water composition.

11.2. Brackish water RO

If feasible, surface water disposal is the disposal method of choice for brackish water plants due to higher costs of other options. However, surface water discharge is often limited to coastal RO plants; otherwise, the large amount of piping needed to transport concentrate from an inland plant to the sea is financially prohibitive. Some inland brackish water RO plants discharge concentrate to local lakes and rivers. Unlike seawater discharge, brackish water RO concentrate, if discharged to surface water, can change the salinity of the receiving water. The change in salinity can change the concentration of dissolved oxygen in the water and negatively affect aquatic life; the standard limit for surface water discharge is a salinity difference of less than 10% (Mickley, 2004).

If the feed to a brackish water RO plant is groundwater, the plant must often treat the RO concentrate before disposal. Treatment is particularly important if the concentrate is discharged to surface water. Groundwater can have high concentrations of gases, such as carbon dioxide, ammonia,

and hydrogen sulfide, some of which are toxic to aquatic animals (Kimes, 1995; Mickley, 2001). Concentrate treatment can include pH adjustment, chlorination, dechlorination, degasification, and aeration (Mickley, 2001). High levels of fluoride and calcium in RO concentrate have also been shown to be toxic to certain test organisms, although concentrate treatment does not typically remove these contaminants.

With a growing need for inland desalination of brackish water, other disposal options must be used, and thus far, a wide range of methods have been put into practice (Bloetscher et al., 2006; Mickley, 2001, 2004; Van der Bruggen et al., 2003). Each method has advantages and disadvantages, often representing a compromise between cost, local available resources, environmental impact, and technology.

After surface water discharge, concentrate disposal to a combined sewer is often the method of choice. A combined sewer transports concentrate and other wastewaters to a local municipal wastewater treatment plant, where the mix of both waters is treated; some of the salt from the concentrate flow becomes part of the sludge (dewatered solid waste product from the wastewater treatment plant), while the rest remains dissolved and becomes a part of the plant effluent. High salinity can negatively affect biological treatment processes and cause discharge permit violations (Mickley, 2004). Local regulations, the size of the desalination plant, and availability of a nearby wastewater treatment plant often dictate the feasibility of this disposal option. If the concentrate flow is too large or too saline, the desalination plant may not be able to use sewer disposal.

Brackish water desalination plants located near green (parks, golf courses) or agricultural areas can sometimes use the RO concentrate as irrigation water. Although this disposal option is advantageous to the desalination plant, as the concentrate waste becomes a needed product, irrigation with concentrate can cause several problems. The use of saline water for crop irrigation adds salt to the soil and to the local groundwater aquifers (Mickley, 2001). Build-up of the salt in the soil can affect future crop growth, while the groundwater will slowly increase in salinity over time. In addition, high boron concentrations in the irrigation water can cause plant damage (Glueckstern and Priel, 2003). While a useful concentrate disposal solution, the irrigation application eventually creates the need for more desalination in areas that once had fresh groundwater resources.

Evaporation ponds are often considered the basic, conventional concentrate management option and were originally relatively inexpensive to build, given adequate space. However, today, regulations have been put in place to protect the local soil and groundwater from the salts and other potential chemicals that can leach into the ground from evaporation ponds (Nicot et al., 2007). Evaporation ponds have been primarily used in the Middle East and Australia, with some use in the U.S. (Texas) (Glater and Cohen, 2003). Research in the Middle East (Oman, United Arab Emirates, Israel) has shown the need for pond leakage monitoring and for enhanced evaporation strategies (Ahmed et al., 2001; Gilron et al., 2003; Glater and Cohen, 2003). In the U.S., the climate and land available in Texas has allowed this type of disposal to be a viable option. In comparison to Texas, other states considering brackish water RO in the U.S. (California,

Florida, East Coast states) do not have the climate and/or land necessary to build evaporations ponds. Both irrigation and evaporation ponds are disposal methods typically used for smaller size RO plants (<400 m³/day) (Mickley, 2004).

Deep well injection, where the concentrate is injected several hundred to several thousand meters into the ground below the fresh water aquifers, is another concentrate disposal option. This disposal option is practiced worldwide for all types of wastewater (Glater and Cohen, 2003; Mickley, 2001; Saripalli et al., 2000) and is often the most economical solution for inland plants. The disposal method has been used successfully in Florida for concentrate disposal. A plant recently started up (2007) in El Paso, Texas has chosen deep well injection, but the method remains controversial. Complications include appropriate site selection, concentrate conditioning with chemicals, corrosion and leakage from the well, possible damaging seismic activity, and unknown well lifetime, as well as sparingly soluble salt precipitation (Glater and Cohen, 2003). The process has not been in use long enough to know if the injected salt will eventually leach into the fresh water aquifers above. However, if designed appropriately and used for a long period of time, deep well injection could be the least costly disposal option for large inland desalination plants (Mickley, 2001).

For a brackish drinking water RO plant, all of the disposal options represent a loss of water and additional plant costs. In addition to all of the individual negative aspects of each disposal option described, each is costly in terms of water recovery; all of the water in the concentrate is lost during the disposal method. Whether by evaporation, wastewater discharge, alternate use, or injection, the water in the concentrate is not recovered as drinking water. In addition, most of the conventional disposal methods, except for deep well injection, are not applicable to large inland brackish water RO desalination plants in design today (Glater and Cohen, 2003).

The ultimate achievement in concentrate disposal and RO recovery is to operate a system with zero liquid discharge, or ZLD, where the recovery would approach 100%. In ZLD, most of the water in the concentrate is recovered as product by completely separating the salt from the water. ZLD systems include thermal evaporators, crystallizers, brine concentrators, and spray dryers (Mickley, 2001, 2004). A ZLD system combined with a high recovery brackish water RO system can produce permeate with as low as 10 mg/L TDS (Mickley, 2001). ZLD systems can be used in any geographical location and often are easily accepted by the local community due to positive environmental effects and minimal waste production (Mickley, 2001). While these systems are technologically available, the capital cost of such a system is often higher than the cost of the desalination plant alone (Mickley, 2004). In addition, the energy required to achieve near 100% recovery in a ZLD system is high and often not financially possible, except for very small RO systems. Somewhat lower costs may be achieved by combining a brine concentrator system with an evaporation pond, but most often this concentrate disposal method is the most expensive option (Mickley, 2004). Today, efforts to reduce the cost of ZLD technology continue (Bond and Veerapaneni, 2007; Sandia, 2003) bench- and pilot-scale tests have shown that ZLD processes can be applied to

a variety of water compositions and community needs (Bond and Veerapaneni, 2007).

A non-thermal process that approaches ZLD is electrodialysis (ED) or electrodialysis reversal (EDR). ED and EDR use ion exchange membranes and electric current to separate ions from water and to create a permeate and a concentrate. ED and EDR have lower salt rejections than RO membranes (up to 60% TDS reduction per stage) but can achieve recoveries of up to 94% in one stage and 97% with multiple stages (Reahl, 2004). EDR is the primary technology currently used; the electric field within the membranes is alternated, reducing membrane scaling. While EDR can operate at higher recoveries than RO, has longer membrane lifetimes, and requires less pre- and post-treatment, the higher capital and energy costs have limited the expansion of EDR. RO remains the primary choice for membrane desalination, while EDR is primarily used for hardness removal of low salinity brackish waters, specific industrial applications, and as a hybrid process with RO.

12. Alternative energy sources

12.1. Seawater RO

The coupling of alternative (renewable) energy sources with RO desalination plants has had increased interest and development. The plants in operation are small-scale (<10 m³/day) plants and represent approximately 0.02% of the total world desalination capacity (Mathioulakis et al., 2007). These plants are largely demonstration or research plants and often operate non-continuously; in addition, renewable energy sources are still more expensive than traditional resources (Helal et al., 2008; Lamei et al., 2008). Therefore, the unit cost operation for RO coupled with renewable energy is higher than for typical RO plants. Communities that would typically benefit from coupled renewable energy–RO systems are located in rural areas, where financial resources and system maintenance personnel are limited. Factors including capital cost, sustainable technology, technical operation, social acceptance, and energy resource availability, have contributed to the slow growth of the renewable energy–RO market (Mathioulakis et al., 2007).

The three main renewable energy sources available are solar (photovoltaic and thermal), wind and geothermal energy. The thermal energy sources are most often used with distillation desalination, while wind and photovoltaic solar energy are commonly paired with RO desalination. The combined choices of energy and process take advantage of matching the type of energy with the type of process (thermal versus mechanical). These RO systems can use seawater or brackish water as the feed source and are typically small to medium plants. Overall, the energy source most often used has been solar energy (70% of market), and RO has the majority (62%) of the renewable energy desalination market (Mathioulakis et al., 2007).

Solar-powered desalination is possibly the most promising alternative energy choice, and both distillation and membrane plants have been designed and operated. In particular, countries already advanced in conventional RO desalination, such as Spain, Italy, and Saudi Arabia, have

successfully implemented solar photovoltaic energy and seawater RO (García-Rodríguez, 2003; Herold et al., 1998; Tzen et al., 1998). In solar photovoltaic cells, the solar energy is directly converted into electricity, providing a direct energy source for RO operation. García-Rodríguez (2003) reported a wide range of permeate production capacities for solar photovoltaic–seawater RO plants, from 0.5 to 120 m³/day.

Spain, the desalination leader of Europe (Graber, 2006), has also paired renewable energy, in the form of a wind farm, with a seawater RO plant in Gran Canaria, Canary Islands (Rybar et al., 2005). Four wind generators provide the main energy source for the 5000 m³/day RO permeate production, along with pressure exchangers and motors with variable frequency that act as an energy recovery system. The power produced by the wind generators is variable throughout the year; at times, excess power is sold to the conventional power network in place, and sometimes, the RO plant consumes supplementary power from the network grid. In 2004, the plant obtained 57% of its power from the wind generators, and sold 95% of the power produced by the wind generators to the power network.

12.2. Brackish Water RO

Renewable energy sources have also been used in conjunction with brackish water RO desalination. In Australia, small-scale (0.4–1 m³/day) RO desalination systems powered by solar energy have recently been tested (Masson et al., 2005; Richards and Schaefer, 2002; Werner and Schaefer, 2007). These systems were used in remote areas of central Australia, where communities are small and fresh water is limited. While large RO plants fill an important role for large coastal cities, many other communities suffer from water needs; small RO plants with simple designs and renewable energy sources can successfully provide water to rural communities (Werner and Schaefer, 2007). The development of small RO systems in rural areas has been limited due to the high capital cost investment required (Ayoub and Alward, 1996), but the use of renewable energy could enable more communities to take advantage of RO technology. Perhaps most importantly, building and operating an RO plant in a rural community requires not only technical evaluation of the site and plant processes, but a study of the social aspects surrounding the eventual water use. Werner and Schaefer (2007) tested a solar energy–RO system in six sites in central Australia and found significant differences in water use, water resource quality, and human resources for system maintenance among the sites. Other countries, with arid, sunny climates, and rural communities that have limited access to electrical power grids or a central water distribution network, have also investigated similar renewable energy–RO systems for both seawater and brackish water sources (Bouguecha et al., 2005; García-Rodríguez, 2003; Herold et al., 1998; Tzen et al., 1998; Weiner et al., 2001); brackish water systems using solar photovoltaic energy have a range of production from 0.1 to 60 m³/day (García-Rodríguez, 2003).

13. Costs

Karagiannis and Soldatos (2008) conducted a review of water desalination cost literature and found that the type of feed

water (seawater or brackish water), as well as the plant size and the energy source, play major roles in the cost of desalinated water (\$/m³). The investment cost per unit of production capacity of seawater RO plants is higher than that of brackish water RO plants; seawater RO capital costs tend to fall between \$600/(m³/day) and \$800/(m³/day) (Reddy and Ghaffour, 2007; Sauvet-Goichon, 2007), while brackish water RO capital costs range from \$240/(m³/day) to \$400/(m³/day) (Vince et al., 2008; Yun et al., 2006). For both seawater and brackish water, small RO plants (<5000 m³/day) have higher unit water costs (\$/m³) than medium (5000–60,000 m³/day) or large (>60,000 m³/day) plants (Karagiannis and Soldatos, 2008).

A cost comparison of fixed and operation and maintenance costs for a large brackish water RO plant and a large seawater RO plant is shown in Table 6 (Reddy and Ghaffour, 2007; Yun et al., 2006). The brackish water RO plant cost data is based on a cost model evaluation of a theoretical plant; a smaller pilot study (1100 m³/day permeate) was performed on the feed water. The hypothetical plant would be located in southern California and operated by the Metropolitan Water District of Southern California. The plant would use a blend of Colorado River water (702 mg/L TDS) and California state project water (394 mg/L TDS) as the feed water (~500 mg/L TDS) and produce 700,300 m³/day permeate (Yun et al., 2006). The goal of this hypothetical plant would be to provide low-TDS permeate to blend with the permeate from the three existing conventional water treatment plants operated by Metropolitan. The existing plants use Colorado River water as the feed source and produce a combined 1.97 million m³/day permeate.

The seawater RO plant is the Ashkelon plant in southern Israel. This plant uses open seawater intake from the Mediterranean Sea (40,700 mg/L TDS) and produces 330,000 m³/day permeate (Sauvet-Goichon, 2007). The energy costs can be compared since both analyses were completed within a similar time period; today, the energy costs are likely to have risen above those calculated several years ago. The capital costs of the seawater plant are five times greater than the brackish water plant; this difference in capital costs is due in part to a more extensive pretreatment system to treat the surface water feed and larger pumping and piping needed to

Table 6 – Comparison of plant costs for a brackish water RO plant and a seawater RO plant (Reddy and Ghaffour, 2007; Yun et al., 2006).

Plant costs	Metropolitan plant (\$/m ³)	Ashkelon plant (\$/m ³)
Water type	Brackish water	Seawater
Fixed costs (capital costs)	0.057	0.311
Energy	0.029	0.134
Labor	0.007	Included in Miscellaneous
Chemicals	0.016	0.021
Membrane replacement	0.010	0.028
Miscellaneous	0.077	0.031
Total	0.134	0.525

move the seawater RO concentrate (because of the lower RO recovery). Energy costs for the seawater RO plant are nearly five times greater than those of the brackish water RO plant, due to higher pressures and lower recovery. The chemical costs are similar, while the membrane replacement costs are greater for the seawater RO system. Seawater RO membranes typically have shorter lifetimes and must be replaced more often due to fouling.

Hafez and El-Manharawy and Yun et al., give more detailed analyses of specific plant capital costs for a set of small seawater RO plants and for a large brackish water RO plant, respectively. The cost data include site-specific feed intake, pretreatment, post-treatment, site development and concentrate treatment costs (Hafez and El-Manharawy, 2002; Yun et al., 2006). A more detailed explanation of the factors affecting seawater RO and brackish water RO costs is presented below.

13.1. Seawater RO

The key to widespread interest and implementation of seawater RO plants has been a significant reduction in capital and operation/maintenance costs over the past 30 years. The unit cost of water from RO production has decreased from close to \$5.00/m³ in the late 1970s to less than \$1.00/m³ in 2004 (Zhou and Tol, 2004). Plants built in the 1990s produced water at a cost of \$1.00/m³–\$2.00/m³ (Wilf and Bartels, 2004), while recent plants built in Israel (Ashkelon plant, production started in 2005) and the United Arab Emirates (Fujairah plant, production started in 2005) cost \$0.53/m³ (in 2005) (Sauvet-Goichon, 2007) and \$0.87/m³ (Dawoud, 2005), respectively. The production capacities and water prices of several seawater RO plants are presented in Table 7 (Reverse Osmosis Desalination Plants Reference List, 2005; Dawoud, 2005; Glueckstern et al.,

2001; Magara et al., 2000; Sauvet-Goichon, 2007; Tsiourtis, 2001; Wilf and Bartels, 2006; Wilf and Klinko, 2001).

Several factors have helped reduce RO energy consumption and costs, including improvements in membrane materials and technology (higher flux, higher salt rejection, lower hydrostatic pressure required, lower materials cost) and the use of pressure recovery devices (Reddy and Ghaffour, 2007; Zhou and Tol, 2004). RO has also become less expensive than thermal processes, which require 10 times the electricity (Service, 2006), as crude oil prices have risen from 20 US\$/bbl (per barrel) in 1997, to approximately 70 US\$/bbl in 2007 (Blank et al., 2007). More recently, oil prices have risen past 100 US\$/bbl (Lee, 2008; Martin, 2007; Pflimlin, 2007). While oil prices increase, many project bids continue to base predicted energy costs on a figure of 5 US\$/bbl (Blank et al., 2007). This incongruence in predicted versus actual energy costs allows thermal desalination plants to still be compared to RO desalination plants as economic options. However, RO desalination is more energy efficient, and the process will continue to improve with new membrane materials and modifications. While most RO plants still use fossil fuels as the electricity source, future RO plants may take advantage of renewable energy resources or nuclear energy to further reduce costs (Mathioulakis et al., 2007; Reddy and Ghaffour, 2007; Rybar et al., 2005; Werner and Schaefer, 2007).

Both the relative capital cost and the energy cost (per unit of plant production capacity) of a RO plant decrease as plant size increases (Almulla et al., 2005; Avlonitis et al., 2003). The energy required by a RO system is primarily used to power pumps; the larger the plant, the greater the required power for larger pumping systems. For seawater RO plants, the power costs can account for up to 50% of the total plant operating and maintenance costs (Younos, 2003). The second largest cost is typically fixed costs (approximately 37%), including capital investment amortization and insurance (Miller, 2003; Younos, 2003). Other costs include maintenance and parts (7%), membrane replacement (5%), labor (4%), and consumable chemicals (3%) (Miller, 2003).

For a specific size plant, the choice in target recovery can greatly affect the cost of the plant (Wilf and Klinko, 2001). Most plants are designed to have a target permeate or product (volumetric) flow (Nadav et al., 2005; Petry et al., 2007; Reverberi and Gorenflo, 2007; Reverter et al., 2001; Sauvet-Goichon, 2007). For a fixed permeate flow, as recovery increases, the required feed flow decreases because more of the feed flow passes through the membranes as permeate. Thus, for higher recovery, all associated equipment (i.e., piping, pumps, storage tanks, pretreatment equipment, chemical dosing systems, concentrate outfall) can be sized smaller, and costs decrease. However, the increasing osmotic pressure with increasing recovery eventually overtakes the benefit of smaller feed flow, and at seawater RO recoveries above 55–60%, total water cost begins to increase (Wilf and Klinko, 2001). This optimal recovery percent range increases for less saline feed water.

RO pretreatment contributes significantly to the overall equipment cost of a plant. Conventional pretreatment (multi-media filtration) has traditionally cost less than other recent pretreatment alternatives (membrane filtration by MF, UF, or NF). However, the capital investment required for membrane pretreatment systems has decreased. While membrane

Table 7 – Summary of production and water prices for several large seawater RO desalination plants.

Plant location	1st year of production	RO production capacity (m ³ /day)	Water price (\$/m ³)
Galder-Agaete, Spain	1989	3500	1.94
Jeddah, Saudi Arabia	1989	23,000	1.31
Ad Dur, Bahrain	1990	45,000	1.30
Lanzarote III and Agragua, Spain	1991	10,000 × 2	1.62/1.34
Santa Barbara, California, USA	1992	25,000	1.51
Dhkelia, Cyprus	1997	40,000	1.46
Mallorca and Marbella, Spain	1998	42,000/56,400	1.03/1.00
Eilat, Israel	1998	10,000	0.72
Larnarca, Cyprus	2001	56,000	0.83
Eilat, Israel	2003	10,000	0.81
Tampa Bay, Florida, USA	2003	94,600	0.55
Fujairah, United Arab Emirates	2005	170,500	0.87
Ashkelon, Israel	2005	320,000	0.53

capital costs remain slightly higher than multi-media filtration capital costs, membrane pretreatment can substantially lower RO operation and maintenance costs, particularly the cost of RO membrane replacement. However, membrane pretreatment still tends to be more expensive than conventional pretreatment due to the higher cost of membrane replacement relative to the operating and maintenance costs of media filtration. As more and more seawater RO plants are forced to use open water intake feed sources, membrane pretreatment is likely to become the pretreatment method of choice.

Some full-scale plants, particularly in the Middle East and Asia, have already started using membrane pretreatment (Bartels et al., 2006; Bu-Rashid and Czolkoss, 2007). These plants, including those in Addur, Bahrain; Kindasa, Saudi Arabia; Fukuoka, Japan; and Yu-Han, China, have all seen consistent, improved water quality as a result of using membrane pretreatment. Bartels et al. (2006) noted that the capital cost of a RO plant utilizing membrane pretreatment could be as much as 30% higher than if conventional pretreatment had been chosen. The authors state that reduced operational and chemical costs can reduce the overall difference to between 2 and 3% (Bartels et al., 2006). Pearce (2007) described potential areas of cost savings due to improvements in RO operations. Membrane pretreatment can decrease costs associated with the plant footprint, RO membrane replacement, pre-RO cartridge filters, and chemical costs, as well as allow increases in recovery and permeate flux (due to lower fouling rates). In the case study presented, data are based on an eastern Mediterranean Sea feed water, and estimates are given for cost decreases for each of the parameters mentioned above. RO permeate flux can increase by 25%, while the plant footprint and RO membrane replacement can decrease by 33%. Chemical costs can decrease by 45–65% and pre-RO cartridge filters are not necessary. The total water cost for both the conventional pretreatment-RO system and the membrane pretreatment-RO system is \$0.90/m³ (Pearce, 2007). While the Mediterranean Sea has relatively good quality feed water and requires only one stage of media filtration (for conventional pretreatment), other feed sources, such as those in the Middle East (Persian Gulf), are poorer in quality and require more extensive pretreatment (Bu-Rashid and Czolkoss, 2007; Pearce, 2007).

13.2. Brackish water RO

The key difference in cost distribution between seawater RO and brackish water RO plants is the electrical power required. Brackish water RO plants require a much lower hydrostatic pressure to produce permeate because of the lower salt content of the feed water. In addition, brackish water RO membranes often have a lower salt rejection and are more permeable than seawater RO membranes. The power cost of a typical brackish water RO represents only 11% of the total cost (44% for seawater RO), and the largest costs are fixed costs (capital amortization and insurance), at 54% (Miller, 2003; Younos, 2003). Due to the decreased contribution of power in the overall cost distribution, other factors, such as maintenance (9%), membrane replacement (7%), labor (9%), and consumable chemicals (10%), increase (Miller, 2003).

The unit water price for brackish water RO ranges between \$0.10/m³ and \$1.00/m³ (Miller, 2003; Sethi, 2007); this price range is lower than that of seawater RO (\$0.53/m³–\$1.50/m³). Costs can increase due to pumping requirements (i.e., pumping groundwater from wells). Factors resulting from lower feed salinity, including the lower energy requirements noted above and less frequent membrane replacement, help decrease the water cost. As in seawater RO, pretreatment costs can represent a significant portion of the capital cost, particularly if membrane pretreatment is chosen. However, often conventional media filtration pretreatment is used because many brackish feed waters are groundwaters.

A key cost for brackish water RO systems is concentrate disposal. For brackish water RO plants, surface water outfall (typically located in the same body of water providing the feed water) is the least expensive disposal option. While still considered an important cost to seawater RO plants (Ravizky and Nadav, 2007), due to lower recoveries, surface water disposal is relatively inexpensive for brackish water systems. Apart from surface water disposal, combined sewer disposal is often the next choice as a relatively low-cost disposal option. This option is often not available, however, and plants must choose from more expensive options, depending on local regulations and available land. A summary of costs for concentrate disposal and the RO membrane system is shown in Table 8 (Graves and Choffel, 2004; Koyuncu et al., 2001; Mickley, 2001, 2004; Miller, 2003; Sethi, 2007; Thomas, 2006); the cost of producing desalinated water becomes more costly as the concentrate disposal costs increase. Surface water disposal is by far the least expensive option, although piping and pumping costs can significantly increase when the plant is not located on the coast. Evaporation ponds and brine concentrators are the most expensive options due to stringent groundwater regulations and energy requirements, respectively.

Table 8 – Cost comparison of membrane unit and concentrate disposal options.

Cost	Cost (\$/m ³)	Critical factors
Membrane replacement	0.008–0.05	Scaling, fouling
Chemicals	0.008–0.05	Cleaning frequency
<i>Concentrate disposal^a</i>		
Surface Water	0.03–0.30	Piping, pumping, and outfall construction
Evaporation pond	1.18–10.04	Pond size and depth, salt concentration, evaporation rate, disposal rate, pond liner cost
Deep well injection	0.33–2.64	Tubing diameter and depth, injection rate, chemical costs
Sewer	0.30–0.66	Disposal rate, salinity, sewer capacity, fees
Brine concentrator (ZLD)	0.66–26.41	Disposal rate, energy costs, salinity
a Costs for concentrate disposal options include capital and operations and maintenance (O&M) costs.		

14. Technological challenges and the future of RO

An emerging application of RO membranes is in wastewater treatment and trace organic contaminant removal. A host of new organic contaminants have been identified (Richardson et al., 2007), and RO technology is a potential treatment candidate. Particularly for hydrophilic organic compounds, including many disinfection by-products and pharmaceutical compounds, traditional treatment processes (coagulation and flocculation) are not effective at removal. However, RO membranes may remove these compounds through both charge repulsion and size exclusion phenomena. Using RO membranes in wastewater treatment presents unique process challenges; calcium phosphate precipitation can occur, and wastewaters tend to have much higher organic carbon content than drinking water feed streams. Membrane fouling and pretreatment design will be primary concerns as RO systems are developed for wastewater treatment.

The development of energy recovery devices and hybrid desalination/power plants has allowed significant advances in energy recovery. In addition, new RO membrane module design, including larger diameter spiral wound modules (Yun et al., 2006) and high-flux membranes, has provided cost and energy efficiency improvements to the typical RO system design. Further research and technology development in energy recovery and system design will allow additional gains in energy recovery and cost reduction.

A key limitation to commercial polyamide RO membranes and treatment system design is membrane degradation through contact with chlorine, one of the common disinfectants used in water and wastewater treatment. Recent research in novel membrane materials and polymer chemistry (Park et al., 2008) has resulted in the development of sulfonated polysulfone composite membranes that are highly resistant to chlorine attack. Commercial development of chlorine-resistant membranes would eliminate the need for dechlorination of the RO feed and rechlorination after the membrane system, reducing the overall cost of RO.

The need for inland brackish water RO will continue to increase in the future, and the primary limitations to further application of RO inland are the cost and technical feasibility of concentrate disposal. Research on novel concentrate treatment options is ongoing (Rahardianto et al., 2007), and pilot plant demonstrations have shown that significant increases in RO recovery are possible. Optimization of anti-scalant dosing, chemical addition, and pH control is necessary to improve the cost of concentrate treatment. Full-scale use of concentrate treatment is just beginning and will be necessary to allow economic use of inland brackish water resources.

Increasingly stringent water quality standards will cause further optimization and development in RO membrane technology. In particular, the standard for boron has been lowered, and seawater RO plants may need more than one RO pass to achieve the required water quality. Membrane manufacturers are developing new RO membranes with higher boron rejections; future technology may focus on other regulated and emerging contaminants, including disinfection by-products, pharmaceuticals, and endocrine disrupting compounds.

The extensive development of coastal desalination plants that use surface water discharge as concentrate disposal has the potential to negatively affect the local receiving water and the larger surrounding sea. In particular, research and modeling on salinity variations in the Arabian Gulf (Altayaran and Madany, 1992; Purnama et al., 2005; Smith et al., 2007) show that an increase in coastal desalination installations is likely to increase the salinity in the Gulf and cause local variations in oxygen content and temperature. As the use of desalination continues to grow, the impact of desalination plants on local water bodies must be evaluated, and negative impacts must be minimized.

The use of membrane filtration in RO pretreatment will continue to be investigated; as membrane costs decrease, the use of membrane pretreatment will become a more viable alternative to conventional pretreatment. Particularly for surface water sources, membrane pretreatment is a constant barrier to particulate and colloidal RO membrane fouling and can greatly improve RO feed water quality. Research on SDI values and membrane fouling has shown that SDI is not always an appropriate indicator of RO fouling. An improved method for prediction of fouling potential is needed.

15. Conclusions

The field of RO membrane desalination has rapidly grown over the past 40 years to become the primary choice for new plant installations. Membrane technology has improved, allowing significant increases in product production and cost savings. While the basic operating principles remain the same for all RO applications, individualized applications have developed, based on feed water quality. In particular, the two key types of feed water, seawater and brackish water, have distinguishing features that demand specific parameter adjustment and system design. Seawater RO recovery is primarily limited by osmotic pressure increase and organic material fouling; system design typically consists of chemical and filtration pretreatment and one RO stage. However, problematic components, such as boron, can require more complex RO stage design. Brackish water RO membrane systems typically consist of two RO stages in series; key issues include salt precipitation and concentrate management. While both seawater and brackish water RO have been sufficiently developed to be used in large-scale commercial plants, several significant challenges to the RO field remain. Further improvements in membrane technology, energy use, and concentrate treatment will allow a wider application of RO to inland and rural communities.

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