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Reverse Osmosis

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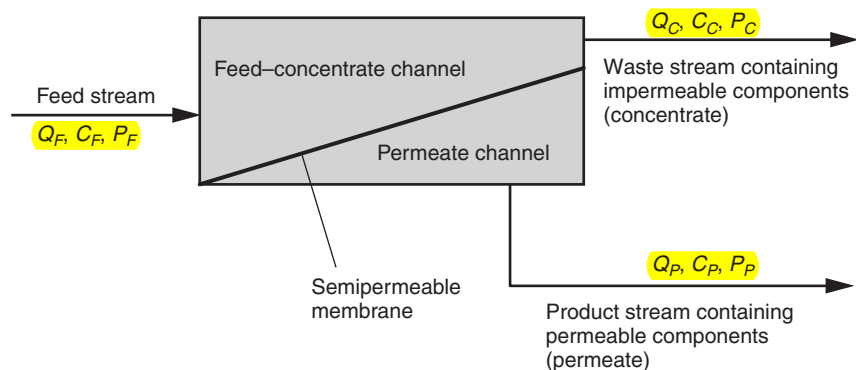
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Terminology for Reverse Osmosis

Term	Definition
Active layer	Layer of membrane that provides the separation capabilities.
Asymmetric structure	Membrane formed of a single material but with multiple layers that are structurally different and have different functions.
Array	Full unit of water production in a reverse osmosis system, which may include multiple stages.
Concentrate	Portion of feed water that has not passed through the membrane. Constituents removed from the permeate are concentrated in the concentrate. Also known as brine.
Concentration polarization	Accumulation of solutes near a membrane surface due to boundary layer effects and the rejection of solutes by the membrane as water passes through the membrane.
Dense membrane	Material that is permeable to certain constituents, such as water, even though it does not have pores.
Limiting salt	Salt that reaches its saturation concentration first as water is concentrated in a reverse osmosis system.
Membrane element	Smallest distinct unit of production capacity in a reverse osmosis system; several membrane elements are arranged in series in a pressure vessel.
Nanofiltration membrane	Reverse osmosis membrane product engineered for selective removal of divalent ions or natural organic matter while allowing passage of smaller monovalent ions.
Osmosis	Flow of solvent through a semipermeable membrane from a dilute solution into a concentrated one.

Term	Definition
Osmotic pressure	Pressure required to balance the difference in chemical potential between two solutions separated by a semipermeable membrane.
Permeate	Portion of feed water that has passed through the membrane. Solutes have been largely removed from this stream so that it is usable for potable purposes. Also known as product water.
Reverse osmosis	Physicochemical separation process in which water flows through a semipermeable membrane due to the application of an external pressure in excess of the osmotic pressure.
Semipermeable membrane	Material that is permeable to some components in a solution but not others; e.g., a material permeable to water but not permeable to salts.
Spiral wound element	Most common type of reverse osmosis membrane element, in which envelopes of membrane material are wrapped around a permeate tube and treated water flows spirally through the envelope to the tube.
Stage	Group of pressure vessels operated in parallel as a single component of water production.
Thin-film composite	Reverse osmosis membranes composed of two or more materials cast on top of one another, where one material is the active layer and other materials form the support layers.

Reverse osmosis (RO) is a membrane treatment process used to separate dissolved solutes from water. It includes any pressure-driven membrane that uses preferential diffusion for separation. A typical RO membrane is made of synthetic semipermeable material, which is defined as a material that is permeable to some components in the feed stream and impermeable to other components and has an overall thickness of less than 1 mm. Water is pumped at high pressure across the surface of the membrane, causing a portion of the water to pass through the membrane, as shown schematically on Fig. 17-1. Water passing through the membrane, called permeate, is relatively free of targeted dissolved solutes, while the remaining water, called concentrate (also commonly called retentate, reject water, or brine), exits at the far end of the pressure vessel. The delineation of membrane processes, applications for RO, a historical perspective, a process description, process fundamentals, and process design are presented in this chapter.

**Figure 17-1**

Schematic of separation process through reverse osmosis membrane.

17-1 Classification of Membrane Processes

Membrane processes were introduced in Chap. 12, where it was noted that the membranes used in municipal water treatment include microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), and reverse osmosis (RO) membranes. From a physicochemical perspective, these four types of membranes are used in two distinct processes in water treatment (1) membrane filtration and (2) reverse osmosis. They are differentiated by the types of materials rejected, characteristic pore dimensions, and operating pressures. Membrane filtration is used primarily for the removal of particulate matter, whereas RO accomplishes a variety of treatment objectives involving the separation of dissolved solutes from water.

Membrane filtration is covered in Chap. 12, in which a hierarchy of membranes used in water treatment is described (Fig. 12-2), and additional details are provided on the delineation between membrane filtration and RO (Sec. 12-1) including a table of significant differences between these processes (Table 12-1). Common membrane nomenclature is included in Chap. 12 as well as here.

Nanofiltration membranes were designed by FilmTec Corporation around 1983 to remove divalent anions from seawater for applications in the oil industry. The word *nanofiltration* was coined because the separation cutoff size was about 1 nm, and the membranes were designed for removal of specific ionic species, whereas other RO membranes of that era were indiscriminate with respect to the ionic species removed. The ability of NF membranes to simultaneously remove divalent cations (hardness) and natural organic matter while achieving only low monovalent ion removal made them ideal for certain water treatment applications. While NF membranes were a unique product in the 1980s, membrane manufacturers have since engineered a variety of RO membranes with different formulations, permeation capabilities, and rejection characteristics. These products provide a full range of different capabilities, and some new RO

membranes have characteristics similar to the original NF membranes. A variety of names have been applied to these new products, including “loose” RO membranes, softening membranes, and low-pressure RO membranes. Manufacturers will continue to develop new RO membranes to achieve specific goals, and NF membranes are just one in a succession of many innovative developments in the field of RO.

17-2 Applications for Reverse Osmosis

Uses for RO in water treatment as well as alternative processes are listed in **Table 17-1**. These objectives encompass the desalination of ocean or brackish water, advanced treatment for water reuse, softening, natural organic matter (NOM) removal for controlling disinfection by-product (DBP) formation, and specific contaminant removal.

The scarcity of freshwater sources may mean a strong future for the use of RO for desalination of ocean water or seawater. About 97.5 percent of the earth’s water is in the oceans, and about 75 percent of the world’s population live in coastal areas (Bindra and Abosh, 2001). The salinity of the ocean ranges from about 34,000 to 38,000 mg/L as total dissolved solids (TDS) (Stumm and Morgan 1996), nearly two orders of magnitude higher than that of potable water [the World Health Organization’s (WHO’s

Desalination of Ocean Water or Seawater

Table 17-1

Reverse osmosis objectives and alternative processes

Process Objective	Membrane Process Name	Alternative Processes
Ocean or seawater desalination	High-pressure RO, seawater RO	Multistage flash (MSF) distillation, multieffect distillation (MED), vapor compression distillation (VCD)
Brackish water desalination	RO, low-pressure RO, NF	Multistage flash distillation, ^a multieffect distillation, ^a vapor compression, ^a electrodialysis, electrodialysis reversal
Softening	Membrane softening, NF	Lime softening, ion exchange
NOM removal for DBP control	NF	Enhanced coagulation/softening, GAC
Specific contaminant removal ^b	RO	Ion exchange, activated alumina, coagulation, lime softening, electrodialysis, electrodialysis reversal
Water reuse	RO	Advanced oxidation
High-purity process water	RO	Ion exchange, distillation

^aMSF, MED, and VCD are rarely competitive economically for brackish water desalination.

^bApplicability of alternative processes depends on the specific contaminants to be removed and their concentration.

Table 17-2

Typical concentration of important solutes in seawater

Salt	Concentration, mg/L
Cations	
Sodium, Na ⁺	10,800
Magnesium, Mg ²⁺	1,290
Calcium, Ca ²⁺	412
Potassium, K ⁺	399
Strontium, Sr ²⁺	7.9
Barium, Ba ²⁺	0.02
Anions	
Chloride, Cl ⁻	19,400
Sulfate, SO ₄ ²⁻	2,700
Total carbonate, CO ₃ ²⁻	142
Bromide, Br ⁻	67
Fluoride, F ⁻	1.3
Phosphate, HPO ₄ ²⁻	0.5
Total	35,200

Source: Stumm and Morgan (1996).

guidance level for TDS is 1000 mg/L and the United States has a secondary standard of 500 mg/L)]. The concentration of important ions in seawater is shown in Table 17-2. Seawater also contains several important neutral species, including 3 mg/L of silicon (present as H₄SiO₄) and 4.6 mg/L of boron (present as H₃BO₃). Boron is a concern because neutral species are poorly removed by conventional RO membranes, as will be presented later, and California has a notification limit of 1 mg/L for boron in drinking water.

Desalination costs are dropping, and the process is becoming more competitive with other treatment options in areas where freshwater is scarce, although desalination of ocean water is an energy-intensive process.

The Middle East is currently the most prominent market for desalination of ocean water. Virtually 100 percent of the drinking water in Kuwait and Qatar and 40 to 60 percent of the drinking water in Bahrain, Saudi Arabia, and Malta is produced by desalination (Bremere et al., 2001). Thermal processes such as multistage flash (MSF) distillation and multieffect distillation (MED) are common in the Middle East, which has vast energy resources but little freshwater. Worldwide, 43 percent of desalination is done with thermal processes and 56 percent is done with membrane processes (NRC, 2008).

Interest in the oceans as a source water is growing in other areas, including coastal areas of the United States. Tampa, Florida, commissioned

a 95,000-m³/d (25-mgd) RO plant in 2003, and a number of communities in California are considering the Pacific Ocean as a source of municipal water.

Interest in desalination of brackish groundwater has increased in areas short on freshwater, such as the southwest region of the United States. Communities in that area are rapidly growing beyond the availability of local freshwater supplies. Brackish groundwater with low to moderate salinity (1000 to 5000 mg/L TDS) are relatively common, and use of these resources has become reasonable as desalination costs have dropped and costs to obtain additional freshwater resources has increased. The difference in feed water quality between brackish water and seawater can lead to differences in design and operation, including differences in pretreatment, feed pressure, configuration of stages, water recovery, fouling prevention, and waste disposal (Greenlee et al., 2009). Since energy consumption is directly related to feed water TDS, brackish water desalination is not nearly as energy intensive as seawater desalination. However, disposal of the concentrate is a significant challenge.

Desalination of Brackish Groundwater

Along with brackish groundwater as an alternative source of water, many communities in water-scarce areas are considering the increased use of recycled treated wastewater. Water reuse for nonpotable uses such as irrigation of municipal greenspaces (parks, golf courses, road medians, etc.) and industrial process water is practiced in some areas, but treating wastewater to sufficient quality for potable reuse would increase flexibility for using the resource and eliminate the need for community dual-pipe systems. A concern in potable reuse applications, however, is the presence of pharmaceuticals, personal care products, endocrine disrupting compounds, unregulated contaminants, and other contaminants of emerging concern. RO's ability to remove virtually all contaminants in water, including many synthetic organic chemicals, has increased the interest in incorporating RO into wastewater treatment process trains as an advanced treatment process.

Water Reuse

Nanofiltration or softening membranes are capable of removing 80 to 95 percent of divalent ions such as calcium and magnesium with low removal of low-molecular-weight (MW) monovalent ions such as sodium and chloride. By allowing passage of sodium and chloride, the osmotic pressure differential is minimized. Nanofiltration membranes can soften water without the voluminous sludge production of lime softening, although concentrate disposal can be a significant regulatory obstacle. Nanofiltration membranes that effectively remove hardness are also effective at removing NOM, making them an excellent treatment option for color removal and DBP formation control because removing NOM and color from water before disinfection with free chlorine typically reduces the formation of DBPs. Nanofiltration membranes have widespread use in Florida,

Softening and NOM Removal

where the groundwater is either brackish or very hard, highly colored freshwater.

Specific Contaminant Removal

An additional use for RO is specific contaminant removal. The EPA has designated RO as a best available technology (BAT) for removal of numerous inorganic contaminants, including antimony, arsenic, barium, fluoride, nitrate, nitrite, and selenium, and radionuclides, including beta-particle and photon emitters, alpha emitters, and radium-226. Reverse osmosis has also been demonstrated to be effective for removing larger MW synthetic organics such as pesticides (Baier et al., 1987). Use of RO for specific contaminants, however, is less common because alternative technologies are frequently more cost effective and the disposal of the concentrate stream may present challenges.

17-3 History of Reverse Osmosis in Water Treatment

The process of **osmosis** through semipermeable membranes was first observed in **1748** by Jean Antoine Nollet (Laidler and Meiser, 1999). The feasibility of desalinating seawater with semipermeable membranes was first seriously investigated in 1949 at the University of California at Los Angeles (UCLA) and in about 1955 at the University of Florida, with funding provided by the newly formed U.S. Department of Interior Office of Saline Water (Glater, 1998). Researchers at both UCLA and the University of Florida successfully produced freshwater from seawater in the mid-1950s, but the flux was too low to be commercially viable. Research focused on reducing the membrane thickness, and in 1959, Loeb and Sourirajan of UCLA succeeded in producing the first asymmetric RO membrane (Lonsdale, 1982). *Asymmetric* membranes are formed from a single material that develops into active and support layers during the casting process (in other words, the membranes are chemically homogeneous but physically heterogeneous). Due to the thinness of the *active layer*, which provides separation capabilities, the asymmetric membrane was a major breakthrough. That advancement, along with the development of the spiral wound element to increase packing density and thin-film composite membranes, led to the commercial viability of membrane desalination.

In June of 1965, the first commercial membrane desalination plant began providing potable water to the City of Coalinga, California. The plant, with combined experimental and production capabilities, produced $19 \text{ m}^3/\text{d}$ (0.005 mgd) of potable water from 2500 mg/L TDS feed water by operating at 41 bar (600 psi) pressure, $34 \text{ L}/\text{m}^2 \cdot \text{h}$ ($20 \text{ gal}/\text{ft}^2 \cdot \text{d}$) flux, and 50 percent recovery (Stevens and Loeb, 1967). Other plants soon followed. The construction of Water Factory 21 in California helped the industry standardize on specific configurations, such as the 8-in. spiral-wound element. In the mid-1970s, RO applications were extended from

desalting to the softening applications mentioned earlier. The first membrane softening plant was built in Pelican Bay, Florida, in 1977 (AWWA, 2007). The use of membranes to remove NOM paralleled the development of membrane softening (Taylor et al., 1987) because many groundwater supplies in Florida are both hard and colored, and NOM and hardness can be removed simultaneously by membranes.

By the end of 2008, the total installed capacity of desalination plants was $42 \times 10^6 \text{ m}^3/\text{d}$ (11 billion gallons per day) worldwide. Over 1100 RO plants are operating in the United States with a total capacity of around $5.7 \times 10^6 \text{ m}^3/\text{day}$ (1500 mgd) (NRC, 2008), which represents about 3 percent of water withdrawn by public water systems. Reverse osmosis plants have been built in every state in the United States.

The future of RO is promising. Growth in the world population, the urbanization of coastal and arid areas, the scarcity of freshwater supplies, the increasing contamination of freshwater supplies, greater reliance on oceans and poorer quality supplies (brackish groundwater, treated wastewater), and improvements in membrane technology suggest continued rapid growth of reverse osmosis installations. The installation of desalination facilities is expected to double between 2005 and 2015 (Wang et al., 2010).

17-4 Reverse Osmosis Process Description

Reverse osmosis relies on differences between the physical and chemical properties of the solutes and water to achieve separation. A high-pressure feed stream is directed across the surface of a semipermeable material, and due to a pressure differential between the feed and permeate sides of the membrane, a portion of the feed stream passes through the membrane. As water passes through the membrane, solutes are rejected and the feed stream becomes more concentrated. The permeate stream exits at nearly atmospheric pressure, while the concentrate remains at nearly the feed pressure. Reverse osmosis is a continuous separation process; that is, there is no periodic backwash cycle.

A typical RO facility is shown on Fig. 17-2. The smallest unit of production capacity in a membrane plant is called a *membrane element*. The membrane elements are enclosed in pressure vessels mounted on skids, which have piping connections for feed, permeate, and concentrate streams. A group of pressure vessels operated in parallel is called a *stage*. The concentrate from one stage can be fed to a subsequent stage to increase water recovery (a multistage system, sometimes called a brine-staged system) or the permeate from one stage can be fed to a second stage to increase solute removal (a two-pass system, also sometimes called a permeate-staged system). In multistaged systems, the number of pressure vessels decreases in each succeeding stage to maintain sufficient velocity in the feed channel



Figure 17-2
Typical reverse osmosis facility.

as permeate is extracted from the feed water stream. A unit of production capacity, which may contain one or more stages, is called an array. Schematics of various arrays are shown on Fig. 17-3. The ratio of permeate to feed water flow (recovery) ranges from about 50 percent for seawater RO systems to about 90 percent for low-pressure RO systems. Several factors limit recovery, most notably osmotic pressure, concentration polarization, and the solubility of sparingly soluble salts.

Pretreatment and Posttreatment

A schematic of an RO system with typical pretreatment and posttreatment processes is shown on Fig. 17-4 and described below.

PRETREATMENT

Feed water pretreatment is required in virtually all RO systems. When sparingly soluble salts are present, one purpose of pretreatment is to

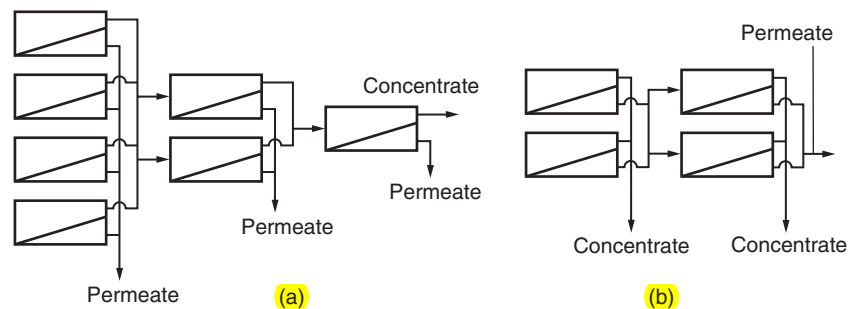


Figure 17-3
Array configurations of reverse osmosis facilities:
(a) $4 \times 2 \times 1$ concentrate-staged array,
(b) two-pass system.

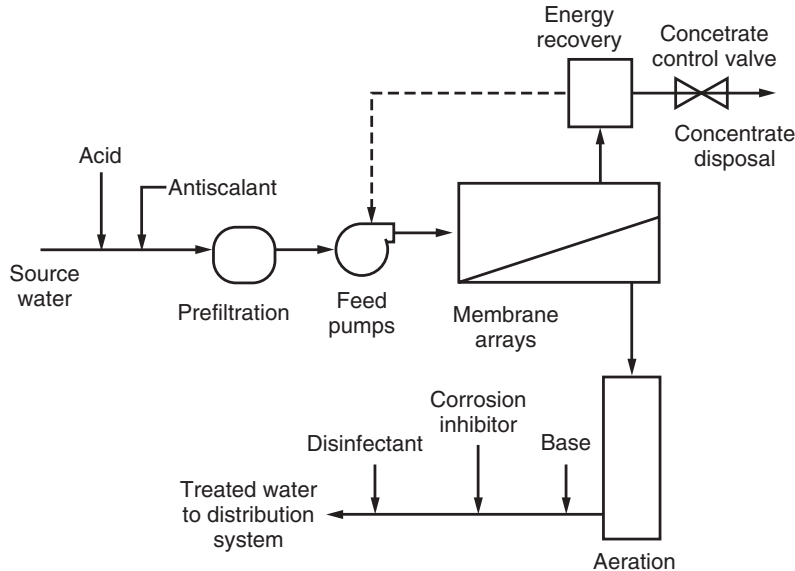


Figure 17-4
Schematic of typical reverse osmosis facility.

prevent scaling. Solutes are concentrated as water is removed from the feed stream, and the resulting concentration can be higher than the solubility product of various salts. Without pretreatment, these salts can precipitate onto the membrane surface and irreversibly damage the membrane. **Scale control consists of pH adjustment and/or antiscalant addition.** Adjusting the pH changes the solubility of precipitates and antiscalants interfere with crystal formation or slow the rate of precipitate formation.

The second pretreatment process is filtration to remove particles. Without a backwash cycle, particles can clog the feed channels or accumulate on the membrane surface unless the concentration is low. As a minimum, cartridge filtration with a 5- μm strainer opening is used, although granular filtration or membrane filtration pretreatment is often necessary for surface water sources. Disinfection is another typical pretreatment step and is used to prevent biofouling. Some membrane materials are incompatible with disinfectants, so the disinfectant can only be applied in specific situations and must be matched to the specific membrane type.

After pretreatment, the feed water is pressurized with feed pumps. The feed water pressure ranges from 5 to 10 bar (73 to 145 psi) for NF membranes, from 10 to 30 bar (145 to 430 psi) for low-pressure and brackish water RO, and from 55 to 85 bar (800 to 1200 psi) for seawater RO.

POSTTREATMENT

Permeate typically requires posttreatment, which consists of removal of dissolved gases and alkalinity and pH adjustment. Membranes do not efficiently remove small, uncharged molecules, in particular dissolved gases.

If hydrogen sulfide is present in the source groundwater, it must be stripped prior to distribution to consumers. If sulfides are removed in the stripping process, provisions must be made to scrub the sulfides from the stripping tower off-gas to prevent odor and corrosion problems. The stripping of carbon dioxide raises pH and reduces the amount of base needed to stabilize the water. Permeate is typically low in hardness and alkalinity and frequently has been adjusted to an acidic pH value to control scaling. Consequently, the permeate is corrosive to downstream equipment and piping. Alkalinity and pH adjustments are accomplished with various bases, and corrosion inhibitors are used to control corrosion.

Concentrate Stream

The concentrate stream is under high pressure when it exits the final membrane element. This pressure is dissipated through the concentrate control valve, which can be a significant waste of energy. Seawater RO systems utilize energy recovery equipment on the concentrate line, and some brackish water RO systems are starting to use energy recovery as well. Unlike cross-flow membrane filtration, the concentrate stream is not recycled to the head of the plant but is a waste stream that must be discarded. Concentrate disposal can be a significant issue in the design of RO facilities and the concentrate may require treatment before disposal. Methods for concentrate disposal are discussed in Chap. 21 and include ocean, brackish river, or estuary discharge; discharge to a municipal sewer; and deep-well injection. Other concentrate disposal options, including evaporation ponds, infiltration basins, and irrigation, are used by a small number of facilities.

Membrane Element Configuration

Reverse osmosis membrane elements are fabricated in either a spiral-wound configuration or a hollow-fine-fiber (HFF) configuration.

SPIRAL-WOUND MODULES

Spiral-wound modules are constructed of several elements in series. The basic construction of a spiral-wound element is shown on Fig. 17-5, and a photograph of typical elements is shown on Fig. 17-6. An envelope is formed by sealing two sheets of flat-sheet membrane material along three sides, with the active membrane layer facing out. A permeate carrier spacer material inside the envelope prevents the inside surfaces from touching each other and provides a flow path for the permeate inside the envelope. The open ends of several envelopes are attached to a perforated central tube known as a permeate collection tube. Feed-side mesh spacers are placed between the envelopes to provide a flow path and create turbulence in the feed water. By rolling the membrane envelopes around the permeate collection tube, the exterior spacer forms a spirally shaped feed channel. This channel, exposed to element feed water at one end and concentrate at the other end, is known as the feed-concentrate channel. Membrane feed water passes through this channel and is exposed to the membrane surface.

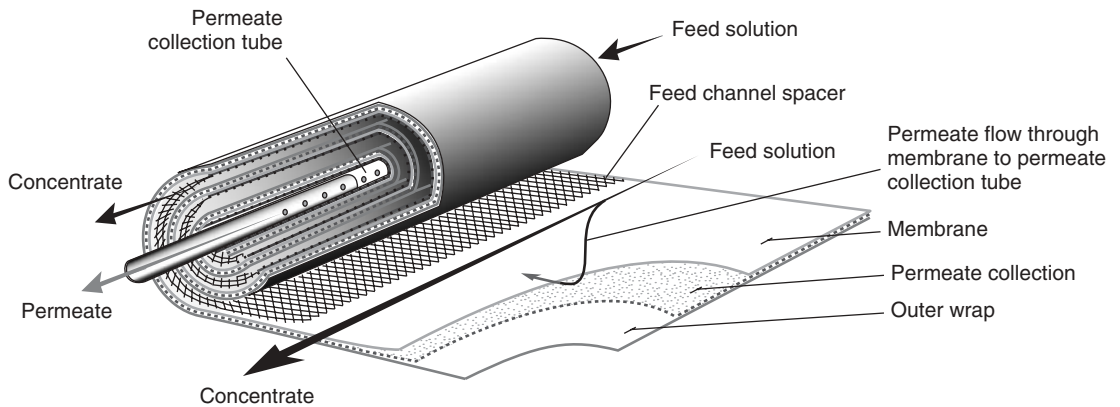


Figure 17-5
Construction of spiral-wound membrane element.



Figure 17-6
Photograph of spiral-wound membrane elements. (Courtesy GE Infrastructure Water Technologies.)

Spiral-wound elements are typically 1 m (40 in.) to 1.5 m (60 in.) long and 0.1 m (4 in.) to 0.46 m (18 in.) in diameter, although 0.2 m (8 in.) diameter elements are most common. Four to seven elements are arranged in series in a pressure vessel, with the permeate collection tubes of the spiral-wound elements coupled together.

During operation, pressurized feed water enters one side of the pressure vessel and encounters the first membrane element. As the water flows tangentially across the membrane surface, a portion of the water passes through the membrane surface and into the membrane envelope and flows spirally toward the permeate collection tube. The remaining feed water, now concentrated, flows to the next element in series, and the process is repeated until the concentrate exits the pressure vessel. Individual spiral-wound membrane elements have a permeate recovery of 5 to 15 percent per element. Head loss develops as feed water flows through the feed channels and spacers, which reduces the driving force for flow through the membrane surface. This feed-side head loss across a membrane element is low, typically less than 0.5 bar (7 psi) per element.

HOLLOW-FINE-FIBER MODULES

The HFF configuration is similar to the hollow fibers used in membrane filtration. Feed water passes over the outside of the fiber and is forced through the wall of the fiber, and the permeate is collected in the lumen (or inner annulus) of the fiber. The original manufacturer of HFF membranes was DuPont, which manufactured fiber with an outside diameter (OD) of 0.085 mm (about the thickness of human hair) and inside diameter of 0.042 mm, considerably thinner than the hollow fibers used in membrane filtration, which have an OD of 1 to 2 mm (about the thickness of pencil lead) (Lonsdale, 1982). The active surface of the membrane is on the outside surface of the fiber and is 0.1 to 1 μm thick. DuPont HFFs are still in widespread use but are no longer commercially available. The only current manufacturer of hollow-fiber RO membranes is Toyobo in Japan. In a typical HFF module, the feed enters one end of the module and the concentrated brine exits from the opposite end. The fibers are folded and suspended lengthwise in the module, with the open ends of a set of fibers exposed at each end of the module. The fiber bundles are wound helically around a center tube. A single module can contain several hundred thousand fibers and have surface area up to 10 times that of spiral wound elements. Product water recovery per element is 30 percent.

17-5 Reverse Osmosis Fundamentals

The fundamentals of RO include the membrane material properties, the phenomenon of osmotic pressure, the mechanisms for water and solute permeation, the equations used to predict water and solute flux, and the phenomenon of concentration polarization. These topics are addressed in this section.

An understanding of the mechanisms that control RO begins with an understanding of the membrane. Important properties include the physical structure, chemistry, and rejection capabilities of the membranes.

MEMBRANE STRUCTURE

The resistance to flow through a membrane is inversely proportional to thickness. To achieve any appreciable water flux, the active membrane layer must be extremely thin, which in RO and NF membranes ranges from about 0.1 to 2 μm . Material this thin lacks structural integrity, so these membranes are comprised of several layers, with a thin active layer providing separation capabilities and thicker, more porous layers providing structural integrity. Multilayer membranes are fabricated in two ways. As previously mentioned, asymmetric membranes are formed from a single material that develops into active and support layers during the casting process (in other words, the membranes are chemically homogeneous but physically heterogeneous). *Thin-film composite* membranes are composed of two or more materials cast on top of one another. An advantage of thin-film membranes is that separation and structural properties can be optimized independently using appropriate materials for each function. A cross section of an RO membrane is shown on Fig. 17-7.

The active layer of RO membranes must selectively allow water to pass through the material while rejecting dissolved solutes that may have

Membrane Structure, Material Chemistry, and Rejection Capabilities

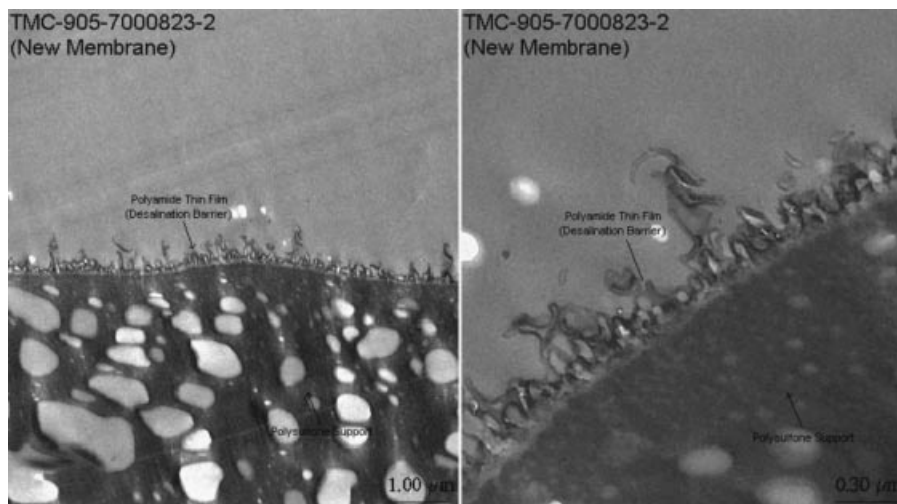


Figure 17-7
Microphotographs of asymmetric reverse osmosis membrane. (TEM images courtesy Bob Riley.)

dimensions similar to water molecules. Separation of small ions cannot be accomplished if they are convectively carried with liquid water. Thus, RO membranes are fabricated of a *dense* material, meaning a permeable but not porous material with no void spaces through which liquid water travels. Water and solutes dissolve into the solid membrane material, diffuse through the solid, and reliquefy on the permeate side of the membrane. The mechanics of permeation through a dense material will be discussed in detail later in this chapter. Low-pressure RO or NF membranes may have void spaces large enough for the convective flow of liquid water through the membrane.

MEMBRANE MATERIAL

Membrane performance is strongly affected by the physical and chemical properties of the material. The ideal membrane material is one that can produce a high flux without clogging or fouling and is physically durable, chemically stable, nonbiodegradable, chemically resistant, and inexpensive. Important characteristics of membrane materials, methods of determination, and effects on membrane performance were discussed in Chap. 12 and shown in Table 12-7. The materials most widely used in RO are cellulosic derivatives and polyamide derivatives.

Cellulose acetate membranes

The original RO membrane developed by Loeb and Sourirajan in 1960 was fabricated of cellulose acetate (CA), and RO membranes using this material are still commercially available. Membranes composed of CA are typically of asymmetric construction. Cellulose acetate is hydrophilic, which helps to maintain high flux values and to minimize fouling. The structural properties of CA are not ideal, however, and the material is not tolerant of temperatures above 30°C, tends to hydrolyze when the pH value is below 3 or above 8, is susceptible to biological degradation, and degrades with free-chlorine concentrations above 1 mg/L, depending on the concentration and duration of contact. In addition, membrane compaction due to the high operating pressure and asymmetric construction causes a reduction of flux over time.

Polyamide membranes

Polyamide (PA) membranes are chemically and physically more stable than CA membranes, generally immune to bacterial degradation, stable over a pH range of 3 to 11, and do not hydrolyze in water. Under similar pressure and temperature conditions, PA membranes can produce higher water flux and higher salt rejection than CA membranes. However, PA membranes are more hydrophobic and susceptible to fouling than CA membranes and are not tolerant of free chlorine in any concentration. Any residual oxidant such as chlorine in the feed will cause rapid deterioration of the

membrane. For most applications, dechlorination is required if the feed water is chlorinated and can be done with sodium bisulfite, sulfur dioxide, or activated carbon. Sensors and instrumentation must be provided to monitor the feed water for oxidants that may damage the material and shut down the system if any are detected. Some PA membranes have a rougher surface than CA membranes, which can increase susceptibility to biological and particulate fouling. Polyamide membranes are typically of thin-film construction. The PAs are used for the active layer, and durable materials such as polyethersulfone are used for the support material. The support layer is essentially a standard UF membrane and provides little resistance to flow.

REJECTION CAPABILITIES

The rejection capabilities of RO and NF membranes are designated with either a percent salt rejection or a molecular weight cutoff (MWCO) value. Salt rejection is typically used for RO membranes:

$$\text{Rej} = 1 - \frac{C_P}{C_F} \quad (17-1)$$

where Rej = rejection, dimensionless (expressed as a fraction)

C_P = concentration in permeate, mol/L

C_F = concentration in feed water, mol/L

Rejection can be calculated for bulk parameters such as TDS or conductivity. For membrane rating, however, rejection of specific salts is specified. Sodium chloride rejection is normally specified for high-pressure RO membranes, whereas MgSO_4 rejection is often specified for NF or low-pressure RO membranes.

Nanofiltration membranes can also be characterized by **MWCO**. The MWCO of NF membranes is typically determined by passage of solutes such as sodium chloride and magnesium sulfate. **The MWCO of NF membranes is typically 1000 Daltons** (Da), also known as atomic mass units (amu), or less.

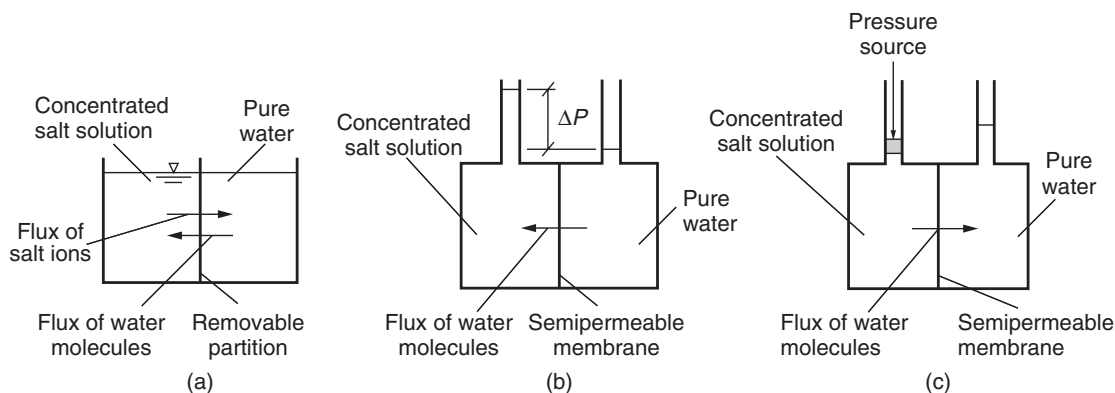
Osmosis is the flow of solvent through a semipermeable membrane, from a dilute solution into a concentrated one. Osmosis reduces the flux through an RO membrane by inducing a driving force for flow in the opposite direction.

The physicochemical foundation for osmosis is rooted in the thermodynamics of diffusion, as described in this section.

DIFFUSION AND OSMOSIS

Consider a vessel with a removable partition that is filled with two solutions to exactly the same level, as shown on **Fig. 17-8a**. The left side is filled with

Osmotic Pressure

**Figure 17-8**

Diffusion sketch for reverse osmosis: (a) diffusion, (b) osmosis, and (c) reverse osmosis.

a concentrated salt solution, the right with pure water, and the partition is gently removed without disturbing the solutions. Initially, the contents are in a nonequilibrium state and the salt will eventually diffuse through the water until the concentration is the same throughout the vessel. With salt ions diffusing from left to right across the plane originally occupied by the partition, conservation of mass requires a flux of water molecules in the opposite direction. Without a flux of water molecules from right to left, mass accumulates on the right side of the container, which is unthinkable with a continuous water surface. Equilibrium requires mass transport in both directions.

On Fig. 17-8b, the top of the vessel has been closed and fitted with manometer tubes and the removable partition has been replaced with a semipermeable membrane. The semipermeable membrane allows the flow of water but prevents the flow of salt. Filling the chambers with salt solution and pure water again creates a thermodynamically unstable system, which must be equilibrated by diffusion. Because the membrane prevents the flux of salt, however, mass accumulates in the left chamber, causing the water level in the left manometer to rise and in the right manometer to drop. This flow of water from the pure side to the salt solution is osmosis. Water flux occurs despite the difference in hydrostatic pressure that develops due to the difference in manometer levels.

OSMOTIC PRESSURE

The driving force for diffusion is typically described as a concentration gradient, although a more rigorous thermodynamic explanation is a gradient in Gibbs energy (Laidler and Meiser, 1999). The concept of Gibbs energy (G) and its relationship to concentration were introduced in Chap. 5. When the vessels on Fig. 17-8 were filled with water and salt solutions, the two

sides had different values of Gibbs energy due to differences in salt concentration. Equilibrium is defined thermodynamically when $\Delta G = 0$, so the gradient in Gibbs energy across the first vessel caused the simultaneous diffusion of salt ions and water molecules, and the system was driven toward an equilibrium condition in which the Gibbs energy (and concentration and water level) was equal throughout the system. In the second vessel, water stops flowing from right to left when the vessel reaches thermodynamic equilibrium but both pressure and concentration are unequal between the chambers. Although Gibbs energy is constant throughout the second vessel at equilibrium, the Gibbs energy includes components to account for both the pressure and concentration differences.

The discussion of Gibbs energy in Chap. 5 was done under conditions of constant temperature and pressure. To describe osmosis, a more general description of Gibbs energy is needed. The general form of the Gibbs function is

$$\partial G = V \partial P - S \partial T + \sum_i \mu_i^\circ \partial n_i \quad (17-2)$$

where $G =$ Gibbs energy, J
 $V =$ volume, m^3
 $P =$ pressure, Pa
 $S =$ entropy, J/K
 $T =$ absolute temperature, K ($273 + ^\circ\text{C}$)
 $\mu_i^\circ =$ chemical potential of solute i , J/mol
 $n_i =$ amount of solute i in solution, mol

Chemical potential is defined as the change in Gibbs energy resulting from a change in the amount of component i when temperature and pressure are held constant:

$$\mu_i^\circ = \left. \frac{\partial G}{\partial n_i} \right|_{P,T} \quad (17-3)$$

Therefore, the last term in Eq. 17-2 ($\mu_i^\circ \partial n_i$) describes the difference in Gibbs energy resulting from the difference in the amount of solute between the chambers (when volume is constant, the difference in amount equals the difference in concentration). Under constant-temperature conditions (i.e., $\partial T = 0$), Eq. 17-2 says equilibrium ($\partial G = 0$) will be achieved when the sum of the Gibbs energy gradient due to chemical potential is offset by the pressure gradient between the two chambers:

$$\partial G = 0 \quad \text{when} \quad V \partial P = - \sum_i \mu_i^\circ \partial n_i \quad (17-4)$$

The pressure required to balance the difference in chemical potential of a solute is called the *osmotic pressure* and is given the symbol π . When the

vessel in the second experiment reaches equilibrium, the difference in hydrostatic pressure between the manometers is equal and opposite to the difference in osmotic pressure between the two chambers. An equation for osmotic pressure can be derived thermodynamically using assumptions of incompressible and ideal solution behavior:

$$\pi = \frac{-RT}{V_b} \ln x_W \quad (17-5)$$

where π = osmotic pressure, bar

V_b = molar volume of pure water, L/mol

x_W = mole fraction of water, mol/mol

R = universal gas constant, 0.083145 L·bar/mol · K

In dilute solution (i.e., $x_W \cong 1$), Eq. 17-5 can be approximated by the van't Hoff equation for osmotic pressure (Eq. 17-6), which is identical in form to the ideal gas law ($PV = nRT$):

$$\pi = \frac{n_S}{V} RT \quad \text{or} \quad \pi = CRT \quad (17-6)$$

where n_S = total amount of all solutes in solution, mol

C = concentration of all solutes, mol/L

V = volume of solution, L

Equation 17-6 was derived assuming infinitely dilute solutions, which is often not the case in RO systems. To account for the assumption of diluteness, the nonideal behavior of concentrated solutions, and the compressibility of liquid at high pressure, a nonideality coefficient (osmotic coefficient ϕ) must be incorporated into the equation:

$$\pi = \phi CRT \quad (17-7)$$

where ϕ = osmotic coefficient, unitless

It should be noted that the thermodynamic equation for osmotic pressure (Eq. 17-5) contains no terms identifying the solute. Osmotic pressure is strictly a function of the concentration, or mole fraction, of water in the system. Solute reduce the mole fraction of water, and the effect of multiple solutes is additive because they cumulatively reduce the mole fraction of water. Solute that dissociate also have an additive effect on the mole fraction of water (e.g., addition of 1 mol of NaCl produces 2 mol of ions in solution, doubling the osmotic pressure compared to a solute that does not dissociate). If multiple solutes are added on an equal-mass basis, the solute with the lowest molecular weight produces the greatest osmotic pressure. The use of Eq. 17-7 is demonstrated in Example 17-1.

Example 17-1 Osmotic pressure calculations

Calculate the osmotic pressure of 1000-mg/L solutions of the following solutes at a temperature of 20°C assuming an osmotic coefficient of 0.95: (1) NaCl, (2) SrSO₄, and (3) glucose (C₆H₁₂O₆). Note that NaCl and SrSO₄ both dissociate into 2 ions when dissolved into water.

Solution

1. Determine the osmotic pressure for NaCl, first by calculating the molar concentration of ions and then using Eq. 17-7:

$$C = \frac{(2 \text{ mol ion/mol NaCl})(1000 \text{ mg/L})}{(10^3 \text{ mg/g})(58.4 \text{ g/mol})} = 0.0342 \text{ mol/L}$$

$$\begin{aligned} \pi &= \phi CRT = (0.95)(0.0342 \text{ mol/L})(0.083145 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(293 \text{ K}) \\ &= 0.79 \text{ bar} \end{aligned}$$

2. Determine the osmotic pressure for SrSO₄:

$$C = \frac{(2 \text{ mol ion/mol SrSO}_4)(1000 \text{ mg/L})}{(10^3 \text{ mg/g})(183.6 \text{ g/mol})} = 0.0109 \text{ mol/L}$$

$$\begin{aligned} \pi &= (0.95)(0.0109 \text{ mol/L})(0.083145 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(293 \text{ K}) \\ &= 0.25 \text{ bar} \end{aligned}$$

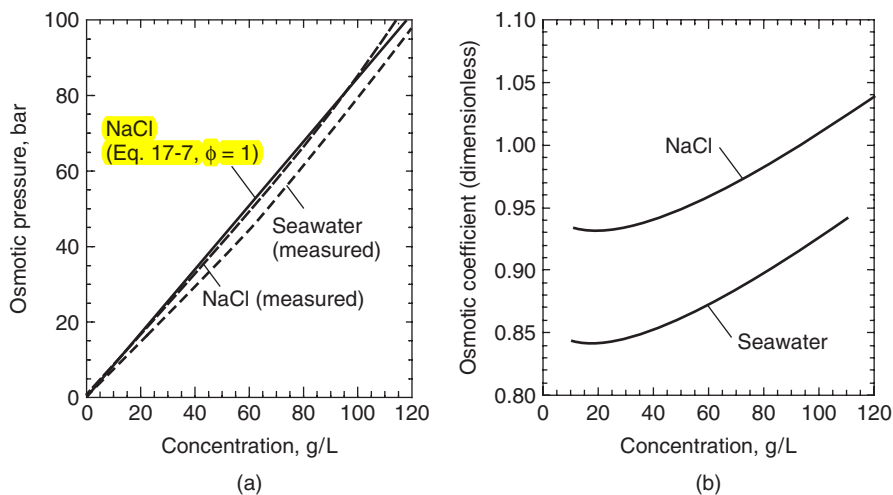
3. Determine the osmotic pressure for glucose (no dissociation):

$$C = \frac{1000 \text{ mg/L}}{(10^3 \text{ mg/g})(180 \text{ g/mol})} = 0.0056 \text{ mol/L}$$

$$\begin{aligned} \pi &= (0.95)(0.00556 \text{ mol/L})(0.083145 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(293 \text{ K}) \\ &= 0.13 \text{ bar} \end{aligned}$$

Comment

Each solution contains the same mass of solute. Because NaCl and SrSO₄ dissociate into two ions, the molar ion concentration is double the molar concentration of added salt. The NaCl has a higher osmotic pressure because it has a lower molecular weight. Even though SrSO₄ and glucose have nearly the same molecular weight, the osmotic pressure of SrSO₄ is nearly double that of glucose because it dissociates.

**Figure 17-9**

(a) Osmotic pressure of aqueous solutions of sodium chloride. (b) Osmotic coefficients for sodium chloride and seawater (osmotic coefficient for seawater with the van't Hoff equation is based on a concentration of NaCl equal to the TDS of the seawater).

The osmotic pressure of a solution of sodium chloride, calculated with Eq. 17-7 and $\phi = 1$, is shown on Fig. 17-9a along with experimentally determined values. Over the range of salt concentrations of interest in seawater desalination, the osmotic coefficient for sodium chloride ranges from 0.93 to 1.03 and is shown as a function of solution concentration on Fig. 17-9b. Osmotic coefficients for other electrolytes are available in Robinson and Stokes (1959). The deviation between measured and calculated values of osmotic pressure can be significantly greater for other solutes and higher concentrations, as shown for sucrose solutions on Fig. 17-10.

Reported values for the osmotic pressure of seawater (Sourirajan, 1970) are about 10 percent below measured values for sodium chloride, as shown on Fig. 17-9a, due to the presence of compounds with a higher molecular weight than sodium chloride. The osmotic pressure for seawater can be calculated with Eq. 17-7 and an equivalent concentration of sodium chloride by using the osmotic coefficient for seawater shown on Fig. 17-9b.

Two opposing forces contribute to the rate of water flow through the semipermeable membrane on Fig. 17-8b: (1) the concentration gradient and (2) the pressure gradient. These opposing forces are exploited in RO. Consider a new experiment using the apparatus on Fig. 17-8, modified so that it is possible to exert an external force on the left side, as shown on Fig. 17-8c. Applying a force equivalent to the osmotic pressure places the system in thermodynamic equilibrium, and no water flows. Applying a force in excess of the osmotic pressure places the system in nonequilibrium,

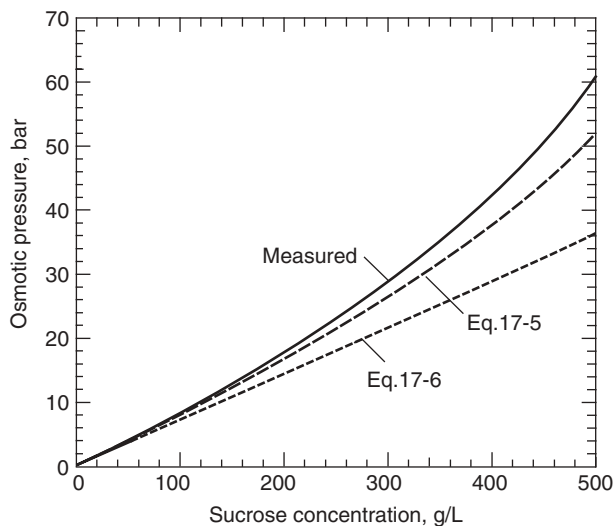


Figure 17-10

Osmotic pressure of aqueous solutions of sucrose.

with a pressure gradient exceeding the chemical potential gradient. Liquid would flow from left to right, that is, from the concentrated solution to the dilute solution. The process of causing water to flow from a concentrated solution to a dilute solution across a semipermeable membrane by the application of an external pressure in excess of the osmotic pressure is called reverse osmosis.

Models have been developed to describe the flux of water and solutes through RO membranes using two basic approaches. The first approach relies on fundamental thermodynamics and does not depend on a physical description of the membrane. The other approach uses physical and chemical descriptions of the membrane and feed solution, such as membrane thickness and porosity. Mathematical development of the models that include descriptions of the membrane and feed solution is beyond the scope of this text but can be found in the published literature (Cheryan and Nichols, 1992; Lonsdale, 1972; Lonsdale et al., 1965; Merten, 1966; Noordman and Wesselingh, 2002; Reid, 1972; Spiegler and Kedem, 1966; Wiesner and Aptel, 1996). For a student learning about RO, the important issue is to develop a conceptual understanding of how water and solutes pass through RO membranes. To promote this understanding, a basic qualitative description of the solution–diffusion, pore flow, and preferential sorption–capillary flow models are presented in the following sections.

SOLUTION–DIFFUSION MODEL

The solution–diffusion model (Lonsdale et al., 1965) describes permeation through a dense membrane where the active layer is permeable but does not

**Models for Water
and Solute
Transport
through RO
Membranes**

have pores. Water and solutes dissolve into the solid membrane material, diffuse through the solid, and reappear on the permeate side of the membrane. Dissolution of water and solutes into a solid material occurs if the solid is loose enough to allow individual water and solute molecules to travel along the interstices between polymer molecules of the membrane.

Liquids behave as liquids because of attractive interactions with surrounding liquid molecules. Thus, even if water molecules travel along a defined path (which hypothetically could be called a pore), they are surrounded by polymer molecules and not other water molecules and therefore are dissolved in the solid, not present as a liquid phase. Diffusion occurs by movement of the water and solute molecules in the direction of the Gibbs energy gradient. Separation occurs when the flux of the water is different from the flux of the solutes.

Equation 17-7 describes a proportionality between osmotic pressure and concentration. Therefore, the driving force (Gibbs energy gradient) for any component can be written equivalently in terms of either pressure or concentration provided the mass transfer coefficient has the proper units. For water, the driving force is expressed in terms of the net pressure gradient, that is, the applied pressure in excess of the osmotic pressure. Solute transport is expressed in terms of the concentration gradient, and most models neglect the effect of applied pressure on solute transport. Flux through the membrane is determined by both solubility and diffusivity. Components of low solubility have a low driving force, and components of low diffusivity have a low diffusion coefficient. The solution–diffusion model predicts that separation occurs because the solubility, diffusivity, or both of the solutes are much lower than those of water, resulting in a lower solute concentration in the permeate than in the feed.

PORE FLOW MODELS

The solution–diffusion model does not consider convective flow through the membrane. Other models consider RO membranes to have void spaces (pores) through which liquid water travels. The pore flow models consider water and solute fluxes to be coupled, meaning the solutes are convected through the membrane with the water. Thus, rejection occurs through mechanisms similar to those described in Chap. 12 for membrane filtration, meaning the solute molecules are “strained” at the entrance to the pores. Because solute and water molecules are similar in size, the rejection mechanism is not a physical sieving and must consider chemical effects such as electrostatic repulsion between the ions and membrane material.

PREFERENTIAL SORPTION–CAPILLARY FLOW MODEL

A third description of water and salt permeation through membranes is provided by the preferential sorption–capillary flow model, which assumes that the membrane has pores. Separation occurs when one component of the feed solution (either the solute or the water) is preferentially

adsorbed to the pore walls and is transported through the membrane by surface diffusion. Membrane materials with a low dielectric constant, such as cellulose acetate, repel ions and preferentially adsorb water, forming a sorbed layer with a reduced concentration of salts. The sorbed layer moves through the membrane by surface diffusion, leaving behind solution components that are repelled from the membrane surface. Separation is a function of the surface chemistry of the membrane and solutes, rather than pore dimensions, although the maximum pore dimension to effect good removal of solutes is two times the thickness of the adsorbed layer, as shown on Fig. 17-11.

COUPLING

Other models consider a combination of permeation mechanisms. The solution-diffusion-imperfection model (Sherwood et al., 1967) assumes that water and solute permeate the membrane by both solution-diffusion and pore flow. The permeation by solution-diffusion is uncoupled but the pore flow is completely coupled. The flux of water by solution-diffusion is proportional to the net applied pressure ($\Delta P - \Delta \pi$), the diffusion of solutes is proportional to the concentration gradient (ΔC), and pore flow is proportional to the applied pressure gradient (ΔP). To achieve high rejection, the pore flow must be a small fraction of the total flow.

In addition to coupling between water and solutes, coupling between solutes must be considered. Electroneutrality must be maintained in both the permeate and the concentrate streams. Thus, preferential transport of ions of one charge can influence the transport of ions of the opposite charge. For instance, negative rejection of hydrogen ions (the concentration of hydrogen ions in the permeate is higher than in the feed solution, manifested as a lower pH in the permeate) is typically observed in RO operations. This occurs because of higher flux of negatively charged ions, such as chloride, than the salt's coion, sodium. Because hydrogen ions are more mobile than sodium ions, the flux of hydrogen ions increases to maintain electroneutrality in the permeate.

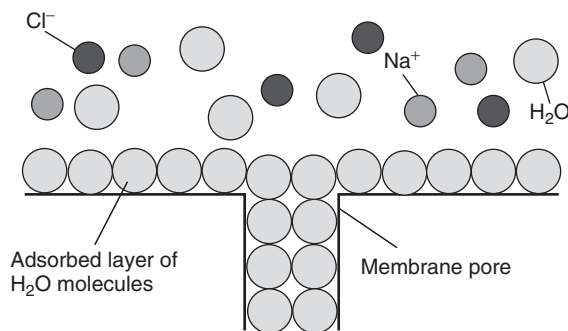


Figure 17-11

Preferential-sorption capillary flow model. Ions are repelled from the membrane surface, resulting in an adsorbed layer of water. The adsorbed water flows through capillary pores in the membrane surface, and the repelled species are left in the feed solution. Good separation can be obtained if the pore diameter is less than 2 times the adsorbed layer thickness.

Mechanisms of Solute Rejection

The membrane permeation models suggest various mechanisms for rejection. The basic mechanisms of rejection are electrostatic repulsion at the membrane surface, solubility and diffusivity through the membrane material due to chemical effects, or straining due to the size and other chemical properties of molecules.

Reverse osmosis and NF membranes are often negatively charged in operation because of the presence of ionized functional groups, such as carboxylates, in the membrane material. Negatively charged ions may be rejected at the membrane surface due to electrostatic repulsion, and positively charged ions may be rejected to maintain electroneutrality in the feed and permeate solutions. The presence of polar and hydrogen-bondable functional groups in the membrane increases the solubility of polar compounds such as water over nonpolar compounds, providing a mechanism for greater flux of water through the membrane. Large molecules would be expected to have lower diffusivity through the membrane material or be unable to pass through the membrane at all.

Experimental observations are consistent with these rejection mechanisms. Small polar molecules such as water generally have the highest flux. Dissolved gases such as H_2S and CO_2 , which are small, uncharged, and polar, also permeate RO membranes well. Monovalent ions such as Na^+ and Cl^- permeate better than divalent ions (Ca^{2+} , Mg^{2+}) because the divalent ions have greater electrostatic repulsion. Acids and bases (HCl , NaOH) permeate better than their salts (Na^+ , Cl^-) because of decreased electrostatic repulsion.

Silica is present in water as uncharged silicic acid (H_4SiO_4) below the $\text{p}K_a$ of 9.84 and is poorly rejected by RO membranes. Similarly, boron is present in water as uncharged boric acid (H_3BO_3) below the $\text{p}K_a$ of 9.24 and also permeates well. The poor removal of boron, coupled with a 1 mg/L notification level in California, often requires specific design considerations for seawater RO systems in that state, such as design of two-pass systems. Increasing the pH to above the $\text{p}K_a$ values results in good removal for both silica and boron.

Within a homologous series, permeation increases with decreasing molecular weight. High-molecular-weight organic materials do not permeate RO membranes at all. Reverse osmosis membranes are capable of rejecting up to 99 percent of monovalent ions. Nanofiltration membranes reject between 80 and 99 percent of divalent ions while achieving low rejection of monovalent ions.

Equations for Water and Solute Flux

Based on the models presented above, a variety of equations have been developed for the rate of water and solute mass transfer through an RO membrane. Ultimately, these models express flux as the product of a mass transfer coefficient and a driving force. The driving force for water flux

through RO membranes is the net pressure differential, or the difference between the applied and osmotic pressure differentials:

$$\Delta P_{\text{NET}} = \Delta P - \Delta \pi = (P_F - P_P) - (\pi_F - \pi_P) \quad (17-8)$$

where ΔP_{NET} = net transmembrane pressure, bar

Subscripts F and P refer to the **feed** and **permeate**, respectively.

The water flux through RO membranes is described by the expression

$$J_W = k_W(\Delta P - \Delta \pi) \quad (17-9)$$

where J_W = volumetric flux of water, $\text{L}/\text{m}^2 \cdot \text{h}$

k_W = mass transfer coefficient for water flux, $\text{L}/\text{m}^2 \cdot \text{h} \cdot \text{bar}$

Water flux is normally reported as a volumetric flux ($\text{L}/\text{m}^2 \cdot \text{h}$ or $\text{gal}/\text{ft}^2 \cdot \text{d}$) and the mass transfer coefficient is typically reported with units of $\text{L}/\text{m}^2 \cdot \text{h} \cdot \text{bar}$ or $\text{gal}/\text{ft}^2 \cdot \text{d} \cdot \text{atm}$. Equation 17-9 is valid at any point on the membrane surface between the feed water entrance and concentrate discharge in a membrane element, but it should be noted that both applied and osmotic pressures change continuously along the length of a spiral-wound element due to head loss and the changing solute concentration. As a result, overall flux must be determined by integrating Eq. 17-9 across the length of the membrane element, as will be demonstrated in the design section of this chapter.

The **driving force for solute flux** is the concentration gradient, and the flux of solutes through RO membranes is expressed as

$$J_S = k_S(\Delta C) \quad (17-10)$$

where J_S = mass flux of solute, $\text{mg}/\text{m}^2 \cdot \text{h}$

k_S = mass transfer coefficient for solute flux, $\text{L}/\text{m}^2 \cdot \text{h}$ or m/h

ΔC = concentration gradient across membrane, mg/L

Solute flux is normally reported as a **mass flux** with units of $\text{mg}/\text{m}^2 \cdot \text{h}$ or $\text{lb}/\text{ft}^2 \cdot \text{d}$. Values of k_W and k_S are determined experimentally by membrane manufacturers. The solute concentration in the permeate is the ratio of the fluxes of solutes and water, as shown by

$$C_P = \frac{J_S}{J_W} \quad (17-11)$$

Thus, the lower the flux of solutes or the higher the flux of water, the better removal of solutes is achieved and the permeate will have a lower solute concentration. The ratio of permeate flow to feed water flow, or **recovery**, is calculated as

$$r = \frac{Q_P}{Q_F} \quad (17-12)$$

where $Q = \text{flow, m}^3/\text{s}$
 $r = \text{recovery, dimensionless}$

Using flow and mass **balance principles**, the solute concentration in the concentrate stream can be calculated from the recovery and solute rejection. The pertinent flow and mass balances using flow and concentration terminology as shown on Fig. 17-1 are

$$\text{Flow balance: } Q_F = Q_P + Q_C \quad (17-13)$$

$$\text{Mass balance: } C_F Q_F = C_P Q_P + C_C Q_C \quad (17-14)$$

where $C = \text{concentration, mol/L or mg/L}$

Combining the mass and flow balances with Eq. 17-1 (rejection) and Eq. 17-12 (recovery) yields the following expression for the solute concentration in the concentrate stream:

$$C_C = C_F \left[\frac{1 - (1 - \text{Rej})r}{1 - r} \right] \quad (17-15)$$

where $\text{Rej} = \text{rejection (dimensionless, expressed as a fraction)}$

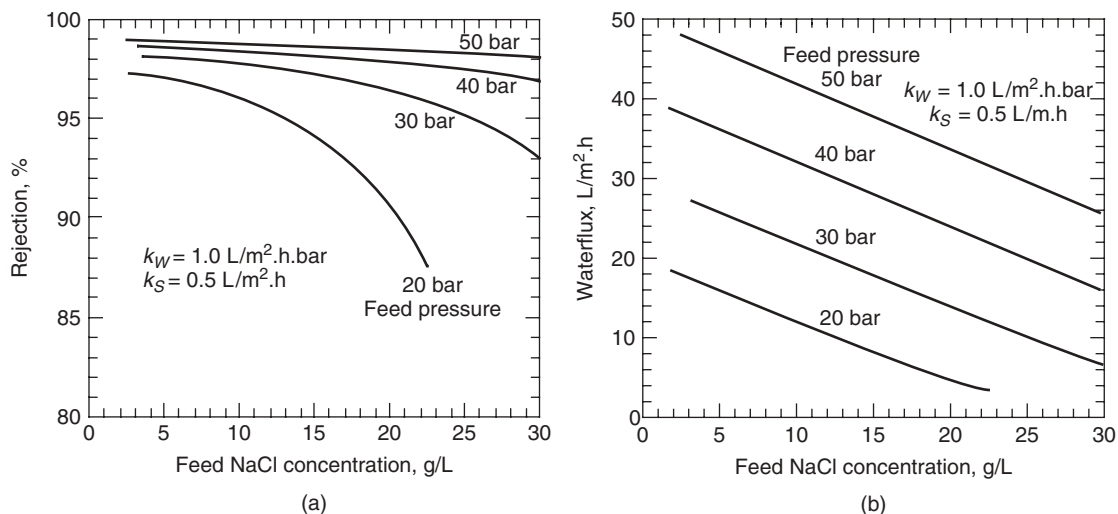
Rejection is frequently close to 100 percent, in which case Eq. 17-15 can be simplified as follows:

$$C_C = C_F \left(\frac{1}{1 - r} \right) \quad (17-16)$$

As shown in Eqs. 17-9 and 17-10, **water flux depends on the pressure gradient** and **solute flux depends on the concentration gradient**. As feed water solute concentration increases at constant pressure, **the water flux decreases (because of higher $\Delta\pi$)** and the solute flux increases (because of **higher ΔC**), which reduces rejection and causes a deterioration of permeate quality. As the feed water pressure increases, water flux increases but the solute flux is essentially constant. Therefore, as the water flux increases, the permeate solute concentration decreases, and the rejection increases. These relationships are illustrated on **Fig. 17-12**.

Temperature and Pressure Dependence

Membrane performance declines (water flux decreases, solute flux increases) due to fouling and membrane aging. However, fluxes of water and solute also vary because of changes in feed water temperature, pressure, velocity, and concentration. To evaluate the true decline in system performance due to fouling and aging, permeate flow rate and salt passage must be compared at standard conditions. Reverse osmosis design manuals present equations for normalizing RO membrane performance in slightly different ways; the equations presented here are adapted from ASTM (2001e) and AWWA (2007). These procedures incorporate the

**Figure 17-12**

Effect of feed water concentration and pressure on (a) percent solute rejection and (b) water flux.

use of **temperature and pressure correction** factors, evaluated at **standard** (subscript **S**) and measured (subscript **M**) conditions:

$$J_{W,S} = J_{W,M} (\text{TCF}) \frac{\text{NDP}_S}{\text{NDP}_M} \quad (17-17)$$

or

$$Q_{P,S} = Q_{P,M} (\text{TCF}) \frac{\text{NDP}_S}{\text{NDP}_M} \quad (17-18)$$

where TCF = temperature correction factor (defined below), dimensionless

NDP = net driving pressure (defined below), bar

Temperature affects the fluid viscosity and the membrane material. The relationship between membrane material, temperature, and flux is specific to individual products, so TCF values should normally be obtained from membrane manufacturers, who determine values experimentally. If manufacturer TCF values are unavailable, the relationship between flux and fluid viscosity can be approximated by the following expression, which may be appropriate for membranes containing pores:

$$\text{TCF} = (1.03)^{T_S - T_M} \quad (17-19)$$

where T = temperature, $^{\circ}\text{C}$

The standard temperature is typically taken to be 25°C for reverse osmosis operation.

The net driving pressure accounts for changes in feed and permeate pressures, feed channel head loss, and osmotic pressure. In spiral-wound elements, the applied pressure decreases and osmotic pressure increases continuously along the length of the feed–concentrate channel as permeate flows through the membrane. Thus, the net driving pressure must take average conditions into account, as shown in

$$\text{NDP} = \Delta P - \Delta \pi = (P_{\text{FC,ave}} - P_P) - (\pi_{\text{FC,ave}} - \pi_P) \quad (17-20)$$

where $P_{\text{FC,ave}}$ = average pressure in the feed–concentrate channel, bar
 $= \frac{1}{2} (P_F + P_C)$
 P_P = permeate pressure, bar
 $\pi_{\text{FC,ave}}$ = average feed–concentrate osmotic pressure (see below), bar
 π_P = permeate osmotic pressure, bar

Feed, concentrate, and permeate pressures are easily measured using system instrumentation. Osmotic pressure must be calculated from solute concentration using Eq. 17-7. Although osmotic pressure increases continuously along the length of a spiral-wound element, solute concentration normally can only be measured in the feed and concentrate streams. Manufacturers use various procedures for determining the average concentration in the feed–concentrate channel and must be contacted for procedures for calculating the average concentration in the feed–concentrate channel. The two most common approaches for determining the average concentration in the feed channel are (1) an arithmetic average (Eq. 17-21) and (2) the log mean average (Eq. 17-22):

$$C_{\text{FC,ave}} = \frac{1}{2} (C_F + C_C) \quad (17-21)$$

$$C_{\text{FC,ave}} = \frac{C_F}{r} \ln \left(\frac{1}{1-r} \right) \quad (17-22)$$

Because head loss is a function of feed flow and osmotic pressure is a function of solute concentration, the system design must establish standard conditions for these parameters in addition to applied pressure.

Solute flux across the membrane is affected by temperature and solute concentration, so it is standardized by multiplying the measured flux by the TCF and ratio of concentration at standard and measured conditions, as follows:

$$J_{S,S} = J_{S,M} (\text{TCF}) \frac{C_{\text{FC,S}}}{C_{\text{FC,M}}} \quad (17-23)$$

Membrane performance, however, is usually evaluated in terms of salt passage rather than solute flux. Salt passage is defined as the ratio of permeate concentration to feed concentration:

$$SP = \frac{C_P}{C_F} = 1 - Re_j \quad (17-24)$$

where SP = salt passage

By rearranging and substituting Eqs. 17-11, 17-17, and 17-24 into Eq. 17-23, standard membrane performance in terms of salt passage is obtained (ASTM, 2001e) as follows:

$$SP_S = SP_M \left(\frac{NDP_M}{NDP_S} \right) \left(\frac{C_{FC,S}}{C_{FC,M}} \right) \left(\frac{C_{F,M}}{C_{F,S}} \right) \quad (17-25)$$

Rearranging Eq. 17-25 in terms of rejection yields the expression

$$Re_{j_S} = 1 - (1 - Re_{j_M}) \left(\frac{NDP_M}{NDP_S} \right) \left(\frac{C_{F,M}}{C_{F,S}} \right) \left(\frac{C_{FC,S}}{C_{FC,M}} \right) \quad (17-26)$$

In multistage systems, it is necessary to standardize the water flux and recovery for each stage independently. The procedures for standardizing RO performance data are shown in [Example 17-2](#).

Example 17-2 Standardization of RO operating data

An RO system uses a shallow brackish groundwater that averages around 4500 mg/L TDS composed primarily of sodium chloride. Permeate flow is maintained constant, but temperature, pressure, and feed concentration change over time as shown in the table below. The operators need to determine whether fouling has occurred between January and May.

Parameter	Unit	January 1	May 31
Permeate flow	m ³ /d	7500	7500
Feed pressure	bar	34.5	32.1
Concentrate pressure	bar	31.4	29.1
Permeate pressure	bar	0.25	0.25
Feed TDS concentration	mg/L	4612	4735
Permeate TDS concentration	mg/L	212	230
Recovery	%	0.69	0.72
Water temperature	°C	11	18

The pressure vessels contain seven membrane elements. The manufacturer has stated that performance data for this membrane element were developed using the following standard conditions:

Parameter	Unit	Standard
Temperature	°C	25
Feed pressure	bar	30
Permeate pressure	bar	0
Head loss per element	bar	0.4
Feed TDS concentration	mg/L	2000
Permeate TDS concentration	mg/L	100
Recovery	%	80

Determine the change in system performance (permeate flow and salt passage) that has occurred between January 1 and May 31. Assume $\phi = 1.0$.

Solution

1. Calculate the TCF for the **January** operating condition:

$$TCF_{\text{Jan}} = (1.03)^{T_s - T_M} = (1.03)^{25 - 11} = 1.512$$

2. Calculate the **NDP** for the January operating condition.
 - a. Calculate the average molar solute concentration in the feed-concentrate channel using **Eq. 17-22**:

$$C_{CF, \text{Jan}} = \frac{C_F}{r} \ln \left(\frac{1}{1-r} \right) = \frac{4612 \text{ mg/L}}{0.69} \ln \left(\frac{1}{1-0.69} \right)$$

$$= 7828 \text{ mg/L}$$

$$C_{CF, \text{Jan}} = \frac{(7828 \text{ mg/L})(2 \text{ mol ions/mol NaCl})}{(10^3 \text{ mg/g})(58.4 \text{ g/mol})}$$

$$= 0.268 \text{ mol/L}$$

- b. Calculate the osmotic pressure in the feed-concentrate channel using **Eq. 17-7**:

$$\pi_{CF, \text{Jan}} = \phi CRT$$

$$= (0.268 \text{ mol/L})(0.083145 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(284 \text{ K})$$

$$= 6.33 \text{ bar}$$

- c. Calculate the molar concentration and osmotic pressure in the permeate:

$$C_{P, \text{Jan}} = \frac{(212 \text{ mg/L})(2 \text{ mol ions/mol NaCl})}{(10^3 \text{ mg/g})(58.4 \text{ g/mol})} = 0.0073 \text{ mol/L}$$

$$\pi_{P, \text{Jan}} = (0.0073 \text{ mol/L})(0.083145 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(284 \text{ K})$$

$$= 0.17 \text{ bar}$$

- d. Calculate the NDP for the January operating condition using Eq. 17-20:

$$P_{FC,ave} = \frac{1}{2} (P_F + P_C) = \frac{1}{2} (34.5 + 31.4) = 32.95 \text{ bar}$$

$$\begin{aligned} \text{NDP} &= (32.95 \text{ bar} - 0.25 \text{ bar}) - (6.33 \text{ bar} - 0.17 \text{ bar}) \\ &= 26.5 \text{ bar} \end{aligned}$$

3. Repeat the calculations in steps 1 and 2 for the standard condition and the May operating condition. The concentrate pressure is not given for the standard operating condition, but can be calculated from the given head loss information:

$$h_L = (0.4 \text{ bar/element})(7 \text{ elements}) = 2.8 \text{ bar}$$

$$P_C = 30 \text{ bar} - 2.8 \text{ bar} = 27.2 \text{ bar}$$

The remaining calculations are summarized in the table below:

Parameter	Unit	Standard Conditions	January 4 Conditions	May 23 Conditions
TCF		1.0	1.51	1.23
$C_{CF,ave}$	mg/L	4024	7828	8372
π_{CF}	bar	3.36	6.33	6.94
π_P	bar	0.08	0.17	0.19
$P_{CF,ave}$	bar	28.6	32.95	30.6
NDP	bar	25.3	26.5	23.6

4. Calculate the standard permeate flow for each date using Eq. 17-17:

$$\begin{aligned} Q_{W,S(\text{Jan})} &= 7500 \text{ m}^3/\text{d} (1.51) \left(\frac{25.3 \text{ bar}}{26.5 \text{ bar}} \right) \\ &= 10,800 \text{ m}^3/\text{d} \end{aligned}$$

$$\begin{aligned} Q_{W,S(\text{May})} &= 7500 \text{ m}^3/\text{d} (1.23) \left(\frac{25.3 \text{ bar}}{23.6 \text{ bar}} \right) \\ &= 9900 \text{ m}^3/\text{d} \end{aligned}$$

5. Calculate the actual salt passage for each date using Eq. 17-24:

$$SP_{M,\text{Jan}} = \frac{212 \text{ mg/L}}{4612 \text{ mg/L}} = 0.046$$

$$SP_{M,\text{May}} = \frac{230 \text{ mg/L}}{4735 \text{ mg/L}} = 0.049$$

6. Calculate the **standard salt passage** for each date using **Eq. 17-25**:

$$\begin{aligned} SP_{S(\text{Jan})} &= (0.046) \left(\frac{26.5 \text{ bar}}{25.3 \text{ bar}} \right) \left(\frac{4612 \text{ mg/L}}{2000 \text{ mg/L}} \right) \left(\frac{4024 \text{ mg/L}}{7828 \text{ mg/L}} \right) \\ &= 0.057 \end{aligned}$$

$$\begin{aligned} SP_{S(\text{May})} &= (0.049) \left(\frac{23.6 \text{ bar}}{25.3 \text{ bar}} \right) \left(\frac{4735 \text{ mg/L}}{2000 \text{ mg/L}} \right) \left(\frac{4024 \text{ mg/L}}{8372 \text{ mg/L}} \right) \\ &= 0.052 \end{aligned}$$

Comment

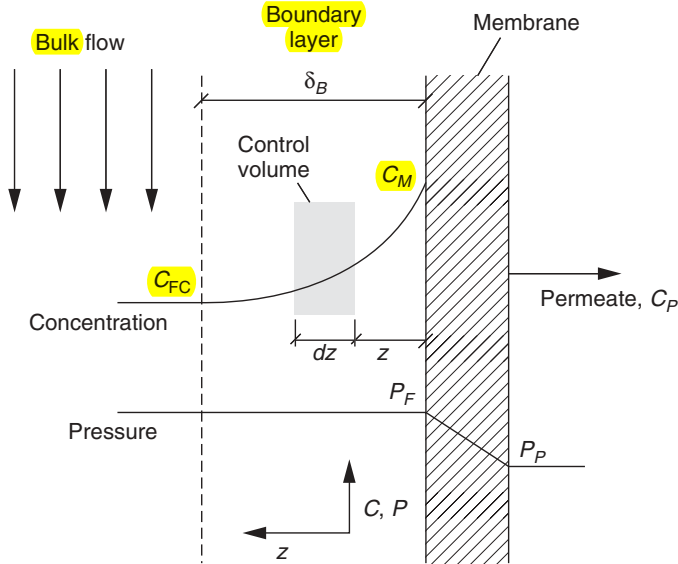
Even though the membrane system is producing the same permeate flow with less pressure in May than in January, there has been a **8 percent loss** of system performance because the standard permeate flow has declined **from 10800 to 9900 m³/d**. The **standard salt passage also decreased** between January and May, even though a higher permeate concentration was observed.

Concentration Polarization

Concentration polarization (CP) is **the accumulation of solutes near the membrane surface** and has **adverse effects** on membrane performance. The flux of water through the membrane **brings feed water** (containing water and solute) to the membrane surface, and as clean water flows through the membrane, **the solutes accumulate near the membrane surface**. In **membrane filtration**, particles contact the membrane and form a **cake layer**. Because the rejection **mechanisms for reverse osmosis are different**, solutes stay in solution and form a **boundary layer of higher concentration at the membrane surface**. Thus, the concentration in the feed solution becomes polarized, with **the concentration at the membrane surface higher** than the concentration in the **bulk** feed water in the feed channel.

Concentration polarization **has several negative impacts on RO** performance:

1. Water flux is lower because the osmotic pressure gradient is higher due to the higher concentration of solutes at the membrane surface.
2. Rejection is lower due to an increase in solute transport across the membrane from an increase in the concentration gradient and a decrease in the water flux.
3. Solubility limits of solutes may be exceeded, leading to precipitation and scaling.

**Figure 17-13**

Schematic of concentration polarization.

Equations for concentration polarization can be derived from film theory (see Chap. 7) and mass balances. In the membrane schematic shown on Fig. 17-13, feed water is traveling vertically on the left side of the membrane and water is passing through the membrane to the right. According to film theory, a boundary layer forms at the surface of the membrane. Water and solutes move through the boundary layer toward the membrane surface. As water passes through the membrane, the solute concentration at the membrane surface increases. The concentration gradient in the boundary layer leads to diffusion of solutes back toward the bulk feed water. During continuous operation, a steady-state condition is reached in which the solute concentration at the membrane surface is constant with respect to time because the convective flow of solutes toward the membrane is balanced by the diffusive flow of solutes away from the surface. **The solute flux toward the membrane surface due to the convective flow of water is described by the expression**

$$J_s = J_w C \quad (17-27)$$

A mass balance can be developed at the membrane surface as follows:

$$\text{Mass accumulation} = \text{mass in} - \text{mass out} \quad (17-28)$$

With **no accumulation of mass at steady state**, the solute flux toward the membrane surface must be balanced by fluxes of solute flowing away from

the membrane (due to diffusion) and through the membrane (into the permeate) as follows:

$$\frac{dM}{dt} = 0 = J_W Ca - D_L \frac{dC}{dz} a - J_W C_P a \quad (17-29)$$

where M = mass of solute, g

t = time, s

D_L = diffusion coefficient for solute in water, m²/s

z = distance perpendicular to membrane surface, m

a = surface area of membrane, m²

Equation 17-29 applies not only at the membrane surface but also at any plane in the boundary layer because the net solute flux must be constant throughout the boundary layer to prevent the accumulation of solute anywhere within that layer (the last term in Eq. 17-29 represents the solute that must pass through the boundary layer and the membrane to end up in the permeate). Rearranging and integrating Eq. 17-29 across the thickness of the boundary layer with the boundary conditions $C(0) = C_M$ and $C(\delta_B) = C_{FC}$, where C_{FC} is the concentration in the feed-concentrate channel and C_M is the concentration at the membrane surface, are done in the following equations:

$$D_L \int_{C_M}^{C_{FC}} \frac{dC}{C - C_P} = -J_W \int_0^{\delta_B} dz \quad (17-30)$$

Integrating yields

$$\ln \left(\frac{C_M - C_P}{C_{FC} - C_P} \right) = \frac{J_W \delta_B}{D_L} \quad (17-31)$$

$$\frac{C_M - C_P}{C_{FC} - C_P} = e^{(J_W \delta_B)/D_L} = e^{J_W/k_{CP}} \quad (17-32)$$

where $k_{CP} = D_L/\delta_B$ concentration polarization mass transfer coefficient, m/s

The concentration polarization mass transfer coefficient describes the diffusion of solutes away from the membrane surface. Concentration polarization is expressed as the ratio of the membrane and feed-concentrate channel solute concentrations as follows:

$$\beta = \frac{C_M}{C_{FC}} \quad (17-33)$$

where β = concentration polarization factor, dimensionless

Combining Eq. 17-33 with Eqs. 17-1 and 17-32 results in the following expression:

$$\beta = (1 - \text{Rej}) + \text{Rej} (e^{J_W/k_{CP}}) \quad (17-34)$$

If rejection is high (greater than 99 percent), Eq. 17-34 can be reasonably simplified as follows:

$$\beta = e^{J_W/k_{CP}} \quad (17-35)$$

To predict the extent of concentration polarization, the value of the concentration polarization mass transfer coefficient is needed. As demonstrated in Chap. 7, mass transfer coefficients are often calculated using a correlation between Sherwood (Sh), Reynolds (Re), and Schmidt (Sc) numbers. Correlations for mass transfer coefficients depend on physical characteristics of the system and the flow conditions (e.g., laminar or turbulent). To promote turbulent conditions and minimize concentration polarization in RO membrane elements, spiral-wound elements contain mesh feed channel spacers and maintain a high velocity flow parallel to the membrane surface. The feed channel spacer complicates the flow patterns and promotes turbulence. The superficial velocity (assuming an empty channel) in a spiral-wound element typically ranges from 0.02 to 0.2 m/s, but the actual velocity is higher because of the space taken up by the spacer.

In the spacer-filled feed channel of a spiral-wound element, Schock and Miquel (1987) found that the concentration polarization mass transfer coefficient could be predicted by the following equation, when calculations for the velocity in the channel and the hydraulic diameter took the presence of the spacer into account:

$$k_{CP} = 0.023 \frac{D_L}{d_H} (\text{Re})^{0.875} (\text{Sc})^{0.25} \quad (17-36)$$

$$\text{Re} = \frac{\rho v d_H}{\mu} \quad (17-37)$$

$$\text{Sc} = \frac{\mu}{\rho D_L} \quad (17-38)$$

where

- Re = Reynolds number, dimensionless
- Sc = Schmidt number, dimensionless
- v = velocity in feed channel, m/s
- ρ = feed water density, kg/m³
- μ = feed water dynamic viscosity, kg/m · s
- d_H = hydraulic diameter, m

The hydraulic diameter is defined as

$$d_H = \frac{4 (\text{volume of flow channel})}{\text{wetted surface}} \quad (17-39)$$

For hollow-fiber membranes (circular cross section), the hydraulic diameter is equal to the inside fiber diameter. Spiral-wound membranes can be approximated by flow through a slit, where the width is much larger than the feed channel height ($w \gg h$). In an empty channel (i.e., the spacer is neglected), the hydraulic diameter is twice the feed channel height, as shown in the equation

$$d_H = \frac{4wh}{2w + 2h} \approx 2h \quad (17-40)$$

where h = feed channel height, m
 w = channel width, m

The feed channel height in typical spiral-wound elements ranges from about 0.4 to 1.2 mm and is governed by the thickness of the spacer.

Because the mesh spacer affects mass transfer in the feed channel and many feed spacer configurations have been developed, numerous other correlations have been developed for the mass transfer coefficient. Mariñas and Urama (1996) developed a correlation using the channel height and the superficial velocity, which eliminates the task of determining the parameters of the spacer. Their correlation is

$$k_{CP} = \lambda \frac{D_L}{d_H} (\text{Re})^{0.50} (\text{Sc})^{1/3} \quad (17-41)$$

where λ ranged from 0.40 to 0.54 for different elements. Many spacer configurations have been evaluated in small flat-sheet membrane cells instead of spiral-wound elements, and in those cases, the mass transfer correlation often has an additional term for the ratio of the channel height (d_H) to channel length (L). For instance, the correlation presented by Shakaib et al. (2009) for spacers with axial and transverse filaments is

$$k_{CP} = 0.664 \frac{D_L}{d_H} (\text{Re})^{0.5} (\text{Sc})^{0.33} \left(\frac{d_H}{L} \right)^{0.5} \quad (17-42)$$

Concentration polarization varies along the length of a membrane element; the parameters that change most significantly are the velocity in the feed channel (v) and the permeate flux (J_W). Variation in the concentration polarization factor as a function of these parameters is shown on Fig. 17-14. As might be expected, concentration polarization increases as the permeate flux increases and as the velocity in the feed channel decreases.

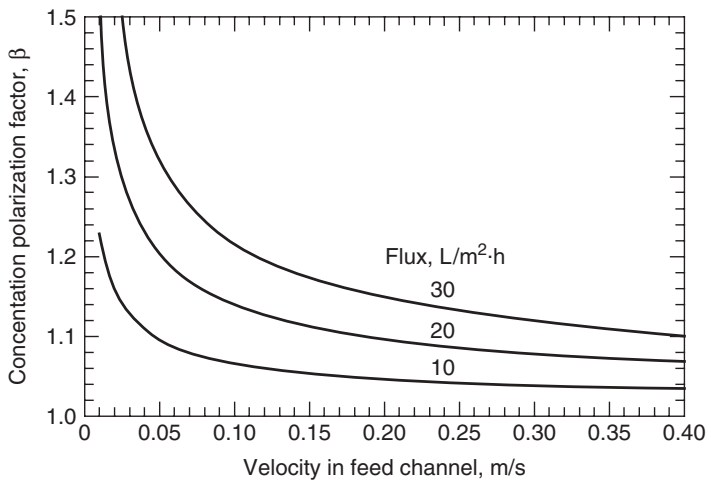


Figure 17-14

Concentration polarization factors as function of feed channel velocity and permeate flux.

The **maximum concentration polarization** allowed for membrane elements is specified by manufacturers; $\beta = 1.2$ is a typical value. The importance of maintaining a high velocity in the feed–concentrate channel, particularly for membranes that achieve higher permeate flux, is clearly demonstrated on Fig. 17-14. Calculation of the concentration polarization factor is illustrated in Example 17-3.

Example 17-3 Concentration polarization

For a **spiral-wound element**, calculate the concentration polarization factor and the concentration of sodium at the membrane surface given the following information: water temperature 20°C, feed channel velocity 0.15 m/s, feed channel height 0.86 mm, permeate flux 25 L/m² · h, sodium concentration 6000 mg/L, and diffusivity of sodium in water 1.35 × 10⁻⁹ m²/s. Use the correlation in Eq. 17-41 and a value of 0.47 for the coefficient. Assume that the **rejection is high enough** that the impact of sodium flux through the membrane can be ignored. Water density and viscosity at 20°C can be found in Table C-1 in App. C.

Solution

1. Calculate the Reynolds and Schmidt numbers using Eqs. 17-37 and 17-38. Because the feed channel height is 0.86 mm, the **hydraulic diameter is 1.72 mm**:

$$Re = \frac{\rho v d_H}{\mu} = \frac{(998 \text{ kg/m}^3)(0.15 \text{ m/s})(1.72 \text{ mm})}{(1.0 \times 10^{-3} \text{ kg/m} \cdot \text{s})(10^3 \text{ mm/m})} = 257$$

$$Sc = \frac{\mu}{\rho D_L} = \frac{1.0 \times 10^{-3} \text{ kg/m} \cdot \text{s}}{(998 \text{ kg/m}^3)(1.35 \times 10^{-9} \text{ m}^2/\text{s})} = 742$$

2. Calculate **k_{CP}** using Eq. 17-41:

$$k_{CP} = \frac{(0.47)(1.35 \times 10^{-9} \text{ m}^2/\text{s})(257)^{0.5}(742)^{1/3}}{(1.72 \text{ mm})(10^{-3} \text{ m/mm})} = 5.36 \times 10^{-5} \text{ m/s}$$

3. Because the rejection is high, β can be calculated using Eq. 17-35 (otherwise, Eq. 17-34 must be used):

$$\beta = \exp\left(\frac{J_W}{k_{CP}}\right) = \exp\left[\frac{(25 \text{ L/m}^2 \cdot \text{h})(10^{-3} \text{ m}^3/\text{L})}{(5.36 \times 10^{-5} \text{ m/s})(3600 \text{ s/h})}\right] = 1.14$$

4. Calculate the sodium concentration at the membrane surface using Eq. 17-33:

$$C_M = (1.14)(6000 \text{ mg/L}) = 6840 \text{ mg/L}$$

17-6 Fouling and Scaling

Nanofiltration and RO membranes are susceptible to fouling via a variety of mechanisms. The primary sources of fouling and scaling are particulate matter, precipitation of insoluble inorganic salts, oxidation of soluble metals, and biological matter.

Particulate Fouling

Particulate fouling is a concern in RO because the operational cycle does not include a backwashing step to remove accumulated solids (in fact, backwashing might cause the active layer of thin-film membranes to separate from the support layers). Virtually all RO systems require pretreatment to minimize particulate fouling. Fouling by residual particulate matter affects the cleaning frequency.

PLUGGING AND CAKE FORMATION

Both inorganic and organic materials, including microbial constituents and biological debris, can cause particulate fouling, which includes plugging and cake formation. Plugging is the entrapment of large particles in the feed channels and piping. Hollow-fine-fiber membranes are reported to have more significant plugging problems because the high packing density of the fibers inside the pressure vessel results in very small spaces between the fibers. The mesh spacers in spiral-wound elements are sized to minimize plugging, but an excessive load of particulate matter may cause plugging anyway. Plugging is minimized by prefiltration of the feed water, and RO membrane manufacturers recommend prefiltration through 5- μm cartridge filters as a minimum prefiltration step for protection of the membrane elements.

Particulate matter forming a cake on the membrane surface adds resistance to flow and affects system performance. Source waters with excessive potential for particulate fouling require advanced pretreatment to lower the particulate concentration to an acceptable level. Coagulation and filtration (using sand, carbon, or other filter media) are sometimes used for pretreatment as well as MF and UF.

ASSESSMENT OF PARTICLE FOULING

It is important to assess the fouling tendency prior to design and construction of an RO facility and to monitor the fouling tendency during operation. Empirical tests have been developed to assess particulate fouling, including the silt density index (SDI) and the modified fouling index (MFI). The SDI (ASTM, 2001b) is a timed filtration test using three time intervals through a gridded membrane filter with a mean pore size of $0.45 \pm 0.02 \mu\text{m}$ and a diameter of 47 mm at a constant applied pressure of 2.07 bar (30 psi). The first interval is the duration necessary to collect 500 mL of permeate. Filtration continues through the second interval without recording volume until 15 min has elapsed (including the first time interval). Occasionally,

a duration shorter than 15 min is used for waters with high fouling tendency. At the end of 15 min, the third interval is started, during which an additional 500-mL aliquot of water is filtered through the now-dirty membrane, and the time is recorded. The SDI is calculated from these time intervals:

$$\text{SDI} = \frac{100(1 - t_I/t_F)}{t_T} \quad (17-43)$$

where SDI = silt density index, min^{-1}

t_I = time to collect first 500-mL sample, min

t_F = time to collect final 500-mL sample, min

t_T = duration of first two test intervals (15 min)

The MFI (Schippers and Verdouw, 1980) uses identical test equipment but different procedures from the SDI. The volume filtered is recorded at 30-s intervals during the MFI test. The flow rate is determined from volume and time data, and the inverse of the flow rate is plotted as a function of volume filtered. An example of the plotted data is shown on Fig. 17-15. A portion of the graph is generally linear, and the MFI is the slope of the graph in this region, that is,

$$\frac{\Delta t}{\Delta V} = \frac{1}{Q} = (\text{MFI})V + b \quad (17-44)$$

where MFI = modified fouling index, s/L^2

V = volume of permeate, L

b = intercept of linear portion of graph

The SDI and MFI have been criticized as being too simplistic to accurately predict RO membrane fouling. They operate in a dead-end, constant-pressure filtration mode, whereas full-scale RO systems operate with a significant cross flow and constant flux. They use a $0.45\text{-}\mu\text{m}$ filter so they only nominally measure fouling by material larger than that size. Research suggests that colloidal matter smaller than $0.45\ \mu\text{m}$ may cause significant fouling of RO membranes. As a result, a revised MFI test that uses a 13-kDa UF membrane has also been developed (Boerlage et al., 2002, 2003).

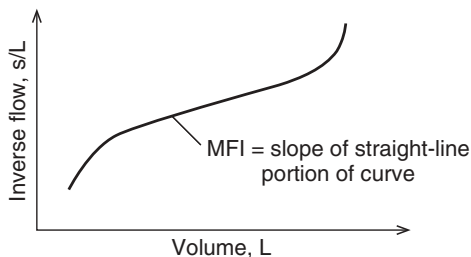
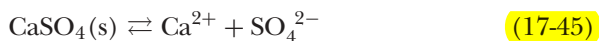


Figure 17-15
Determination of modified fouling index (MFI).

The SDI and MFI might best be considered as screening tests that can indicate unacceptable feed water quality. A high value is a good indicator of fouling problems in RO systems, but a low value does not necessarily mean the source water has a low fouling tendency. RO manufacturers typically specify a maximum SDI value of 4 to 5 min^{-1} . High SDI or MFI values indicate pretreatment is required to remove particulate matter. When lower SDI or MFI values are measured, pilot tests are often necessary to determine the appropriate level of pretreatment to minimize fouling.

Precipitation of Inorganic Salts and Scaling

Inorganic scaling occurs when salts in solution are concentrated beyond their solubility limits and form precipitates. Common sparingly soluble salts are listed in Table 17-3. If the ions that comprise these salts are concentrated past the solubility product, precipitation occurs. Precipitation reactions and solubility calculations were introduced in Chap. 5. The precipitation reaction for a typical salt is as follows:



The solubility product is written as

$$K_{\text{SP}} = \{\text{Ca}^{2+}\}\{\text{SO}_4^{2-}\} = \gamma_{\text{Ca}}[\text{Ca}^{2+}]\gamma_{\text{SO}_4}[\text{SO}_4^{2-}] \quad (17-46)$$

where

$$\begin{aligned} K_{\text{SP}} &= \text{solubility product} \\ \{\text{Ca}^{2+}\} &= \text{calcium activity} \\ \{\text{SO}_4^{2-}\} &= \text{sulfate activity} \\ \gamma_{\text{Ca}} &= \text{activity coefficient for calcium} \\ \gamma_{\text{SO}_4} &= \text{activity coefficient for sulfate} \end{aligned}$$

Table 17-3

Typical limiting salts and their solubility products

Salt	Equation	Solubility Product ($\text{p}K_{\text{sp}}$ at 25°C)
Calcium carbonate (aragonite)	$\text{CaCO}_3(\text{s}) \rightleftharpoons \text{Ca}^{2+} + \text{CO}_3^{2-}$	8.2
Calcium fluoride	$\text{CaF}_2(\text{s}) \rightleftharpoons \text{Ca}^{2+} + 2\text{F}^-$	10.3
Calcium orthophosphate	$\text{CaHPO}_4(\text{s}) \rightleftharpoons \text{Ca}^{2+} + \text{HPO}_4^{2-}$	6.6
Calcium sulfate (gypsum)	$\text{CaSO}_4(\text{s}) \rightleftharpoons \text{Ca}^{2+} + \text{SO}_4^{2-}$	4.6
Strontium sulfate	$\text{SrSO}_4(\text{s}) \rightleftharpoons \text{Sr}^{2+} + \text{SO}_4^{2-}$	6.2
Barium sulfate	$\text{BaSO}_4(\text{s}) \rightleftharpoons \text{Ba}^{2+} + \text{SO}_4^{2-}$	9.7
Silica, amorphous	$\text{SiO}_2(\text{s}) + 2\text{H}_2\text{O} \rightleftharpoons \text{Si}(\text{OH})_4(\text{aq})$	2.7

^aFrom Stumm and Morgan (1996).

$[\text{Ca}^{2+}]$ = calcium concentration, mol/L

$[\text{SO}_4^{2-}]$ = sulfate concentration, mol/L

The ionic strength of feed solutions for RO is sufficiently high that ion products must be calculated using activity, rather than the common practice of assuming that activity is equal to concentration. Several factors in RO operation affect how much ions are concentrated. The system recovery is the most important factor because the concentration of the rejected solutes increases as more clean water is withdrawn from solution. In fact, precipitation is one of the important factors that limit recovery in RO systems (osmotic pressure being the other). The rate of ion or salt rejection is also important, as an ion with 99 percent rejection will be concentrated more than one with 10 percent rejection. Finally, the degree of concentration polarization is important because precipitation occurs in the more concentrated zone near the membrane surface. The inorganic scale that forms on the membrane surface can reduce water permeability or permanently damage the membrane.

In the absence of pretreatment, precipitation must be avoided by minimizing concentration polarization, limiting salt rejection, or limiting recovery. Concentration polarization is minimized by promoting turbulence in the feed channels and maintaining minimum velocity conditions specified by equipment manufacturers. Limiting rejection is impractical because it conflicts with process objectives. Limiting recovery, however, is often necessary to prevent precipitation. The highest recovery possible before any salts precipitate is the *allowable recovery*, and the salt that precipitates at this condition is the *limiting salt*. The most common scales encountered in water treatment applications are calcium carbonate (CaCO_3) and calcium sulfate (CaSO_4).

The allowable recovery without pretreatment that can be achieved in RO is determined by performing solubility calculations for each of the possible limiting salts. The highest solute concentrations occur in the final membrane element immediately prior to the feed water exiting the system as the concentrate stream, so concentrate stream concentrations are used to evaluate solubility limits. In addition, the concentration in the concentrate stream must be adjusted for the level of concentration polarization that is occurring. Incorporating the concentration polarization factor defined in Eq. 17-40 with the expression for the solute concentration in the concentrate stream defined by Eq. 17-15 yields

$$C_M = \beta C_F \left[\frac{1 - (1 - \text{Rej})r}{1 - r} \right] \quad (17-47)$$

Allowable recovery is determined by substituting the activities at the membrane into a solubility product calculation (from Chap. 5) and solving for the recovery, as demonstrated in Example 17-4.

Example 17-4 Allowable recovery from limiting salt calculations

Determine the limiting salt and allowable recovery for a brackish water RO system containing the following solutes: calcium 74 mg/L, barium 0.008 mg/L, and sulfate 68 mg/L. Assume 100 percent rejection of all solutes and a polarization factor of 1.15 and ignore activity coefficients (i.e., activity = concentration).

Solution

1. Calculate the molar concentration for each component:

$$[\text{Ca}^{2+}] = \frac{74 \text{ mg/L}}{(40 \text{ g/mol})(10^3 \text{ mg/g})} = 1.85 \times 10^{-3} \text{ mol/L}$$

$$[\text{Ba}^{2+}] = \frac{0.008 \text{ mg/L}}{(137.3 \text{ g/mol})(10^3 \text{ mg/g})} = 5.83 \times 10^{-8} \text{ mol/L}$$

$$[\text{SO}_4^{2-}] = \frac{68 \text{ mg/L}}{(96 \text{ g/mol})(10^3 \text{ mg/g})} = 7.08 \times 10^{-4} \text{ mol/L}$$

2. Simplify the expression for concentration at the membrane. Let $y = 1 - r$. Because $\text{Rej} = 1$, Eq. 17-47 becomes

$$C_M = \frac{\beta C_F}{y}$$

3. Substitute the concentrations at the membrane surface into the equation for solubility products and calculate recovery. Solubility product constants are available in Table 17-3.
 - a. For calcium sulfate,

$$K_{\text{sp}} = 10^{-4.6} = [\text{Ca}^{2+}]_M [\text{SO}_4^{2-}]_M = \left(\frac{\beta [\text{Ca}^{2+}]_F}{y} \right) \left(\frac{\beta [\text{SO}_4^{2-}]_F}{y} \right)$$

$$= \frac{\beta^2}{y^2} [\text{Ca}^{2+}]_F [\text{SO}_4^{2-}]_F$$

$$y = \left(\frac{\beta^2}{K_{\text{sp}}} [\text{Ca}^{2+}]_F [\text{SO}_4^{2-}]_F \right)^{1/2}$$

$$= \left[\frac{(1.15)^2}{10^{-4.6}} (1.85 \times 10^{-3} \text{ mol/L})(7.08 \times 10^{-4} \text{ mol/L}) \right]^{1/2}$$

$$= 0.26$$

$$r = 1 - y = 1 - 0.26 = 0.74$$

b. For barium sulfate,

$$y = \left[\frac{(1.15)^2}{10^{-9.7}} (5.83 \times 10^{-8} \text{ mol/L})(7.08 \times 10^{-4} \text{ mol/L}) \right]^{1/2}$$

$$= 0.52$$

$$r = 1 - y = 1 - 0.52 = 0.48$$

Comments

1. The allowable recovery before barium sulfate precipitates is 48 percent, compared to 74 percent before calcium sulfate precipitates. Therefore, barium sulfate is the limiting salt and the allowable recovery is 48 percent.
2. Activity coefficients affect solubility calculations and, therefore, recovery. The ionic strength of the feed solution can be calculated from feed ion concentrations. However, the activity coefficients must be calculated from the ionic strength of the concentrate at the allowable recovery, so a simultaneous solution procedure must be used.

The complexity of limiting salt calculations is greatly oversimplified in Example 17-4. As noted above, activity coefficients cannot be ignored. The ionic strength is dependent on recovery and rejection, however, so the activity coefficients cannot be calculated until the recovery is determined. Ignoring ionic strength may yield a significantly lower value for allowable recovery than could actually be achieved. The assumption of 100 percent rejection is often justified because divalent ions typically have rejection near 100 percent. An assumption of 100 percent rejection yields a slightly conservative value for allowable recovery because lower rejection will produce concentrate stream concentrations that are actually slightly lower. For NF and low-pressure RO systems that have divalent ion rejection significantly below 100 percent, however, this assumption would be inappropriate.

Another complicating factor is the formation of ion complexes. For instance, calcium and sulfate form a neutral CaSO_4^0 species that increases the solubility of $\text{CaSO}_4(\text{s})$. The solubility of calcium sulfate in distilled water would be calculated as 680 mg/L as CaSO_4 using Eq. 17-48 if ionic strength and complexation were ignored. With ionic strength and complexation, the solubility of calcium sulfate in distilled water is 2170 mg/L, an error of over 200 percent.

Several models are available to calculate activity coefficients, and the applicability of each model depends on the ionic strength. Seawater has an ionic strength of about 0.7 M. Assuming 50 percent recovery, the ionic strength of the concentrate from a seawater RO plant would be about 1.4 M. This ionic strength is significantly above the range of applicability of the

extended Debye–Huckel or Davies equations. The specific interaction model or Pitzer model are suitable for calculating activity coefficients when the ionic strength is above 1 M (Pitzer, 1975).

Another complicating factor is that carbonate and phosphate concentrations are dependent on pH. As can be imagined, accounting for ionic strength, recovery, complexation, and pH in the calculations in Example 17-4, and then calculating activity coefficients with the Pitzer equations, would result in equations that cannot be easily manipulated algebraically.

Furthermore, the calculations must be repeated for each limiting salt in Table 17-3. Example 17-4 demonstrates that barium was a limiting solute even though its concentration in the feed water was very low. When alternative systems with different rejection capabilities are being evaluated, the calculations must be repeated for each rejection scenario. Temperature and supersaturation considerations further complicate the calculations. Clearly, the computational requirements of limiting salt calculations can be daunting and are rarely done manually. Membrane manufacturers provide computer programs to perform these calculations. These programs account for the concentration polarization factor and rejection capabilities of specific products, temperature and pH effects, and the degree of supersaturation that can be accommodated with various pretreatment strategies. Use of an equilibrium speciation program (Visual MINTEQ) to solve Example 17-4 reveals that the barium sulfate reaches saturation at 84 percent recovery instead of 48 percent recovery.

ACID ADDITION AND ANTISCALANTS TO PREVENT SCALING

Pretreatment is necessary in virtually all RO systems to prevent scaling due to precipitation of sparingly soluble salts. Calcium carbonate precipitation is common, and most systems require pretreatment for this compound. In addition to the limiting salt calculations presented in the above example, calcium carbonate solubility can also be expressed in terms of the Langelier saturation index (LSI) and Stiff and Davis stability index (ASTM, 2001a, 2001f), and manufacturers' solubility programs often report these values. Calcium carbonate precipitation can be prevented by adjusting the pH of the feed stream with acid to convert carbonate to bicarbonate and carbon dioxide. Sulfuric or hydrochloric acids are normally used, but using sulfuric acid can increase the sulfate concentration enough to cause precipitation of sulfate compounds. The pH of most RO feed waters is adjusted to a pH value of 5.5 to 6.0. At this pH, most carbonate is in the form of carbon dioxide and passes through the membrane.

Scaling of other limiting salts is commonly prevented with the addition of antiscalant chemicals. Antiscalants allow supersaturation without precipitation occurring by preventing crystal formation and growth. At one time, sodium hexametaphosphate (SHMP) was commonly used as an antiscalant, but it is rarely used anymore because it has limited ability to extend the supersaturation range and adds phosphate compounds to the concentrate,

which causes disposal problems. SHMP has been largely replaced with polymeric antiscalants. The degree of supersaturation allowed because of antiscalant addition depends on properties of the antiscalant, which are often proprietary, and characteristics of specific equipment configurations. It is appropriate to rely on the recommendations of equipment and antiscalant manufacturers when determining appropriate antiscalant selection and doses necessary for a specific feed water analysis and design recovery.

In addition to acid and antiscalant addition, newer installations are incorporating a variety of strategies to minimize scaling with the goal of reducing the quantity of waste concentrate that must be disposed and increasing the recovery of water. These strategies are discussed in more detail in Sec. 17-7 under the heading Concentrate Management.

SILICA SCALING

Silica scaling can be particularly problematic because silica chemistry is complex and silica can occur in several forms in groundwater, including monomeric, polymeric, and colloidal forms. Many brackish groundwater sources in the Southwestern United States have sufficiently high silica concentrations such that silica is the species that limits recovery. Silica precipitates in an amorphous rather than crystalline form; thus, antiscalants that prevent crystal growth are ineffective for preventing silica precipitation. The presence of metals can increase silica precipitation and change its form (Sahachaiyunta et al., 2002; Sheikholeslami and Bright, 2002), complicating the presence of silica in RO feed water. Recent advances and new antiscalant formulations are now available for both minimizing silica precipitation and cleaning silica from membranes, but these proprietary compounds have had varying degrees of success. When high silica concentrations are present, high-pH softening (resulting in co-precipitation with magnesium hydroxide) may be necessary to remove silica from the feed water to prevent precipitation on the membrane.

A cost trade-off exists between methods of preventing scaling: operating at a lower recovery or the use of pretreatment processes and chemicals. In some cases, it may be more cost effective to operate at a lower recovery to minimize pretreatment costs. Pretreatment and membrane equipment costs must be considered simultaneously and the design recovery set at the point that minimizes overall system costs.

Groundwater used as the source water for RO and NF systems is often anaerobic. Iron and manganese, soluble compounds in their reduced states, can oxidize, precipitate, and foul membranes if oxidants enter the feed water system. Iron fouling is more prevalent and can occur rapidly if any air enters the feed system. Fouling may be avoided by preventing oxidation or removing the iron or manganese after oxidation. If iron concentrations are low, precautions to prevent air from entering the feed system may be sufficient;

**Metal Oxide
Fouling**

antiscalants often include additives to minimize fouling by low concentrations of iron. Pretreatment to control iron might include oxidation with oxygen or chlorine followed by adequate mixing and hydraulic detention time and granular media or membrane filtration or greensand filtration in which oxidation and filtration take place simultaneously. When oxidants are used, precautions must be made to prevent them from reaching the membranes, particularly for polyamide membranes or other materials that are not oxidant resistant. Iron-fouling deposits are usually removable from RO membrane surfaces by commercially available cleaning agents and procedures.

An additional constituent present in many anaerobic groundwaters is hydrogen sulfide. If air enters the feed water system, hydrogen sulfide can oxidize to colloidal sulfur, which can foul membranes. As with iron oxidation, precautions to prevent air from entering the feed system are important to prevent colloidal sulfur fouling. Sulfur deposits on membrane surfaces are typically irreversible.

Biological Fouling

Biological fouling refers to the attachment or growth of microorganisms or extracellular soluble material on the membrane surface or in the membrane element feed channels. Biological fouling is common in many RO systems and can have a variety of negative effects on performance, including loss of flux, reduced solute rejection, increased head loss through the membrane modules, contamination of the permeate, degradation of the membrane material, and reduced membrane life (Ridgway and Flemming, 1996). An example of biological fouling is shown on Fig. 17-16. The primary source of microbial contamination is the feed water. Biological fouling is a significant problem in many RO systems.

Biological fouling is prevented by maintaining proper operating conditions, applying biocides, and flushing membrane elements properly when not in use. Many RO and NF feed waters (groundwater in many cases) have low microbial populations. When operated properly, the shear in the feed channels helps to keep bacteria from accumulating or growing to unacceptable levels. When membrane trains are out of service, however, bacteria can quickly multiply. To avoid this problem, membranes should be flushed with permeate periodically or filled with an approved biocide if out of service for any significant period. Chlorine solutions can be used as a biocide for cellulose acetate membranes within recommended limits, but other chemicals such as sodium bisulfite must be used with polyamide membranes because of their susceptibility to degradation by chlorine. An excellent review of the issues involved in biological fouling of membranes is provided in Ridgway and Flemming (1996).

The feed water to cellulose acetate membranes can be continuously chlorinated within limited concentrations to prevent biological growth, if necessary. Ultraviolet radiation, chloramination, or chlorination followed by dechlorination can sometimes be used for polyamide membranes.

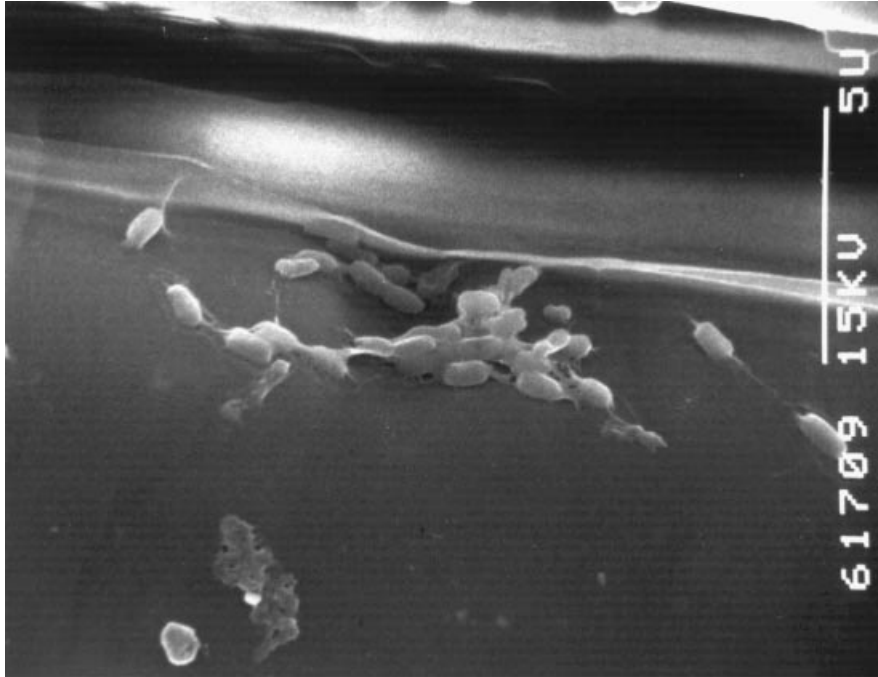


Figure 17-16

Scanning electron micrograph (SEM) image of biological fouling of membrane. (Courtesy Orange County Water District.)

17-7 Reverse Osmosis Process Design

During preliminary design of an RO system, the design engineer must perform the following activities:

1. Select the basic performance criteria: capacity, recovery, rejection, and permeate solute concentrations.
2. Evaluate alternatives for membrane equipment and operation, select the type of membrane element, and determine the array configuration (number of stages, number of passes, number of elements in a pressure vessel, number of vessels in each stage, feed pressure).
3. Select feed water pretreatment requirements (methods to control fouling).
4. Select permeate posttreatment requirements.
5. Select concentrate management and disposal requirements.
6. Select ancillary membrane system features such as permeate back-pressure control and interstage booster pumps.
7. Select equipment and procedures for process monitoring.

These elements of design are not independent of one another. For instance, recovery is often constrained by the solubility of limiting salts. As a result, selection of pretreatment requirements, recovery, and array design must be done simultaneously and iteratively to determine the most economical design.

The basis for design information typically includes characteristics of the feed water (solute concentrations, turbidity, SDI and MFI values) from laboratory or historical data, required treated-water quality (established by the client or regulatory limits), and required treated-water capacity (established by demand requirements). The process design criteria for a hypothetical brackish water RO facility are shown in Table 17-4. Frequently, pilot testing is part of the design process.

The following discussion focuses primarily on the design of the membrane components of an RO system. Design of additional components, such as intakes and pretreatment systems, are available in design manuals such as AWWA (2007).

Element Selection and Membrane Array Design

Membrane array design involves determination of the quantity and quality of water produced by each membrane element in an array. This involves calculation of the flow, velocity, applied pressure, osmotic pressure, water flux, and solute flux in each element, which leads to the determination of the number of stages, number of passes, number of elements in each pressure vessel, and number of vessels in each stage. Membrane array design is a complex and iterative process using a large number of interrelated design parameters. Several important design parameters such as mass transfer coefficients are specific to individual products and available only from membrane manufacturers. Because of the complexity of the calculations and dependence on manufacturer information, array design is often done with design software provided by membrane manufacturers. Nevertheless, an understanding of the mechanics of the design procedure as described in the following paragraphs is important to interpreting the results from manufacturer design software.

DESIGN CALCULATIONS

The most common type of membrane element in use is the spiral-wound element. As described earlier, feed water enters one end of the pressure vessel and flows through several spiral-wound elements in series. As the water passes through each element, some water passes through the membrane into the permeate carrier channel, resulting in continuously changing conditions along the length of the membrane element. The net transmembrane pressure declines continuously across the length of a membrane element because of changes in both applied pressure (due to head loss in the feed channels) and osmotic pressure (due to concentration of salts). As a result, fluxes of both water and solute are dependent on the position

Table 17-4

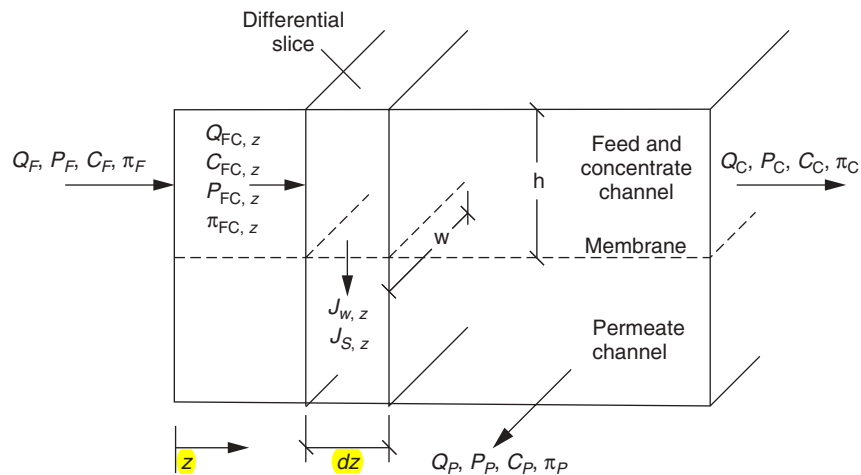
Design criteria for a hypothetical reverse osmosis facility

Operating Parameter	Units	Value
Feedwater pretreatment		
Capacity	m ³ /d	37,900
Strainers		
Number	Number	5
Nominal particle size rating	μm	5
Capacity, each	m ³ /d	9,480
Chemicals		
Sulfuric acid, max. dose	mg/L	200
Scale inhibitor, max. dose	mg/L	2
Feed pumps		
Number	Number	5
Capacity, each	m ³ /d	9,480
Pressure	bar	40
Membrane system		
Feed water flow rate	m ³ /d	37,900
Permeate flow rate	m ³ /d	30,300
Concentrate flow rate	m ³ /d	7,580
Recovery	%	80
Number of arrays	Number	4
Capacity per array	m ³ /d	9,480
Array design criteria		
Membrane area per element	m ²	32.5
Elements per pressure vessel	Number	7
Number of stages per array	Number	2
Number of pressure vessels (stage 1)	Number	40
Stage 1 avg. permeate flux	L/m ² · h	21
Number of pressure vessels (stage 2)	Number	20
Stage 2 avg. permeate flux	L/m ² · h	17
Posttreatment ^a		
Caustic soda, max. dose	mg/L	10
Corrosion inhibitor, max. dose	mg/L	1
Chlorine, max. dose	mg/L	2
Fluoride, max. dose	mg/L	1
Concentrate disposal	Deep-well injection	

^aPosttreatment may also include a countercurrent packed tower for hydrogen sulfide or carbon dioxide removal. See Chap. 14 for details of packed-tower design.

within a spiral-wound element, and the design procedure must integrate along the length of the membrane element.

A differential slice of a membrane element is shown on Fig. 17-17. In this figure, the center plane represents the membrane surface, with the feed–concentrate channel above the membrane and the permeate channel

**Figure 17-17**

Differential slice of spiral-wound membrane element. Because the feed flows axially along the pressure vessel and the permeate flows spirally toward the center of the vessel, the feed and permeate flows are perpendicular to each other.

below the membrane. The fluxes of water and solute are described by Eqs. 17-9 and 17-10, but the applied pressure differential, osmotic pressure differential, and concentration differential depend on the location within the pressure vessel:

$$J_{W,Z} = k_W(\Delta P_Z - \Delta \pi_Z) = k_W[(P_{FC,Z} - P_{P,Z}) - (\pi_{M,Z} - \pi_{P,Z})] \quad (17-48)$$

$$J_{S,Z} = k_S(\Delta C_Z) = k_S(C_{M,Z} - C_{P,Z}) \quad (17-49)$$

where $C_{M,Z}$ = concentration at the membrane surface,

$$C_{M,Z} = \beta_Z C_{FC,Z}, \text{ mol/L}$$

$\pi_{M,Z}$ = osmotic pressure at the membrane surface, bar

Other terms are defined on Fig. 17-17.

The water and solute mass transfer coefficients (k_W and k_S) are dependent on the properties and configurations of specific membrane elements and cannot be generalized. These values are embedded in the manufacturer's design software and are typically not publicized but can be generated from pilot data if they cannot be obtained from the manufacturer.

Solute flux calculations are complicated by the presence of multiple solutes, which may have different value for the mass transfer coefficient. For instance, a low-pressure NF membrane has low rejection of monovalent ions but high rejection of divalent ions, and the mass transfer coefficients would reflect this difference in rejection.

The permeate flow and mass solute flow through the membrane are equal to the flux times the membrane area in the differential element, and the cumulative transfer of water and solute across the membrane is

determined by integrating the flow between the feed end and the position z within the pressure vessel, as shown in the following:

$$Q_{P,Z} = \int_0^z J_{W,Z} w \, dz \quad (17-50)$$

$$M_{S,Z} = \int_0^z J_{S,Z} w \, dz \quad (17-51)$$

where w = effective width of feed–concentrate flow channel, m
 $M_{S,Z}$ = mass of solute transferred, mg/s

The water flow rate (and velocity) in the feed–concentrate channel declines as permeate is produced, and the flow rate at any point in the channel can be determined by subtracting the net permeate production up to that point from the feed water flow rate as follows:

$$Q_{FC,Z} = Q_F - Q_{P,Z} \quad (17-52)$$

Similarly, the solute concentration in the feed–concentrate channel can be determined by performing a mass balance on the solute as follows:

$$C_{FC,Z} = \frac{Q_F C_F - M_{S,Z}}{Q_{FC,Z}} \quad (17-53)$$

Water and solute flux are affected by concentration polarization and the concentration of solute at the membrane surface. Some manufacturers have developed relationships describing concentration polarization for specific element designs, and these relationships should be used if available. If no manufacturer information is available, the correlations presented earlier in this chapter can be used to estimate the concentration polarization factor. Because both flux and velocity are changing, β must be calculated using Eq. 17-41, but as a function of position, as shown in the equation

$$\beta_Z = \text{Rej}(e^{J_{W,Z}/k_{CP,Z}}) + (1 - \text{Rej}) \quad (17-54)$$

The mass transfer coefficient k_{CP} depends on velocity in the feed–concentrate channel, which can be calculated from the expression

$$v_Z = \frac{Q_{FC,Z}}{hw} \quad (17-55)$$

where h = height of feed–concentrate channel, m

The solute concentration at the membrane surface is defined by Eq. 17-40, using concentrations as a function of position.

$$C_{M,Z} = \beta_Z C_{FC,Z} \quad (17-56)$$

Pressure in the feed channel drops due to head loss, but head loss is not constant across the length of the membrane element. Turbulent

conditions are maintained, so head loss in the channel is proportional to the square of the velocity and the first power of length (consistent with the Darcy–Weisbach equation) as given by the expression

$$h_L = \delta_{HL} v^2 L \quad (17-57)$$

where h_L = head loss in feed–concentrate channel, bar
 δ_{HL} = head loss coefficient, bar · s²/m³
 v = water velocity in feed–concentrate channel, m/s
 L = channel length, m

Finally, the permeate solute concentration can be calculated from the ratio of the solute and water fluxes per Eq. 17-11:

$$C_{P,Z} = \frac{J_{S,Z}}{J_{W,Z}} \quad (17-58)$$

Additional design calculations, such as the calculation of osmotic pressure from concentration, have been presented earlier in this chapter. The use of these equations in system array design is demonstrated in Example 17-5.

Example 17-5 Calculation of permeate flux and concentration

Calculate the quantity and quality of water produced by a single membrane element (permeate concentration, rejection, and recovery) given the following information:

Parameter	Unit	Value
Membrane properties		
Element length	m	1
Element membrane area	m ²	32.5
Effective feed channel height	mm	0.125
Water mass transfer coefficient (k_W)	L/m ² · h · bar	2.87
Solute mass transfer coefficient (k_S)	m/h	6.14×10^{-4}
Element head loss (at design velocity of 0.5 m/s)	bar	0.2
Operating conditions		
Feed flow (Q_F)	m ³ /d	270
Feed pressure (P_F)	bar	14.2
Feed concentration (C_F)	mg/L NaCl	2000
Feed temperature (T_F)	°C	20
Permeate pressure (P_P)	bar	0.3
Osmotic coefficient (ϕ)		1.0

Assume $D_{NaCl} = 1.35 \times 10^{-9}$ m²/s, $\phi = 1$, and $MW_{NaCl} = 58.4$.

Solution

The basic solution strategy is to (1) divide the membrane element into a number of increments; (2) determine P , v , C , and π on both sides of the membrane in the first increment; (3) calculate the water and solute flux across the membrane in the first increment; (4) determine Q , P , C , v , and π on both sides of the membrane in the next increment; (5) calculate the water and solute flux across the membrane in the next increment; and (6) repeat steps 4 and 5 for all remaining increments.

Part 1

Divide the element into 10 increments 0.1 m length each. Determine v , P , C , and π on both sides of the membrane in the first increment. The subscript FC is used to designate the feed-concentrate side of the membrane, and the subscript P designates the permeate side of the membrane.

1. The following values are given in the problem statement:

$$Q_{FC,Z} = Q_F = 270 \text{ m}^3/\text{d}$$

$$P_{FC,Z} = P_F = 14.2 \text{ bar}$$

$$P_{P,Z} = 0.3 \text{ bar}$$

2. The feed channel velocity is determined by dividing the feed flow by the channel cross-sectional area. The effective channel height is given as 0.125 mm, but the width is not given. The width can be determined by dividing the membrane area by the element length, both of which are readily available information:

$$w = \frac{a}{L} = \frac{32.5 \text{ m}^2}{1 \text{ m}} = 32.5 \text{ m}$$

It should be noted that the element is not 32.5 m wide. Spiral-wound elements are typically 0.2 to 0.3 m in diameter, and 32.5 m is the unit width of the membrane surface (which includes multiple feed channels because multiple envelopes are used, see Sec. 17-4) as wrapped around the permeate tube. Then,

$$Q_{FC,Z} = \frac{270 \text{ m}^3/\text{d}}{86,400 \text{ s/d}} = 3.125 \times 10^{-3} \text{ m}^3/\text{s}$$

$$V_Z = \frac{Q_{FC,Z}}{hw} = \frac{(3.125 \times 10^{-3} \text{ m}^3/\text{s})(10^3 \text{ mm/m})}{(0.125 \text{ mm})(32.5 \text{ m})}$$

$$= 0.769 \text{ m/s}$$

3. Calculate the osmotic pressure in the feed channel using Eq. 17-7:

$$\pi_{FC,Z} = \frac{(2 \text{ mol ion/mol NaCl})(1.0)(2000 \text{ mg/L})(0.0831451 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(293 \text{ K})}{(10^3 \text{ mg/g})(58.4 \text{ g/mol})}$$

$$= 1.67 \text{ bar}$$

4. The water and solute fluxes depend on the concentration and osmotic pressure in the permeate, which of course depend on the water and solute fluxes. Although a simultaneous numerical solution procedure could be used, it is acceptable to assume C_P and π_P are zero in the first increment for this example. Values calculated in the first increment will be used as an approximation of the values in the next increment.

Part 2

Calculate the water and solute flux and flow rate across the membrane in the first increment.

1. The concentration and osmotic pressure at the membrane wall are higher than in the feed channel because of concentration polarization. However, the concentration polarization factor is dependent on permeate flux, so values for the concentration polarization factor and permeate flux must be determined concurrently by simultaneously solving Eqs. 17-48 and 17-54.

a. Calculate the Reynolds number, Schmidt number, and k_{CP} using Eqs. 17-36, 17-37, and 17-35. The hydraulic diameter is $2h = 2 \times (0.125 \text{ mm}) = 0.25 \text{ mm}$. Water density and viscosity at 20°C are $\rho_W = 998 \text{ kg/m}^3$ and $\mu_W = 10^{-3} \text{ kg/m} \cdot \text{s}$ (Table C-1, App. C):

$$\text{Re} = \frac{\rho v d_H}{\mu} = \frac{(998 \text{ kg/m}^3)(0.769 \text{ m/s})(0.25 \text{ mm})}{(1.0 \times 10^{-3} \text{ kg/m} \cdot \text{s})(10^3 \text{ mm/m})} = 192$$

$$\text{Sc} = \frac{\mu}{\rho D_L} = \frac{1.0 \times 10^{-3} \text{ kg/m} \cdot \text{s}}{(998 \text{ kg/m}^3)(1.35 \times 10^{-9} \text{ m}^2/\text{s})} = 742$$

$$k_{CP} = \frac{(0.023)(1.35 \times 10^{-9} \text{ m}^2/\text{s})(192)^{0.83}(742)^{0.33}}{(0.25 \text{ mm})(10^{-3} \text{ m/mm})}$$

$$= 8.64 \times 10^{-5} \text{ m/s}$$

b. The parameter β can be calculated using Eq. 17-41. Re_j is not yet known and is assumed to be 1.0 in the first increment.

In subsequent increments, Re_j will be taken as equal to the value calculated in the previous increment:

$$\begin{aligned}\beta_Z &= \exp\left(\frac{J_{W,Z}}{k_{CP}}\right) Re_j + (1 - Re_j) \\ &= \exp\left[\frac{(J_{W,Z} \text{ L/m}^2 \cdot \text{h})(10^{-3} \text{ m}^3/\text{L})}{(8.64 \times 10^{-5} \text{ m/s})(3600 \text{ s/h})}\right] \quad (\text{a})\end{aligned}$$

- c. The osmotic pressures in the feed water and at the membrane surface are related by β_Z :

$$C_{M,Z} = \beta_Z C_{FC,Z}$$

Therefore

$$\pi_{M,Z} = \beta_Z \pi_{FC,Z} \quad (\text{b})$$

- d. Substituting Eq. (b) into Eq. 17-48 yields

$$J_{W,Z} = k_W[(P_{FC,Z} - P_{P,Z}) - (\beta_Z \pi_{FC,Z} - \pi_{P,Z})] \quad (\text{c})$$

- e. Solving Eqs. (a) and (c) simultaneously using values given in the problem statement yields $\beta_Z = 1.12$ and $J_{W,Z} = 35.1 \text{ L/m}^2 \cdot \text{h}$.
2. The permeate flow rate is calculated by multiplying the flux by the area of the increment:

$$\begin{aligned}Q_{P,Z} &= J_{W,Z}(w)(dz) = \frac{(35.1 \text{ L/m}^2 \cdot \text{h})(32.5 \text{ m})(0.1 \text{ m})}{(10^3 \text{ L/m}^3)(3600 \text{ s/h})} \\ &= 3.17 \times 10^{-5} \text{ m}^3/\text{s}\end{aligned}$$

3. The solute flux can be calculated using Eq. 17-49 after substituting in Eq. (b) (see step 1c above):

$$J_{S,Z} = k_S(\beta_Z C_{FC,Z} - C_{P,Z})$$

$$J_{S,Z} = (6.14 \times 10^{-4} \text{ m/h})[(1.12)(2000 \text{ mg/L}) - 0 \text{ mg/L}](10^3 \text{ L/m}^3)$$

$$J_{S,Z} = 1375 \text{ mg/m}^2 \cdot \text{h}$$

4. Calculate the solute transport across the membrane:

$$M_{S,Z} = J_{S,Z}(w)(dz) = \frac{(1375 \text{ mg/m}^2 \cdot \text{h})(32.5 \text{ m})(0.1 \text{ m})}{3600 \text{ s/h}} = 1.24 \text{ mg/s}$$

Part 3

Determine P , C , and π on both sides of the membrane in the next increment along with v in the feed channel.

1. The flow in the feed channel is equal to the influent flow minus any permeate production and is calculated using Eq. 17-52:

$$\begin{aligned} Q_{FC,Z} &= Q_F - Q_{P,Z} = 3.125 \times 10^{-3} \text{ m}^3/\text{s} - 3.17 \times 10^{-5} \text{ m}^3/\text{s} \\ &= 3.09 \times 10^{-3} \text{ m}^3/\text{s} \end{aligned}$$

2. Calculate feed channel velocity:

$$v_Z = \frac{Q_{FC,Z}}{hw} = \frac{(3.09 \times 10^{-3} \text{ m}^3/\text{s})(10^3 \text{ mm/m})}{(0.125 \text{ mm})(32.5 \text{ m})} = 0.761 \text{ m/s}$$

3. The solute concentration in the feed channel of the next increment can be calculated using Eq. 17-53:

$$\begin{aligned} C_{FC,Z} &= \frac{Q_F C_F - M_{S,Z}}{Q_{FC,Z}} \\ &= \frac{[(3.125 \times 10^{-3} \text{ m}^3/\text{s})(2000 \text{ mg/L})(10^3 \text{ L/m}^3)](-1.24 \text{ mg/s})}{(3.09 \times 10^{-3} \text{ m}^3/\text{s})(10^3 \text{ L/m}^3)} \\ &= 2020 \text{ mg/L} \end{aligned}$$

4. The solute concentration in the permeate of the next increment can be calculated from the water and solute fluxes in the first increment using Eq. 17-58:

$$C_{P,Z} = \frac{J_{S,Z}}{J_{W,Z}} = \frac{1371 \text{ mg/m}^2 \cdot \text{h}}{35.1 \text{ L/m}^2 \cdot \text{h}} = 39.2 \text{ mg/L}$$

5. Calculate the feed channel and permeate osmotic pressures using Eq. 17-7:

$$\pi_{FC,Z} = \frac{(2 \text{ mol ion/mol NaCl})(1.0)(2020 \text{ mg/L})(0.0831451 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(293 \text{ K})}{(10^3 \text{ mg/g})(58.4 \text{ g/mol})}$$

$$= 1.68 \text{ bar}$$

$$\pi_{P,Z} = \frac{(2 \text{ mol ion/mol NaCl})(1.0)(39.2 \text{ mg/L})(0.0831451 \text{ L} \cdot \text{bar/K} \cdot \text{mol})(293 \text{ K})}{(10^3 \text{ mg/g})(58.4 \text{ g/mol})}$$

$$= 0.03 \text{ bar}$$

6. The pressure in the feed channel drops due to head loss through the channel, and the head loss is a function of the feed velocity. The head loss in the first increment and pressure in the next increment can be calculated:

- a. The head loss in an incremental length of the membrane element as a function of velocity must be determined from the given head

loss information using Eq. 17-57 rearranged as follows:

$$\delta_{HL} = \frac{h_L}{v^2 L} = \frac{0.2 \text{ bar}}{(0.5 \text{ m/s})^2 (1 \text{ m})} = 0.8 \text{ bar} \cdot \text{s}^2/\text{m}^3$$

b. Determine the head loss in the increment using Eq. 17-57:

$$h_{L,Z} = \delta_{HL} v_Z^2 dz = (0.8 \text{ bar} \cdot \text{s}^2/\text{m}^3)(0.769 \text{ m/s})^2(0.1 \text{ m}) = 0.047 \text{ bar}$$

c. Determine pressure in the next increment:

$$P_{FC,Z} = 14.2 \text{ bar} - 0.047 \text{ bar} = 14.15 \text{ bar}$$

Part 4

Repeat Parts 2 and 3 for the second and subsequent increments. The results are shown in the table below:

Increment (z)	Unit	1	2	3	4	5	...	10
$Q_{FC,Z}$	m ³ /s	3.125×10^{-3}	3.093×10^{-3}	3.062×10^{-3}	3.030×10^{-3}	2.999×10^{-3}		2.845×10^{-3}
v_Z	m/s	0.7692	0.7614	0.7536	0.7459	0.7382		0.7003
$P_{FC,Z}$	bar	14.20	14.15	14.11	14.06	14.02		13.81
$h_{L,Z}$	bar	0.047	0.046	0.045	0.045	0.044		0.039
$C_{FC,Z}$	mg/L	2000	2020	2041	2062	2084		2196
$\pi_{FC,Z}$	bar	1.67	1.68	1.70	1.72	1.74		1.83
$Q_{P,Z}$	m ³ /s	3.17×10^{-5}	3.16×10^{-5}	3.14×10^{-5}	3.13×10^{-5}	3.11×10^{-5}		3.03×10^{-5}
$P_{P,Z}$	bar	0.3	0.3	0.3	0.3	0.3		0.3
$C_{P,Z}$	mg/L	0	39.2	39.0	39.6	40.3		43.6
$\pi_{P,Z}$	bar	0	0.03	0.03	0.03	0.03		0.04
$k_{CP,Z}$	m/s	8.64×10^{-5}	8.56×10^{-5}	8.49×10^{-5}	8.42×10^{-5}	8.35×10^{-5}		7.99×10^{-5}
β_Z		1.120	1.120	1.121	1.121	1.122		1.124
$J_{W,Z}$	L/m ² · h	35.1	35.0	34.8	34.7	34.5		33.6
$J_{S,Z}$	mg/m ² · h	1374.77	1365.46	1380.43	1395.16	1410.15		1489.01
M_Z	mg/s	1.24	1.23	1.25	1.26	1.27		1.34
Re_{jz}		0.980	0.981	0.981	0.981	0.981		0.981

Part 5

After calculating Part 4 for all increments in the element, the overall performance can be determined.

1. Permeate production from the element is the sum of the permeate produced in each increment:

$$Q_P = \sum_{Z=1}^{10} Q_{P,Z} = 3.1 \times 10^{-4} \text{ m}^3/\text{s}$$

2. Salt transfer from the element is the sum of the salt transferred in each increment:

$$M_S = \sum_{Z=1}^{10} M_{S,Z} = 12.8 \text{ mg/s}$$

3. Permeate concentration:

$$C_P = \frac{M_S}{Q_P} = \frac{12.8 \text{ mg/s}}{(3.10 \times 10^{-4} \text{ m}^3/\text{s})(10^3 \text{ L/m}^3)} = 41.3 \text{ mg/L}$$

4. Rejection (Eq. 17-1):

$$\text{Rej} = 1 - \frac{C_P}{C_F} = 1 - \frac{41.3 \text{ mg/L}}{2000 \text{ mg/L}} = 0.98$$

5. Recovery (Eq. 17-12):

$$r = \frac{Q_P}{Q_F} = \frac{3.1 \times 10^{-4} \text{ m}^3/\text{s}}{3.12 \times 10^{-3} \text{ m}^3/\text{s}} = 0.099$$

Comment

In this example, the performance of a single membrane element has been determined. The concentrate from this element becomes the feed to the next element in series; that is, $Q_{C,1}$, $P_{C,1}$, and $C_{C,1}$ are $Q_{F,2}$, $P_{F,2}$, and $C_{F,2}$. The system permeate flow rate is the sum of the permeate flow from each element. The system permeate concentration is the flow-averaged permeate concentration from each element.

MANUFACTURER SOFTWARE

In Example 17-5 pressure was used as an input variable and a value for recovery was generated. Normally, the desired recovery is determined from limiting salt calculations (taking acid and antiscalant addition into account), and design calculations generate the feed pressure required for a particular membrane element. Using these equations, an iterative solution would be necessary. The design calculations are also repeated with varying membrane elements and array configurations. In addition, other process parameters, such as permeate backpressure and interstage booster pumps, can affect system design and performance. Thus, design is an iterative process and typically takes place with the cooperation of several membrane system manufacturers. Manufacturers provide design software to perform these calculations, which are based on the principles presented in this chapter, and incorporate issues such as osmotic pressure, limiting salt solubility, mass transfer rates, concentration polarization, and permeate water quality. As such, manufacturers' software is reliable for predicting effluent water

quality from a specific membrane system design and a given set of operating conditions. An example of the output from a vendor-supplied RO design program is shown in Table 17-5.

FUNCTIONAL SPECIFICATIONS

Because design criteria cannot be developed independently of manufacturer data, procurement of RO systems is often accomplished by means of a functional specification. By this method, an engineer develops the system requirements, designs the pretreatment processes, designs the RO system support facilities, and defines the basic requirements of the RO system. The functional specifications outline the operating requirements of the system, physical constraints of the system, and warranty agreements between the manufacturer and the owner. Bid proposals are returned by the interested manufacturers that outline the particulars of the system being supplied, estimates of system product quality as a function of time, system capital costs, and system operating costs as a function of time. The proposals are typically reviewed by the engineer to determine the optimum life-cycle cost.

An important aspect of long-term RO operation is loss of performance due to compaction, fouling, or degradation of the membrane. Limiting salt calculations can be a good predictor of the recovery that can be achieved without causing scaling. Antiscalants can allow supersaturation (i.e., higher recovery) without scaling, but their effectiveness might be dependent on other water quality parameters. SDI and MFI tests can indicate when feed water quality is unacceptable, but low values do not assure that fouling will be minimal. Therefore, it is necessary to perform pilot testing for nearly all RO installations. Pilot testing is guided by membrane system selection and operating conditions developed during array design and serves to verify the array design criteria and identify pretreatment requirements to prevent excessive fouling.

Pilot Testing

COMMERCIAL RO PILOT PLANTS

Reverse osmosis pilot plant systems are typically available from membrane manufacturers or consulting engineering firms. A typical commercially available skid-mount system is shown on Fig. 17-18. This skid unit contains six pressure vessels, each containing spiral-wound membrane elements in series. The pressure vessels can be operated as two independent systems, with each system containing three pressure vessels that can be piped as a 2×1 array, which allows membranes from two manufacturers to be tested simultaneously. The pilot plant system is operated with a programmable logic controller (PLC). Chemicals are added to the feed water to prevent fouling of the membrane. Manufacturer-supplied specifications for pilot plant systems are usually provided so that the pilot unit can be properly operated. These specifications are usually obtained from the manufacturer

Table 17-5Example output from vendor-supplied RO design program^a

Hydranautics Membrane System Design Software, v. 8.00 © 2002 3/11/03										
RO program licensed to: K Howe										
Calculation created by: K Howe										
Project name: MWH Example										
HP pump flow:	4666.7 gpm	Permeate flow:	3500.0 gpm							
Recommended pump press:	204.4 psi	Raw-water flow:	4666.7 gpm							
Feed pressure:	175.4 psi	Booster pump pressure:	10.0 psi							
Feed water temperature:	15.0°C (59°F)	Permeate recovery ratio:	75.0%							
Raw water pH:	8.00	Element age:	5.0 years							
Acid dosage, ppm (100%):	131.1 H ₂ SO ₄	Flux decline % per year:	7.0							
Acidified feed CO ₂ :	127.3	Salt passage increase, %/yr:	10.0							
Average flux rate:	15.8 gfd	Feed type:	Well water							
Stage	Perm. Flow, gpm	Flow/Vessel		Flux, gfd	Beta	Concentration and Throt. Pressures		Element Type	Element No.	Element Array
		Feed, gpm	Conc, gpm			psi	psi			
1-1	2623.6	53.0	23.2	17.9	1.16	149.5	0.0	ESPA3	528	88 × 6
1-2	876.4	45.4	25.9	11.7	1.08	133.1	0.0	ESPA3	270	45 × 6
Ion	Raw water		Feed water		Permeate		Concentrate			
	mg/L	CaCO ₃	mg/L	CaCO ₃	mg/L	CaCO ₃	mg/L	CaCO ₃		
Ca	8.0	20.0	8.0	20.0	0.27	0.7	31.2	77.7		
Mg	2.0	8.2	2.0	8.2	0.07	0.3	7.8	32.1		
Na	734.3	1596.3	734.3	1596.3	115.11	250.2	2591.9	5634.5		
K	8.0	10.3	8.0	10.3	1.52	2.0	27.4	35.2		
NH ₄	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0		
Ba	0.004	0.0	0.004	0.0	0.000	0.0	0.016	0.0		
Sr	2.000	2.3	2.000	2.3	0.069	0.1	7.794	8.9		
CO ₃	3.0	5.0	0.2	0.4	0.00	0.0	0.8	1.4		
HCO ₃	631.0	517.2	473.5	388.1	174.26	142.8	1371.3	1124.0		
SO ₄	79.0	82.3	207.5	216.1	7.41	7.7	807.7	841.3		
Cl	730.0	1029.6	730.0	1029.6	72.28	101.9	2703.2	3812.6		
F	1.1	2.9	1.1	2.9	0.28	0.7	3.6	9.4		
NO ₃	0.0	0.0	0.0	0.0	0.00	0.0	0.0	0.0		
SiO ₂	24.0			24.0	5.83		78.5			
TDS	2222.4		2190.6		377.1		7631.2			
pH	8.0		6.8		6.4		7.3			
				Raw Water	Feed Water	Concentrate				
CaSO ₄ /K _{sp} × 100:				0%	0%	2%				
SrSO ₄ /K _{sp} × 100:				2%	5%	29%				
BaSO ₄ /K _{sp} × 100:				7%	17%	97%				
SiO ₂ saturation:				20%	20%	65%				
Langelier saturation index (LSI)				-0.14	-1.47	0.04				
Stiff-Davis saturation index				-0.20	-1.53	-0.24				
Ionic strength				0.03	0.04	0.13				
Osmotic pressure				22.2 psi	21.3 psi	74.2 psi				

Table 17-5 (Continued)

Stage	Element No.	Feed	Pressure	Permeate	Permeate	Concentrate							
		Element Pressure, psi	Drop, psi	Flow, gpm	Flux, gfd	Beta	Permeate TDS	Osmotic Pressure	Concentrate Saturation Level, %		LSI		
								CaSO ₄	SrSO ₄	BaSO ₄	SiO ₂		
1-1	1	175.4	6.5	5.7	20.5	1.11	116.6	23.8	1	6	20	22	-0.9
1-1	2	168.9	5.5	5.4	19.4	1.12	126.5	26.7	1	7	23	25	-0.7
1-1	3	163.4	4.6	5.1	18.3	1.12	137.8	30.2	1	8	27	28	-0.6
1-1	4	158.8	3.8	4.8	17.2	1.13	151.0	34.4	2	9	32	32	-0.4
1-1	5	155.0	3.1	4.5	16.1	1.15	166.2	39.6	2	11	38	36	-0.3
1-1	6	151.8	2.5	4.1	14.9	1.16	203.0	45.9	2	14	47	42	-0.1
1-2	1	156.3	5.4	4.1	14.6	1.09	225.4	49.8	3	16	52	45	0.0
1-2	2	150.9	4.7	3.7	13.4	1.09	251.4	54.0	3	18	59	49	0.1
1-2	3	146.3	4.1	3.4	12.2	1.09	279.6	58.5	3	20	66	53	0.1
1-2	4	142.1	3.6	3.1	11.1	1.09	309.1	63.2	4	22	74	56	0.2
1-2	5	138.5	3.2	2.8	10.0	1.09	341.4	68.2	4	25	84	60	0.3
1-2	6	135.4	2.8	2.5	8.9	1.08	374.9	73.3	5	28	94	64	0.3

^aThese calculations are based on nominal element performance when operated on a feed water of acceptable quality. No guarantee of system performance is expressed or implied unless provided in writing by Hydranautics.

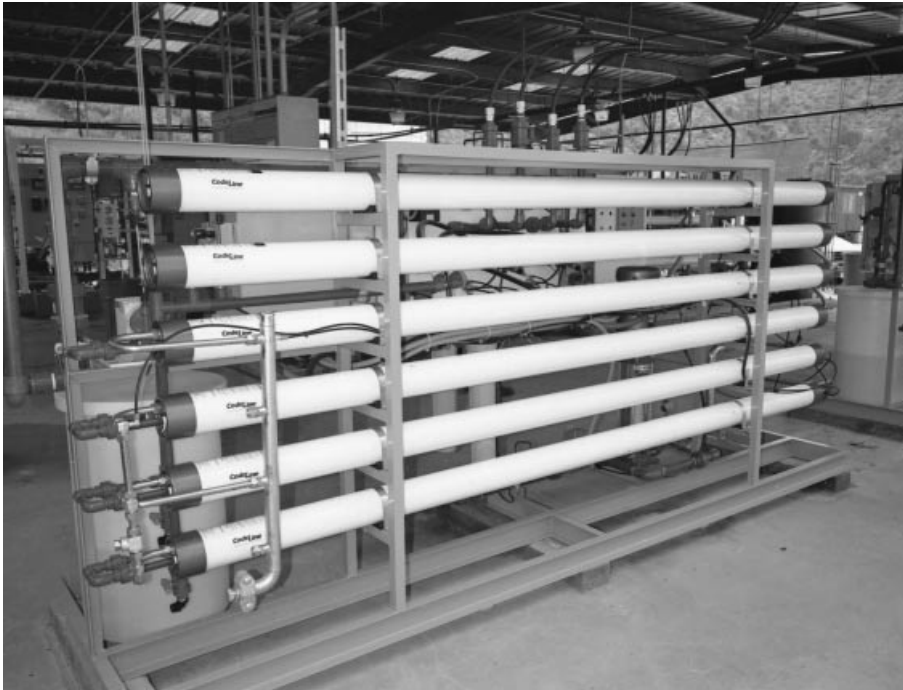


Figure 17-18
Typical reverse osmosis pilot plant.

and provide useful guidelines when planning and operating the pilot plant units.

PILOT TEST PARAMETERS

For most RO pilot studies, the following parameters should be recorded:

1. Date and time of sample analysis
2. Flow rates (feed, concentrate, and permeate)
3. Pressure (feed, concentrate, and permeate)
4. Feed water temperature
5. Conductivity (online reading recommended)
6. Power consumption
7. Chemical usage
8. pH (feed, concentrate, and permeate)

Additional reporting and recording requirements are available elsewhere (ASTM, 2001c, 2001d).

Pretreatment

Pretreatment is necessary to prevent scaling and fouling. The common pretreatment strategies include the injection of acids and antiscalants to prevent the precipitation of sparingly soluble salts and filtration to prevent plugging by particulate matter. Very clean source water (such as groundwater) often can operate with only cartridge filtration prior to the membrane units, but more advanced filtration methods, including coagulation, flocculation, sedimentation, and granular filtration, or membrane filtration, are commonly required with surface water intake facilities. Pretreatment must be selected and designed in concert with the array design because the membrane element performance is dependent on the level of pretreatment. Additional details on the design of pretreatment systems is available in design manuals such as AWWA (2007).

Posttreatment

The permeate from an RO facility typically requires additional treatment. Feed water pH adjustment prior to RO, along with extensive removal of divalent ions by the RO process, produces treated water with low pH, low alkalinity, and low hardness, which are conditions that cause water to be corrosive. Anaerobic groundwater frequently contains hydrogen sulfide, which passes through the membrane and causes odor problems in the treated water. Finally, residual disinfection is always required for municipal water distribution.

PERMEATE STABILITY

A number of strategies can be used to increase the stability (reduce the corrosivity) of the water. When the feed water is acidified for scale control, carbonate alkalinity in the raw water is converted to carbonic acid, which passes through the membrane. Thus, addition of a base such as caustic soda can restore both pH and alkalinity to acceptable levels. Without additional measures, however, such water will still be corrosive. Stability can be improved by adding hardness ions to the water, so base addition with chemicals containing calcium is sometimes preferred over caustic soda. Lime and soda ash are common chemicals for increasing the stability of RO permeate. Small systems sometimes can add an acceptable amount of hardness by passing the permeate through a bed of calcareous media such as dolomite or calcite. In lieu of adding hardness to the water, corrosion inhibitors may be effective. Another strategy for producing a stable finished water is to blend the permeate with a bypass stream of raw water that meets all other water treatment requirements (such as filtration if a surface water source is used). Proper blending of raw and permeate water may produce a finished water with the desired pH, alkalinity, and hardness. However, DBP precursor concentration in the raw water and the potential for DBP formation need to be evaluated when considering blending options. The importance of finished-water stability is discussed in additional detail in Chap. 22.

HYDROGEN SULFIDE

Anaerobic groundwater can contain hydrogen sulfide, a highly odorous compound that is not removed during RO. Hydrogen sulfide can be removed by oxidation or aeration. Oxidation to sulfate can be accomplished with oxidants such as chlorine, but large doses are needed (the stoichiometric chlorine requirement is about 9 times the hydrogen sulfide concentration on a mass basis and insufficient amounts can oxidize sulfide to elemental sulfur, which is equally undesirable). Thus, hydrogen sulfide is commonly removed after the membrane process in an air-stripping process using countercurrent packed towers, which are discussed in Chap. 14. Since hydrogen sulfide is a weak acid, the pH of the water will have a significant impact on its removal efficiency (Howe and Lawler, 1989). Odor control can be a significant issue when stripping water that contains sulfide.

It is necessary to consider all posttreatment goals simultaneously and select treatment options that achieve all objectives. For instance, air stripping to remove sulfide before base addition will strip carbon dioxide and increase the permeate pH; subsequent pH adjustment with caustic soda will not restore alkalinity because the carbonate will be gone. Alternatively, pH adjustment before stripping can prevent effective stripping because

sulfide is present as ionic hydrogen bisulfide rather than gaseous hydrogen sulfide.

DISINFECTION

Chlorine is commonly used for disinfection and is discussed in Chap. 13. The RO process is effective at removing DBP precursors; thus, free chlorination can typically be practiced without forming significant quantities of DBPs. However, care must be used if the RO permeate is to be blended with either raw water (for stability, see above) or a fresh water supply. Blending may increase DBP formation when using free chlorine. Cases have been observed when the blending of desalinated seawater into freshwater can increase the DBP formation of the freshwater, even though the desalinated water has a very low DBP formation potential on its own. Desalinated seawater can have a higher bromide concentration than freshwater sources, so that interactions between bromide from the desalinated seawater and NOM from the freshwater can increase overall DBP formation after chlorination to above what it would be with either water source individually. Thus, bromide removal can be one of the factors that controls the design of RO facilities.

Concentrate Management

A significant concern in the design and operation of inland brackish water RO facilities is the low product water recovery compared to other water treatment processes. Recovery is limited by osmotic pressure in seawater systems and by scaling from sparingly soluble salts in inland brackish water systems. For inland systems, the low recovery has two negative consequences. First, brackish water desalination is typically considered because of a lack of adequate freshwater resources, and inability to recover a high fraction of the feed water is simply a poor use of scarce natural resources. Second, the unrecovered water becomes the concentrate stream and must be disposed of. The increase salinity of the concentrate stream greatly limits available disposal options because of the potential for contaminating the scarce freshwater resources. Thus, there is significant interest in increasing recovery of product water and decreasing the volume of concentrate that must be disposed of.

Increasing recovery from inland brackish water RO facilities involves preventing the precipitation of sparingly soluble salts. As noted earlier, scale inhibitors are used to prevent precipitation and increase recovery up to a point. However, scale inhibitors are limited in their effectiveness, and more aggressive strategies typically must be employed to achieve recovery of greater than 90 percent.

One strategy is to provide an intermediate treatment process between two stages of RO membranes. Since calcium is often the limiting cation,

lime softening can be an effective intermediate strategy. Softening can also be effective at removing other scale-causing constituents. Gabelich et al. (2007) found that increasing pH to between 10.5 and 11.5 with NaOH was able to remove 88 to 98 percent of Ca^{2+} , Ba^{2+} , Sr^{2+} , and 67 percent of silica. However, the high alkalinity and hardness present after a first stage of RO can lead to high doses of lime or NaOH; doses in excess of 1000 mg/L have been reported in experimental studies. Similarly high doses of acid can be necessary to reduce the pH after softening. The high doses also lead to a large amount of waste production. Seeding with calcite or gypsum crystals has also been explored as a way of improving the effectiveness of the intermediate precipitation process (Rahardianto et al., 2007). Fluidized bed crystallization using sand as a seed material has also proved effective in bench-scale testing (Sethi et al., 2008). Ion exchange is another possibility for interstage treatment for the removal of scale-causing constituents that may result in less waste production (Howe et al., 2010).

Several patented or proprietary processes have been developed to increase recovery from brackish RO systems. The patented high-efficiency reverse osmosis (HERO) process involves pretreatment to reducing scaling, followed by pH adjustment and additional stages of reverse osmosis. Hardness is typically removed using a cation exchange column that removes calcium and magnesium, and carbonate is removed by stripping carbon dioxide in a countercurrent packed column (see Chap. 14). The pH is then increased using caustic soda, typically above $\text{pH} = 10$. Since calcium and carbonate have been removed, calcium carbonate scaling at high pH is no longer a concern and the concentrate is fed into another stage of reverse osmosis. At pH above 10, silica and borate are transformed from neutral to ionic species, the solubility of silica is increased and scaling potential is reduced, the rejection of silica and borate is increased, the potential for organic fouling or biofouling is decreased, and cleaning costs are reduced. Recovery of 90 to 98 percent has been achieved.

Another proprietary system is the SAL-PROC system developed by Geo-Processors, Inc. This process uses a variety of treatment steps, including chemical addition, heating, cooling, and sequential concentration steps that may include more RO or evaporation. The SAL-PROC system is potentially capable of producing usable and possibly sellable salt products and slurries from the RO concentrate.

Another option that has been explored in research to prevent scaling and potentially increase recovery include the vibratory shear-enhanced process (VSEP) in which a membrane system is vibrated to prevent scale from forming on the membrane surface (Chang, 2008). Researchers have also explored other electrodialysis reversal, membrane distillation, or other desalination processes as a second-state desalination system after an intermediate-scale reduction process (Sethi et al., 2008).

Brine concentrators and crystallizers are additional technologies to reduce the volume of concentrate, and can lead to zero liquid discharge (ZLD), in which the only residuals from the facility are solids, which are then easier to dispose of (Mickley, 2006). While brine concentrators and crystallizers are used in some industrial processes such as the power generation industry, they are expensive, energy intensive, and have not yet been used in municipal water treatment industry. Brine concentrators and crystallizers are discussed in more detail in Chap. 21.

Disposal of Residuals

Disposal of the concentrate stream is frequently a challenge in RO plant design. The factors that contribute to this problem are identified in Table 17-6. In addition to the concentrate stream, RO plants must also dispose of spent cleaning solutions. Both of these residuals are discussed in this section.

CONCENTRATE

Several surveys of concentrate disposal methods are available (Kenna and Zander, 2001; Mickley et al., 1993; Truesdall et al., 1995). The most common concentrate disposal options in the United States are (1) discharge to a brackish surface water (include oceans, brackish rivers, or estuaries),

Table 17-6
Factors affecting concentrate disposal

Issue	Description
Volume	The waste stream volume from many water treatment processes is less than 5% of the feed stream volume. In RO, the waste stream volume ranges from 15 to 50% of the feed stream volume.
Salinity/toxicity	The high salinity of the concentrate stream makes it toxic to many plants and animals, limiting options for land application or surface water discharge and rendering it unusable for recycling or reuse. Many concentrate streams are anaerobic, which can be toxic to fish without sufficient dilution. In addition, RO processes used for specific contaminant removal (i.e., arsenic, radium) may produce concentrate streams that can be classified as a hazardous material.
Regulations	Concentrate is classified as an industrial waste by the U.S. EPA. Concentrate disposal is regulated under several different federal, state, and local laws, and the interaction between these regulatory requirements can be complex (Kimes, 1995; Pontius et al., 1996). Regulatory considerations are often as important as cost and technical considerations for determining viable concentrate disposal options.

(2) discharge to a municipal sewer, and (3) deep-well injection. In the United States, about half of all plants discharge concentrate to a surface water, a third discharge to a municipal sewer, and about 10 percent discharge to a deep well. Deep-well injection is most common in Florida. Evaporation ponds are used by a small number of facilities. Concentrate disposal is an integral part of the design of RO facilities and disposal options are discussed in more detail in Chap. 21 of this text.

An alternative to disposal of concentrate is to identify beneficial uses for the concentrate or its constituent salts and minerals. Possible beneficial uses that have been explored in various research projects include (1) land application or irrigation of salt-tolerant crops, (2) saline aquaculture, farming of brine shrimp or other saltwater species, (3) restoration of brackish waterways or development of saltwater marshes, wetlands, or habitats, (4) energy generation using solar gradient ponds, (5) industrial uses as feedstock or process stream, (6) production of marketable salts or mineral commodities (Ahuja and Howe, 2005; Everest and Murphree, 1995). At the current time, however, beneficial uses for the concentrate have not been identified at most facilities.

CLEANING SOLUTIONS

Spent cleaning solutions from RO plants are frequently acidic or basic solutions and contain detergents or surfactants. In many cases, the cleaning solution volume is small compared to the concentrate stream and can be diluted into and disposed of with the concentrate. In some cases, treatment of the cleaning solution may be required prior to disposal, but treatment may consist only of pH neutralization. Detergents and surfactants should be selected with disposal issues in mind.

Reverse osmosis is an energy-intensive process. The theoretical thermodynamic minimum energy requirement for desalinating seawater, based solely on the pressure required to overcome the osmotic pressure, is 0.70 kWh/m^3 . This value is significantly higher than the typical energy required for the treatment of freshwater. A significant component of operating costs is electrical power for the feed pumps because of the high pressure necessary to operate RO membranes. Although pressure drops significantly as permeate passes through the membrane, the head loss through the feed channels is relatively small, and the concentrate exits the final membrane element at 80 to 90 percent of the feed pressure, with backpressure maintained by a concentrate control valve. If concentrate is discharged to a deep well, a portion of this pressure can be used to drive the disposal process. If, however, the concentrate is discharged to a surface water, this pressure must be dissipated prior to discharge. Pressure in the concentrate stream

Energy Recovery

dissipated across the concentrate control valve is wasted energy because it performs no useful work in the treatment system. Because the concentrate stream is both high energy and relatively high volume, the amount of wasted energy is substantial.

Energy recovery devices are being used more frequently to reclaim the wasted energy in the concentrate stream. Several types of devices are available, including reverse-running turbines, Pelton wheels, pressure exchangers, and electric motor drives (Geisler et al., 1999; Harris, 1999; Oklejas and Pergande, 2000; Tomkins and Nemeth, 2001). Typically, recovered energy from the residual pressure of the concentrate stream is used to pressurize the feed stream. In some systems, the concentrate stream spins a rotor, losing energy in the process, and exits the energy recovery device at a significantly lower pressure. In the reverse-running turbine and pressure exchanger, the energy recovery device is in contact with both the feed and concentrate streams, with a single rotor transferring pressure from the concentrate to the feed stream. Pressure exchangers allow direct contact between the feed and concentrate streams via a rotating rotor, and are thus able to transfer the pressure from the concentrate stream directly to the feed stream. Pelton wheel devices use a rotor connected directly to the feed pump via an extended shaft, and the energy recovered from the concentrate stream provides hydraulic assistance to the operation of the feed pumps. The main moving part is the Pelton wheel and shaft. Electric motor drives are more complex, utilizing a hydraulic drive system connected to the pump motor.

More than 90 percent of the energy expended to pressurize the concentrate stream can be recovered. Depending on the price for electricity, capital costs of energy recovery equipment may be recouped within 3 to 5 years. Energy recovery devices were first utilized on seawater RO systems because they operate at high pressure and low recovery, compounding the energy loss. Recent trends and improvements in energy recovery equipment and rising electricity prices suggest that energy recovery will be applied in more and more low-pressure systems.

In addition to providing pressure to the feed stream, another application is to use the energy recovery system to add pressure between stages (Duranceau et al., 1999). In normal operation, the second or later stages produce less permeate because of lower applied pressure (due to pressure drop in the first stage) and higher osmotic pressure (due to concentration of the feed stream in the first stage). The lower permeate flow and higher feed concentration also increase salt passage and degrade permeate quality. These effects are sometimes counteracted by installing booster pumps between stages, so that a higher feed pressure is available to offset

the higher osmotic pressure. By using energy recovery devices to boost pressure between stages, the booster pumps can be eliminated, which offsets a portion of the capital cost of the energy control device.

Problems and Discussion Topics

- 17-1 Discuss key similarities and differences between membrane filtration and RO.
- 17-2 Explain why dissolved gases such as CO_2 and H_2S are poorly rejected by RO membranes.
- 17-3 Calculate the total osmotic pressure of seawater at a temperature of 20°C using the ion concentrations shown in Table 17-2 and $\phi = 1$. Calculate the osmotic pressure of a solution containing an equivalent concentration of sodium chloride (i.e., 35,200 mg/L NaCl) also using $\phi = 1$. Explain and discuss the difference between the two results and discuss Fig. 17-9 in the context of these results.
- 17-4 The following solutions are representative of common applications of reverse osmosis. Calculate the osmotic pressure of each at 20°C . Discuss the importance of osmotic pressure and how it affects the applied pressure for these applications.
- NaCl = 35,000 mg/L (representative of seawater RO).
 - NaCl = 8000 mg/L (representative of brackish water RO).
 - Hardness = 400 mg/L as CaCO_3 (representative of softening NF).
 - Dissolved organic carbon (DOC) = 25 mg/L (representative of using NF to control DBP formation by removing DBP precursors. Assume an average MW of 1000 g/mol.).
- 17-5 Seawater RO facilities are restricted to a maximum applied pressure of about 85 bar (1200 psi) because of equipment limitations. Using the seawater composition shown in Table 17-2, calculate the maximum recovery that can be achieved before the osmotic pressure at the membrane surface (at the exit from a membrane module) is equal to the applied pressure. Assume 100 percent rejection, a temperature of 15°C , and a concentration polarization factor of 1.12. Discuss how the results of this calculation compare to the typical recovery achieved by seawater RO facilities. Does osmotic pressure lead to any practical limitations on the size of the waste stream from a seawater RO facility?

- 17-6 Operating data for a low-pressure RO system on two different days are shown in the table below:

	Unit	Day 1	Day 2
Water temperature	°C	13	22
Water flux	L/m ² · h	17.5	18.8
Feed pressure	bar	41.9	38.7
Concentrate pressure	bar	39.0	35.8
Permeate pressure	bar	0.25	0.25
Feed TDS concentration	mg/L	10,500	10,200
Permeate TDS concentration	mg/L	120	120
Recovery	%	66	68

Performance data for this membrane element were developed using the following standard conditions:

	Unit	Standard
Temperature	°C	20
Feed pressure	bar	40
Permeate pressure	bar	0
Head loss per element	bar	0.4
Number of elements	no.	7
Feed TDS concentration	mg/L	10,000
Permeate TDS concentration	mg/L	100
Recovery	%	70

Determine the difference in system performance (water flux and rejection) between the two days using the temperature correction formula in this text and an arithmetic average for the solute concentration in the feed–concentrate channel. Assume the salts in the feed water are sodium chloride for the purpose of calculating osmotic pressures.

- 17-7 In Eq. 17-10 the solute flux is dependent on the concentration gradient and independent of pressure; also it was noted that solute flux is dependent on temperature. However, Eq. 17-26 includes a correction factor for pressure and not temperature, from which it appears that rejection is dependent on pressure and independent of temperature. Show mathematically and explain (1) how rejection can be dependent on pressure when solute flux is independent of pressure and (2) why there is no temperature correction factor for rejection when there is a temperature correction factor for water flux.
- 17-8 Examine the importance of the diffusion coefficient on concentration polarization by plotting β as a function of the diffusion coefficient for diffusion coefficient values between 10^{-10} m²/s

(typical of NOM with a diameter of 5 nm) and $1.35 \times 10^{-9} \text{ m}^2/\text{s}$ (sodium chloride). Use feed channel velocity 0.65 m/s, permeate flux $25 \text{ L}/\text{m}^2 \cdot \text{h}$, hydraulic diameter 0.5 mm, and temperature 20°C . Discuss the implications that this graph has on the accumulation of material at the membrane surface.

- 17-9 Examine the importance of temperature on concentration polarization by plotting β as a function of temperature for values between 1 and 30°C . Use feed channel velocity 0.65 m/s, permeate flux $25 \text{ L}/\text{m}^2 \cdot \text{h}$, hydraulic diameter 0.5 mm, and calculate the diffusion coefficient from the Nernst–Haskell equation given in Chap. 7 (Eq. 7-36) for sodium chloride. Discuss how temperature will impact water and solute flux across the membrane from the perspective of concentration polarization.
- 17-10 An SDI test was performed to evaluate the fouling tendency of potential RO source water. The time to collect 500 mL of water was measured as 24 s. Filtration continued for a total of 15 min, and then a second 500 mL was collected. The time necessary to collect the second 500-mL sample was 32 s. Calculate the SDI.
- 17-11 Calculate the MFI from the following experimental data:

Time, min	Volume Filtered, L	Time, min	Volume Filtered, L	Time, min	Volume Filtered, L
0	0	5.5	5.37	11.0	9.86
0.5	0.63	6.0	5.80	11.5	10.24
1.0	1.17	6.5	6.23	12.0	10.61
1.5	1.68	7.0	6.65	12.5	10.98
2.0	2.16	7.5	7.07	13.0	11.35
2.5	2.64	8.0	7.48	13.5	11.71
3.0	3.11	8.5	7.89	14.0	12.06
3.5	3.58	9.0	8.29	14.5	12.41
4.0	4.03	9.5	8.69	15.0	12.75
4.5	4.48	10.0	9.08		
5.0	4.93	10.5	9.47		

- 17-12 An RO facility is being designed to treat groundwater containing the ions given below. Calculate the allowable recovery before scaling occurs and identify the limiting salt. Assume 100 percent rejection, a concentration polarization factor of 1.08, and $T = 25^\circ\text{C}$, and ignore the impact of ionic strength. The water contains calcium = 105 mg/L, strontium = 2.5 mg/L, barium = 0.0018 mg/L, sulfate = 128 mg/L, fluoride = 1.3 mg/L, and silica = 9.1 mg/L as Si.
- 17-13 A groundwater has a calcium concentration of 125 mg/L, alkalinity of 180 mg/L as CaCO_3 , and pH of 7.1. Calculate the degree of

- supersaturation of calcium carbonate (ratio of actual concentration to the saturated concentration for each ion) at 60 percent recovery. Calculate the adjusted pH value and acid (HCl) dose necessary to prevent calcium carbonate precipitation at this recovery. Assume 100 percent rejection, $\beta = 1.12$, and $T = 25^\circ\text{C}$, and ignore ionic strength.
- 17-14 Feed water to a proposed low-pressure RO facility has a barium concentration of $0.2\ \mu\text{g/L}$ and a sulfate concentration of $420\ \text{mg/L}$. The planned recovery is 80 percent. Calculate the concentration polarization allowable before the solubility of barium sulfate is exceeded. Assume 100 percent rejection and $T = 25^\circ\text{C}$, and ignore the impact of ionic strength.
- 17-15 Reverse osmosis facilities can be designed with multiple stages (concentrate from one stage is fed to the next stage) or multiple passes (permeate from one stage is fed to the next stage). Explain the difference in permeate quantity and quality expected from these systems.
- 17-16 Concentrate-staged membrane arrays can be designed with a booster pump in the concentrate line between stages. Explain the benefits of this interstage booster pump and the impact it has on permeate quantity and quality.
- 17-17 Design criteria for an RO system are given in the following table:

Item	Unit	Value
Membrane properties		
Element length	m	1
Element membrane area	m^2	32.5
Feed channel height (spacer thickness)	mm	0.125
Water mass transfer coefficient (k_W)	$\text{L}/\text{m}^2 \cdot \text{h} \cdot \text{bar}$	1.25
Solute mass transfer coefficient (k_S)	m/h	3.29×10^{-4}
Element head loss (at design velocity of 0.5 m/s)	bar	0.1
Operating conditions		
Feed flow (Q_F)	m^3/d	19,000
Feed pressure (P_F)	bar	34
Feed concentration (C_F)	mg/L NaCl	8500
Feed temperature (T_F)	$^\circ\text{C}$	20
Permeate pressure (P_P)	bar	0.3

The system is to be designed as a 2×1 array with 80 pressure vessels in the first stage and 40 pressure vessels in the second stage, and with 7 membrane elements in each pressure vessel.

- a. Using a spreadsheet or computer program, calculate and graph (1) the feed flow rate entering each element, (2) the feed concentration entering each element, (3) the concentration

- polarization factor β at each element, (4) the permeate flow rate produced by each element, and (5) permeate salt concentration produced by each element. For the purposes of this problem, assume that the operating conditions are constant across the length of each individual element. Assume that the feed water salinity is due entirely to NaCl, $\phi = 0.94$, and $D_{\text{NaCl}} = 0.8 \times 10^{-9} \text{ m}^2/\text{s}$ (from Table 7-4 in Chap. 7).
- Calculate the average permeate flow rate and concentration for each stage and for the whole array.
 - Calculate overall recovery, rejection, and average water flux.
 - Discuss any observations about the quantity and quality of water produced by the first element compared to the last element, and explain the observed trend in β .
- 17-18 Calculate and plot water flux and salt rejection as a function of recovery, for recovery ranging from 50 to 85 percent, given $C_F = 10,000 \text{ mg/L NaCl}$, $\Delta P = 50 \text{ bar}$, $k_W = 2.2 \text{ L/m}^2 \cdot \text{h} \cdot \text{bar}$ and $k_S = 0.75 \text{ L/m}^2 \cdot \text{h}$, $\phi = 1$, and $T = 20^\circ\text{C}$. Comment on the effect of recovery on RO performance.
- 17-19 A new brackish water RO system is being proposed. The water quality is as shown in the table below. Using RO manufacturer design software (provided by the instructor or obtained from a membrane manufacturer website), develop the process design criteria for the plant. The required water demand is $38,000 \text{ m}^3/\text{d}$ and the finished-water TDS should be 500 mg/L or lower.

Constituent	Concentration, mg/L	Constituent	Concentration, mg/L
Ammonia	1.3	Bicarbonate	680
Barium	0.04	Chloride	890
Calcium	20	Fluoride	0.7
Iron	0.5	Orthophosphate	0.7
Magnesium	2.5	Sulfate	105
Manganese	0.02	Silica	21.5
Potassium	17	Nitrate	1.2
Sodium	875	Hydrogen sulfide	0.3
Strontium	2.17		
pH	7.8	Turbidity	0.3 NTU
SDI	$<1 \text{ min}^{-1}$	Temperature	15°C

- 17-20 A new seawater RO system is being proposed. The water quality is as shown in the table below. Using RO manufacturer design software (provided by the instructor or obtained from a membrane manufacturer website), develop the process design criteria for

the plant. The required water demand is 4000 m³/d and the finished-water TDS should be 500 mg/L or lower.

Constituent	Concentration, mg/L	Constituent	Concentration, mg/L
Aluminum	0.15	Strontium	6.6
Ammonia	0.092	Bromide	51
Barium	0.00	Bicarbonate	112
Boron	4.3	Chloride	18,900
Calcium	439	Fluoride	0.61
Iron	0.1	Phosphate	0.12
Magnesium	1,240	Sulfate	2380
Potassium	425	Silica	0.86
Sodium	10,100	Hydrogen sulfide	0.0
Strontium	6.6		
pH	8.0	Turbidity	3.3 NTU
SDI	<1 min ⁻¹	UV ₂₅₄	0.03/cm
Temperature	15°C		

17-21 A new membrane softening system is being proposed. The water quality is as shown in the table below. Using RO manufacturer design software (provided by the instructor or obtained from a membrane manufacturer website), develop the process design criteria for the plant. The required water demand is 14,200 m³/d and the finished-water hardness should be between 50 and 75 mg/L as CaCO₃.

Constituent	Concentration, mg/L	Constituent	Concentration, mg/L
Ammonia	1.5	Bicarbonate	135.1
Barium	0.0	Bromide	0.0
Calcium	100	Carbonate	0.11
Magnesium	10	Chloride	98.8
Manganese	0.002	Fluoride	0.5
Sodium	60	Phosphate	0.5
Strontium	1.0	Sulfate	167.6
		Silica	15.0
pH	7.0	Temperature	20°C
SDI	<1 min ⁻¹		

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