The Guidebook to
Membrane Desalination Technology
Reverse Osmosis, Nanofiltration and Hybrid Systems
Process, Design and Applications
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Mark Wilf

With chapters by Leon Awerbuch, Graeme Pearce, Craig Bartels, Mike Mickley, and Nikolay Voutchkov

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Introduction

Mark Wilf* Chapters 1–15

The reverse osmosis (RO) technology is at present the most versatile desalination method. It is effective in treating water of any salinity, starting with low salinity brackish water up to high salinity seawater. RO units are applied to produce ultrapure water for the semiconductor industry and potable water from high salinity seawater in a cost effective way.

The RO technology started as a scientific experiment at the University of Florida in the 1950s where Reid and Breton (1) were able to demonstrate desalination properties of a cellulose acetate film. At that time reverse osmosis was just one of a number of desalination processes that were evaluated for commercial use (2). After the development of the first asymmetric membrane material from cellulose acetate by Loeb and Sourirajan in the 1950s (3, 8), the subsequent progress included the development of composite membranes with better performing and more robust polyamide based membrane chemistry (4). This was followed by development and optimization of membrane module configurations and the improvement of the process and RO system design. Within the next five decades of continuous development, inventions and improvements, the RO process has been transferred from a scientific curiosity into a self-supporting, rapidly growing industry.

The scientific experiment of the 1950s, which produced a few drops of desalted water per hour, results today (16, 19) in a worldwide network of RO plants of a combined desalting capacity of about 20 million cubic meters per

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day (5.2 billion gallons per day) out of a total of 40 million m$^3$/d. Reverse osmosis technology is used today in large municipal plants to produce potable water quality from brackish and seawater sources, reclaim municipal effluents and reduce water salinity for industrial applications. At the other end of the application spectrum reverse osmosis membrane elements are used in small under the sink units to produce a few liters per day of drinking water.

A wide variety of membrane material chemistry and membrane module configurations have been developed over the years. However, current commercial membrane modules are almost exclusively made of composite aromatic polyamide membrane materials in a spiral wound configuration. A small fraction of RO elements are still made of cellulose acetate polymers in spiral wound and capillary configuration.

Significant efforts have been made to reduce the investment and operating cost of desalination systems. Better understanding of feed water quality requirements and the introduction of organic scale inhibitors resulted in simplification of the feed water pretreatment process and the increase of feed to permeate conversion ratio.

The development of high flux membrane elements and incorporation of variable speed drivers and power recovery equipment into RO system configuration resulted in significant reductions of specific power consumption. One of the remaining unresolved problems of RO technology is effective control of biofouling in RO systems equipped with polyamide membrane elements. This is especially important in applications involving treatment of feed water from surface or wastewater sources.

The initial cellulose acetate membrane manufactured in the late sixties had a specific permeate flux of about 0.32 l/m$^2$/h/bar (0.013 gfd/psi) of net driving pressure and salt transport coefficient of 1.5E-05 cm/s. The early RO membranes required a net driving pressure of over 70 bar (1000 psi) in order to produce a permeate flux rate of 25.5 l/m$^2$/h (15 gfd). The latest generation of polyamide brackish water membranes has specific flux of 5.9 l/m$^2$/h/bar (0.24 gfd/psi) and salt transport coefficient below 1.0E-05 cm/s. The corresponding net driving pressure required to produce a flux rate of 25.5 (15 gfd) is only 4.3 bar (62 psi) with higher salt rejection (about 35% lower salt passage) than the initial CA membranes. This improvement of specific permeate flux translates into over a twenty fold reduction of the specific power consumption of the RO process pumps. The evolution of membrane performance is summarized in Table 1 and Figs. 1-1 and 1-2.
FIG. 1.1 Evolution of performance of brackish RO membranes.

FIG. 1.2 Evolution of performance of seawater RO membranes.
A comprehensive review of early development work and theory of the RO process is included in a book edited by Merten (6). An early review of RO applications and RO systems design is included in a publication edited by Buros (5).

The interesting point is that the basic composition of aromatic polyamide membrane has not changed since its invention in the 1980s. The improvements of performance are the result of better morphology of the composite membrane structure. They were achieved mainly through better control of the manufacturing parameters due to improved understanding of the membrane formation process.

<table>
<thead>
<tr>
<th></th>
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<th></th>
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<th></th>
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</tr>
</thead>
<tbody>
<tr>
<td>Membrane type</td>
<td>CA asymmetric</td>
<td>PA composite</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Specific flux, l/m²/h/bar (gfd/psi)</td>
<td>0.32 (0.013)</td>
<td>0.76 (0.031)</td>
<td>1.24 (0.05)</td>
<td>1.73 (0.07)</td>
<td>2.96 (0.12)</td>
<td>5.93 (0.24)</td>
<td>8.15 (0.31)</td>
</tr>
<tr>
<td>Salt rejection, %</td>
<td>96.0</td>
<td>98.0</td>
<td>98.0</td>
<td>98.5</td>
<td>99.5</td>
<td>99.0</td>
<td>99.2</td>
</tr>
<tr>
<td>Salt transport, cm/s</td>
<td>1.5E-5</td>
<td>1.2E-5</td>
<td>1.9E-5</td>
<td>2.0E-5</td>
<td>5.8E-6</td>
<td>6.0E-6</td>
<td></td>
</tr>
<tr>
<td><strong>NDP</strong> required for flux 26l/m²/h (15 gfd), bar (psi)</td>
<td>79.3 (1150)</td>
<td>33.3 (483)</td>
<td>20.7 (300)</td>
<td>14.8 (214)</td>
<td>8.6 (125)</td>
<td>4.3 (62.5)</td>
<td>3.1 (45.5)</td>
</tr>
</tbody>
</table>

CA asymmetric: Cellulose acetate membrane material, asymmetric structure of membrane layer.
PA composite: Polyamide membrane material, composite structure. Ultra thin membrane barrier cast on a porous support.
Introduction to reverse osmosis: basic terms

2.1. Water salinity—concentration units

The concentration of dissolved ions in water is expressed as parts per million (ppm), which is equivalent to milligrams per liter (mg/l) or grams per cubic meter (g/m³). Sometimes concentration is expressed as a percent of weight of dissolved ions in the total weight of solution. Approximately 10,000 ppm is equal to 1% weight concentration. Concentration can also be expressed as milliequivalents per liter (meq/l). Miliequivalent concentration is calculated by dividing concentration, expressed as ppm, by equivalent weight, which is ion weight divided by valency. For example calcium (Ca) ion concentration of 1000 ppm is equivalent to meq concentration of \( \frac{1000}{(40/2)} = 50 \) meq/l. Natural water sources contain dissolved ions at various concentrations ranging from a few hundreds ppm for low salinity well or surface water to 35,000–45,000 ppm for a seawater source.

Direct determination of total dissolved solids (TDS) concentration is quite cumbersome. It requires evaporation to dryness of accurately weighted volume of water solution and weight determination of the dry residue. The most common approach to measurement of concentration of total dissolved solids in solution is through analytical determination of concentration of dissolved ions and summation of individual ion concentrations. Approximate determination of the concentration of dissolved ions can be accomplished by measurement of electric conductivity of the water solutions according to the equation:

\[
\text{TDS} = K \cdot \text{EC}
\]  

(1)
where TDS is concentration in parts per million (ppm), K is a conversion factor and EC is electric conductivity in μS/cm.

The composite conductivity of ions in solution is easily measurable. It is customarily expressed as micro Siemens per cm (μS/cm). Electric conductivity of solution is affected by temperature and ion composition. Conductivity varies about 2% per degree C and is customarily reported as a corrected value at 25°C. Different ions have different contributions to conductivity. Therefore conductivity conversion to TDS has to be calibrated with different TDS determinations for each site. The common values of conversion factor K are in the range of 0.55 for an RO permeate up to 0.75 for seawater concentrate (7).

2.2. The osmotic process

Osmosis is a natural process involving fluid flows across a semipermeable membrane barrier. It is selective in the sense that the solvent passes through the membrane at a faster rate than the passage of dissolved solids. The difference of passage rate results in solvent-solids separation. The direction of solvent flow is determined by its chemical potential which is a function of pressure, temperature, and concentration of dissolved solids. Pure water in contact with both sides of an ideal semipermeable membrane at equal pressure and temperature has no net flow across the membrane because the chemical potential is equal on both sides. If a soluble salt is added to water on one side of the membrane, the chemical potential of this salt solution is reduced. Osmotic flow from the pure water side across the membrane to the salt solution side will occur until the equilibrium of chemical potential is restored (Figure 2.1a). Equilibrium occurs when the hydrostatic pressure differential resulting from the volume changes on both sides is equal to the osmotic pressure. This is a solution property independent of the membrane. Application of an external pressure to the salt solution side, which is equal to the osmotic pressure, will also cause equilibrium. Additional pressure will raise the chemical potential of the water in the salt solution and cause a solvent flow to the pure water side, because it now has a lower chemical potential. This phenomenon is called reverse osmosis (Figure 2.1b).

The osmotic pressure, \( P_{\text{osm}} \), of a solution can be determined indirectly by measuring the concentration of dissolved salts in solution:

\[
P_{\text{osm}} = R(T + 273) \sum (m_i)
\]  

\[ (2) \]
where $P_{\text{osm}}$ is osmotic pressure (in bar), $R$ is the universal gas constant ($0.082 \text{ L} \cdot \text{atm/mol} \degree\text{K}$), $T$ is the temperature (in $\degree\text{C}$), and $\Sigma(m_i)$ is the sum of molar concentration of all constituents in a solution.

An approximation for $P_{\text{osm}}$ can be made by assuming that 1000 ppm concentration of TDS equals about 0.77 bar (11 psi) of osmotic pressure. For example, in an RO unit operating at 75% recovery rate, feed salinity is 3,000 ppm TDS and concentrate salinity is about 11,500 ppm TDS. Accordingly, osmotic pressure of the feed is 2.3 bar (33 psi) and the concentrate is 8.7 bar (126 psi). Eq. 2 only holds for dilute salt solutions and temperatures close to 25°C. At significantly different conditions a more rigorous calculation that takes into consideration ion activities rather than concentrations has to be applied.

### 2.3. Permeate recovery rate (conversion)

Permeate recovery is one of the more important parameters in the design and operation of RO systems. Recovery or conversion rate of feed water to product (permeate) is defined by Eqs. 3 and 4 and illustrated in Fig. 2.2:

$$R_p = 100\% \left( \frac{Q_p}{Q_f} \right)$$

$$R_p = 100\% \left( \frac{Q_p}{Q_p + Q_c} \right)$$

where $R_p$ is the product recovery rate (in %), $Q_p$ is the product water flow rate, $Q_f$ is the feed water flow rate and $Q_c$ is the concentrate flow rate.

---

**FIG. 2.1** The direct and reverse osmosis process.
The recovery rate affects salt passage and product flow. As the recovery rate increases, the salt concentration on the feed-brine side of the membrane increases, which causes an increase in salt flow rate across the membrane. Also, a higher salt concentration in the feed-brine solution increases the osmotic pressure, reducing the net driving pressure available and consequently reducing the product water flow rate.

Using Eq. 4 and mass balance a relation can be developed between recovery and concentration of various streams:

\[
R_p = \frac{Q_p}{Q_p + Q_c}
\]

\[
Q_f C_f = Q_p C_p + Q_c C_c
\]

\[
Q_f C_f = Q_p C_p + ((Q_f - Q_p)C_c)
\]

\[
Q_f(C_f - C_c) = Q_p(C_p - C_c)
\]

\[
R_p = \frac{Q_p}{Q_f} = \frac{(C_c - C_f)}{(C_c - C_p)}
\]

where \( C_f \) is feed concentration, \( C_p \) is permeate concentration and \( C_c \) is concentrate concentration.

The above relations (Eq. 5) can be applied to determine recovery rate from concentration values of ions in the feed, permeate and concentrate stream. Usually these calculations are based on concentrations of chloride or calcium ions which can be determined easily and with a high degree of accuracy.

**Example 1**

- Designed recovery rate 75%
- Concentrations, ppm of Cl⁻
  - Feed = 1000

**FIG. 2.2** Mass balance in reverse osmosis unit. Recovery rate 75%.
Concentrate = 3800
Permeate = 200
Actual recovery rate, \( R_p = \frac{3800 - 1000}{3800 - 200} = 0.78 \) (78%)

In a multistage RO system the product recovery rate is defined for each stage and for a combined system (Fig. 2.3).

**Example 2**

Designed recovery rate 0.75 (75%)
Feed concentration, ppm Cl\(^-\) = 1000
Permeate concentration, ppm Cl\(^-\) = 200
Calculation of concentrate concentration:
\[
C_c = \frac{(C_f - R_p C_p)}{(1 - R_p)}
\]
\[
C_c = \frac{(1000 - 0.75 \times 200)}{(1 - 0.75)} = 3400
\]

**FIG. 2.3** Recovery rate in a two stage RO unit.

*Stage 1* \( R_{p1} = 65\% \)
*Stage 2* \( R_{p2} = 57\% \) (20%)
Total \( R_t = R_{p1} + (1 - R_{p1}/100) \cdot R_{p2} \)
\[
R_t = 65\% + (1 - (65/100)) \cdot 57\% = 85\%
\]

### 2.4. Average feed salinity

Average feed salinity (AFS) is a representative value of feed concentration used for calculation of performance of membrane element or RO system. AFS accounts for the phenomenon of salinity increase in the RO system from the salinity of feed water at the entrance to the RO device to the final salinity of concentrate leaving the system. The AFS is calculated as an arithmetic (Eq. 6) or logarithmic mean (Eq. 7). AFS salinity can be expressed as a function of recovery rate, assuming at first approximation that ions are totally rejected by RO membranes:
AFS = 0.5 \( C_f (1 + 1/(1 - R)) \) \( \quad \) (6)
AFS = \( C_f \ln \left(1/(1 - R)\right)/R \) \( \quad \) (7)

where \( R \) = recovery rate expressed as a decimal fraction.

The arithmetic mean is usually applied for calculations in cases of low recovery (single element calculations). In case of a high recovery rate a logarithmic mean is used for performance calculations. The values of concentration factors for the arithmetic average, logarithmic average and for concentrate are shown in Fig. 2.4.

![Fig. 2.4 Concentration increase vs. recovery rate.](image)

2.5. Net driving pressure

The net driving pressure (NDP) is the driving force of the water transport through the semipermeable membrane. The value of NDP decreases along the RO unit. Therefore, for the purpose of membrane performance calculations it is defined as an average NDP. The NDP is defined as the fraction of the applied pressure in excess of average osmotic pressure of the feed and any pressure losses in the system according to Eq. 8.

\[
NDP = P_f - P_{os} - P_p - 0.5P_d (+ P_{osp}) \quad (8)
\]

where

- \( P_f \) = feed pressure
- \( P_{os} \) = average feed osmotic pressure
- \( P_p \) = permeate pressure
- \( P_d \) = pressure drop across RO elements
- \( P_{osp} \) = osmotic pressure of permeate.
In regular RO applications osmotic pressure of permeate is negligible. However, in NF applications, where salt rejection is relatively low, permeate salinity is significant compared to the feed concentration. Therefore, osmotic pressure of permeate has to be considered in calculation of NDP in NF systems.

**Example 3**

Brackish RO system operates at recovery rate of 85%

Feed salinity: 2000 ppm TDS
Permeate salinity = 25 ppm TDS
Concentrate salinity = \((2000 – 0.85 \times 25)/(1 – 0.85) = 13191\) ppm
Feed pressure: 13 bar (188.5 psi)
Pressure drop across the system: 3.0 bar (45 psi)
Permeate pressure: 1.0 bar (14.5 psi)
Average feed salinity = 0.5 \((2000 + 13191) = 7596\) ppm
Average feed osmotic pressure = 0.77 \((7596/1000) = 5.9\) bar (85.5 psi)
Permeate osmotic pressure = 0.77 \((25/1000) = 0.02\) bar (0.3 psi)
Average net driving pressure: NDP = 13 – 5.9 – 1.0 – 0.5 \(\times 3.0 = 4.6\) bar (67 psi)

**Example 4**

Nanofiltration RO system operates at recovery rate of 85%

Feed salinity: 1000 ppm TDS
Permeate salinity = 350 ppm TDS
Concentrate salinity = \((1000 – 0.85 \times 350)/(1 – 0.85) = 4683\) ppm
Feed pressure: 7 bar (101 psi)
Pressure drop across the system: 3.0 bar (45 psi)
Permeate pressure: 1.0 bar (14.5 psi)
Average feed salinity = 0.5 \((1000 + 4683) = 2842\) ppm
Average feed osmotic pressure = 0.77 \((2842/1000) = 2.2\) bar (32 psi)
Permeate osmotic pressure = 0.77 \((350/1000) = 0.3\) bar (4 psi)
Average net driving pressure: NDP = 7 – 2.2 – 1.0 – 0.5 \(\times 3.0 + 0.3 = 2.6\) bar (38 psi)

2.6. Water reversed salt separation by reverse osmosis

The mechanism of water and salt separation by reverse osmosis is not fully understood. Current scientific thinking suggests two water transport models: porosity and diffusion. That is, transport of water through the membrane may
be through physical pores present in the membrane (porosity), or by diffusion from one bonding site to another within the membrane. The theory suggests that the chemical nature of the membrane is such that it will absorb and pass water preferentially to dissolved salts at the solid/liquid interface. This may occur by weak chemical bonding of the water to the membrane surface or by dissolution of the water within the membrane structure. Either way, during this process a salt concentration gradient is formed across the solid/liquid interface (Fig. 2.7).

The chemical and physical nature of the membrane determines its ability to allow for preferential transport of solvent (water) over solute (salt ions). It is also known that part of dissolved species rejection is a result of size discrimination, i.e., larger molecules are better rejected by the RO membranes than the small ones. Another part of the rejection process is a result of the repulsion of dissolved ions due to presence of charges (usually negative) on the membrane surface. The membrane repels ions having the same charges as those present on the membrane surface. Usually ions of multiple negative valency (high ion charges) are better rejected than single valency ions. For example, rejection of sulfate ions is higher than rejection of chloride ions. Due to the condition of maintaining electro-neutrality of solutions on both sides of the membrane, repulsion of one type of ion hinders passage of the co-ion and increases overall rejection. Conversely, presence of ions with high passage through the membrane will increase passage of the corresponding co-ion. The extent of this charge-dependant rejection will vary with the composition of solution treated (see also chapter 10. Feed water composition).

2.7. Water transport

The rate of water passage through a semipermeable membrane is defined in Eq. 9:

\[ Q_w = (\Delta P - \Delta P_{osm}) K_w (S/d) \]  

(9)

where \( Q_w \) is the rate of water flow through the membrane, \( \Delta P \) is the hydraulic pressure differential across the membrane, \( \Delta P_{osm} \) is the osmotic pressure differential across the membrane, \( K_w \) is the membrane permeability coefficient for water, \( S \) is the membrane area, and \( d \) is the membrane thickness.

This Eq. can be simplified to:

\[ Q_w = A \cdot S \cdot NDP \]  

(10)
where $A$ is water transport coefficient and represents a unique constant for each membrane material type, and $NDP$ is the net driving pressure or net driving force for the mass transfer of water across the membrane.

$A$ units are: g/cm²/s (gal/ft²/d/psi)

### 2.8. Salt transport

The rate of salt flow through the membrane is defined by Eq. 11:

$$Q_s = \Delta C \cdot K_s \ (S/d)$$  \hspace{1cm} (11)

where $Q_s$ is the flow rate of salt through the membrane, $K_s$ is the membrane permeability coefficient for salt, $\Delta C$ is the salt concentration differential across the membrane, $S$ is the membrane area, and $d$ is the membrane thickness.

$K_s$ units are: cm/s (ft/s)

Eq. 11 can be simplified to:

$$Q_s = B \cdot S \ (\Delta C)$$  \hspace{1cm} (12)

Where $B$ is the salt transport coefficient and represents a unique constant for each membrane type, and $\Delta C$ is the concentration gradient which is the driving force for the transfer of dissolved ions through the membrane.

Eqs. 10 and 12 show that for a given membrane:

a) Rate of water flow through a membrane is proportional to the $NDP$ differential across the membrane.

b) Rate of salt flow is proportional to the concentration differential across the membrane and is independent of applied pressure.

Salinity of the permeate, $C_p$, depends on the relative rates of water and salt transport through the reverse osmosis membrane:

$$C_p = Q_s/Q_w$$  \hspace{1cm} (13)

The fact that water and salt have different mass transfer rates through a given membrane creates the phenomena of water–salt separation and salt rejection. No membrane is ideal in the sense that it absolutely rejects salts; rather the different transport rates of water and dissolved ions create an apparent rejection. Eqs. 9–12 explain important design considerations in RO systems. For
example, an increase in operating pressure will increase water flow without significantly affecting salt flow, thus resulting in lower permeate salinity. On the other hand a higher recovery rate will increase the concentration gradient and result in higher permeate salinity.

2.9. Salt passage and salt rejection

Salt passage is defined as the ratio of concentration of salt on the permeate side of the membrane relative to the average feed concentration. Mathematically, it is expressed in Eq. 14:

$$SP = 100\% \left( \frac{C_p}{C_{fm}} \right)$$  \hspace{1cm} (14)

where $SP$ is the salt passage (in%), $C_p$ is the salt concentration in the permeate, and $C_{fm}$ is the mean salt concentration in the feed stream.

Applying the fundamental equations of water flow and salt flow illustrates some of the basic principles of RO membranes. For example, apparent salt passage is an inverse function of pressure; that is, the salt passage increases as applied pressure decreases. This is because with reduced feed pressure permeate flow rate decreases and hence dilution of salt, on the permeate side of the membrane, decreases as well (the salt flows through the membrane at a constant rate as the rate of flow is independent of pressure).

Salt rejection is the opposite of salt passage, and is defined by Eq. 15:

$$SR = 100\% - SP$$  \hspace{1cm} (15)

where $SR$ is the salt rejection (expressed as a percent), and $SP$ is the salt passage as defined in Eq. 14. Salt rejection is an important performance parameter of RO membranes, determining suitability of given membranes for various applications. The above relations for water and salt transport imply constant values of transport rates. However, salt and water transport rates are strongly affected by temperature, changing at similar rates with temperature fluctuations.

Example 5

RO element is tested at 15 recovery rate, $R_p = 0.15$
Feed salinity: 1500 ppm NaCl
Permeate salinity: 4.5 ppm NaCl
Average feed salinity = $0.5 \times \frac{C_i (1 + 1/(1 - R_p))}{(1 - 0.15)} = 0.5 \times 1500 \times (1 + 1/(1 - 0.15)) = 1632$ ppm
Salt passage = 100% \(\frac{4.5}{1632}\) = 0.28%
Salt rejection = 100% – 0.28% = 99.72%

2.10. Temperature effect on transport rate

Feed water temperature effect rate of diffusive flow through the membrane. For RO calculations the following equation is being used to calculate temperature correction factor (TCF), applied for calculation of water permeability:

\[
TCF = \frac{1}{\exp(C \left( \frac{1}{273 + t} \right) - \frac{1}{298})} \quad (16)
\]

where \(t\) is temperature °C, C is constant, characteristic of membrane barrier material. For polyamide membranes C values of 2500–3000 are being used.

It is customary for RO applications to use a temperature of 25°C (77°F) as the reference temperature, for which TCF = 1.0. The water and salt transport increases about 3% per °C. Fig. 2.5 shows a value of TCF in the temperature range of 5–50°C. By comparison, relative values of water viscosity are also included. There is striking similarity of both curves, suggesting that changes of water permeability with temperature are a result of viscosity changes.

The results in Fig. 2.5 suggest that due to increased permeability with temperature increase the operating feed pressure should be lower at higher temperature. This is indeed the situation in case of processing of low salinity feed (brackish applications). This is also the case for RO seawater applications in the low range of feed water temperatures. However, at feed water temperatures

![FIG. 2.5 Temperature effect on water permeability and viscosity.](image)
above 30°C the subsequent decrease of required feed pressure levels off. The
effect of increased permeability is reduced by increased osmotic pressure of the
seawater feed. In addition, increase of salt passage and potential need for partial
second pass processing, may actually result in higher overall power consump-
tion at the high end of feed water temperature (Fig. 2.6).

Example 6

RO system operates in temperature range of 15–28°C.

At feed water temperature of 28°C the NDP$_1$ required for the design capacity
is 7.5 bar (108.9 psi). What NDP will be required at 15°C to maintain the
design permeate capacity?

Value of constant $C$ (Eq. 16) for the membrane used is 2700.

For 28°C

$\text{TCF}_1 = 1/(\exp(2700(1/(273 + 28) – 1/298)) = 0.914$

For 15°C

$\text{TCF}_2 = 1/(\exp(2700(1/(273 + 15) – 1/298)) = 1.370$

$\text{NDP}_2 = \text{NDP}_1 \frac{\text{TCF}_2}{\text{TCF}_1} = 7.5 \times 1.370/0.914$

$\text{NDP}_2 = 11.2$ bar (162.4 psi)

2.11. Average permeate flux (APF)

Average permeate flux is another important design parameter of the
RO process. APF is combined permeate flow divided by the total membrane
area installed in the RO unit. Units: l/m$^2$/h or gfd (gal/ft$^2$/d)
APF = $Q_p/(EN \times MA)$ \hspace{1cm} (17)

where $Q_p$ = permeate flow rate  
EN = number of elements in the system  
MA = membrane area per element

**Example 7**  
RO system produces 400 m$^3$/d (105,700 gal/d). Membrane array consists of 3 pressure vessels, each housing 6 membrane elements. Each element has 37 m$^2$ of membrane area (400 ft$^2$). APF is calculated as follows:

$APF = 400,000 \text{ l/d} / (3 \times 6 \times 37 \text{ m}^2 \times 24) = 25.0 \text{ l/m}^2/\text{h}$

$APF = 105,700 \text{ gal/d} / (3 \times 6 \times 400 \text{ ft}^2) = 14.6 \text{ gfd}$

Conversely, design APF is used to determine the required number of membrane elements in the RO system for required permeate capacity.

**Example 8**  
RO system will produce 400 m$^3$/h (1,760 gal/min). The design flux rate is 25 l/m$^2$/h (14.7 gfd). Membrane element type selected for this system is ESPA2 with 37 m$^2$ of membrane area (400 ft$^2$) per element.  
Number of membrane elements required (NEPV):

$NEPV = 400,000 \text{ l/h} / (25 \text{ l/m}^2/\text{h} \times 37 \text{ m}^2) = 432 \text{ elements}$

$NEPV = 1,760 \text{ gal/min} \times 1440 \text{ min/d} / (14.7 \text{ gal/ft}^2/\text{d} \times 400 \text{ ft}^2) = 431 \text{ elements}$  
The number of elements required will be rounded up according to the number of elements per vessel. Assuming seven elements per vessel:

$431 / 7 = 61.6 \text{ pressure vessels}$

Number of elements required: $62 \times 7 = 434 \text{ elements}$

**2.12. Specific permeability of a membrane**

Specific permeability, or specific flux (SF), characterizes the membrane material in terms of water flux rate driven by the gradient of applied net driving pressure.

$$SF = \frac{APF}{NDP}$$ \hspace{1cm} (18)

Specific permeability depends on the resistance of the membrane to water flow. This resistance is a composite of the flow resistance of the membrane barrier, support layers and any foulant layer on the membrane surface. It is usually calculated
for the feed water temperature of 25°C. Specific flux units are: l/m²/h-bar (gfd/psi)

**Example 9**

RO membrane element is tested at the following test conditions:

- Feed salinity = 1500 ppm NaCl
- Recovery rate = 15% (0.15)
- Feed pressure = 10.3 bar (150 psi)
- Pressure drop = 0.2 bar (2.9 psi)
- Permeate pressure: 0.1 bar (1.5 psi)
- Permeate flow: 41.6 m³/d (11,000 gpd)
- Membrane area: 39.5 m² (430 ft²)

Average permeate flux = \( \frac{41.6 \times 1000}{24 \times 39.5} \) = 43.9 l/m²/h (25.8 gfd)

Average feed salinity = \( 1500 \times 0.5 \times (1 + 1/(1 - 0.15)) \) = 1632 ppm NaCl

Average osmotic pressure = \( \frac{1632}{1000} \times 0.77 \) = 1.3 bar (19 psi)

Net driving pressure = 10.3 – 1.3 – 0.1 – 0.5 \( \times 0.2 \) = 8.8 bar (128.0 psi)

Specific flux, SF = \( \frac{43.9}{8.8} \) = 4.99 l/m²/h/bar (0.20 gfd/psi)

### 2.13. Concentration polarization

As water flows through the membrane and salts are rejected by the membrane, a boundary layer is formed near the membrane surface. In this layer the salt concentration exceeds the salt concentration in the bulk solution. This in-
crease of salt concentration at the membrane surface is called concentration polarization. As shown on Fig. 2.7, during the RO process there is a convective flow of water and ions toward the membrane surface. Ions rejected by the membrane, diffuse back to the bulk due to the concentration gradient. The observed effect of concentration polarization is reduction of actual product water flow rate and salt rejection versus theoretical estimates.

Concentration polarization has the following effect on the RO process:

1. Increased osmotic pressure at the membrane surface than in the bulk feed solution, $\Delta P_{osm}$, and reduced net driving pressure differential across the membrane ($\Delta P - \Delta P_{osm}$).
2. Reduced water flow across the membrane ($Q_w$).
3. Increased salt flow across the membrane ($Q_s$).
4. Increased probability of exceeding solubility of sparingly soluble salts at the membrane surface, and the distinct possibility of precipitation causing membrane scaling.

The Concentration Polarization Factor (CPF) is defined as a ratio of salt concentration at the membrane surface ($C_s$) to bulk concentration ($C_b$):

$$CPF = \frac{C_s}{C_b}$$

An increase of permeate flux will increase the delivery rate of ions to the membrane surface and increase $C_s$. An increase of feed flow, parallel to the membrane surface, increases turbulence and reduces the thickness of the high concentration layer near the membrane surface. Therefore, the CPF will increase with increase of permeate flow ($Q_p$) due to increase of permeate flux, and will decrease with increase of the average feed flow ($Q_{favg}$).

$$CPF = K_p \exp\left(\frac{Q_p}{Q_{favg}}\right)$$

Where $K_p$ is a constant depending on membrane element geometry, namely configuration and dimensions of feed channels and of feed spacer.

Using the arithmetic average of feed and concentrate flow as an average feed flow, the CPF can be expressed as a function of the permeate recovery rate of a membrane element ($R_i$).

$$CPF = K_p \exp\left(2R_i/(2 - R_i)\right)$$
The value of the Concentration Polarization Factor of 1.2, which is the recommended limit by some membrane manufacturers, corresponds to 18% permeate recovery for a 1 m (40") long membrane element, as shown in Fig. 2.8. The value of CPF is applied in calculations of RO elements performance to express excess concentration adjacent to the membrane surface.
3

**Commercial RO/NF membrane technology**

The semipermeable membrane for nanofiltration and reverse osmosis applications consists of a film of polymeric material composed of a skin layer several thousands angstroms thick and spongy supporting layer approximately 0.25–0.50 mm (0.001–0.002") thick cast on a fabric support. The overall thickness of membrane is 0.15–0.20 mm (0.06–0.08"). The schematic configuration of membrane layers is shown in Fig. 3.1. Scanning electron microscopy (SEM) picture of cross section of cellulose acetate membrane, at 500 times magnification is shown in Fig. 3.2. Corresponding SEM picture of cross section of polyamide membrane is shown in Fig. 3.3.

The commercial grade RO membrane must have high water permeability and a high degree of semipermeability; that is, the rate of water transport must be much higher than the rate of transport of dissolved ions. The membrane material must be stable over a wide range of pH and temperature, and have good mechanical integrity. The stability of membrane performance: permeability and salt rejection, over a period of time at field conditions defines the commercially

![Cross section configuration of flat sheet RO-NF membrane.](image)
FIG. 3.2  SEM picture of cross section of cellulose acetate membrane.

FIG. 3.3  SEM picture of cross section of composite polyamide membrane.
useful membrane life, which is, for the current commercial membranes, in the range of 5 to 10 years.

There are two major groups of polymeric materials, which are used to produce commercial reverse osmosis membranes: cellulose acetate (CA) and polyamide (PA). Polymer structure (Fig. 3.4), chemical tolerance, membrane manufacturing, operating conditions, and performance differ significantly for each group of polymeric material.

3.1. Cellulose acetate membranes

The original cellulose acetate membrane, developed in the late 1950s by Loeb and Sourirajan, was made from cellulose diacetate polymer (3). Current CA membrane is usually made from a blend of cellulose diacetate and triacetate. The membrane is formed by casting a thin film acetone-based solution of cellulose acetate polymer with swelling additives from a trough onto a non-woven polyester fabric (Figs. 3.5 and 3.6).

![Chemical structure of cellulose triacetate and polyamide membrane material.](image)
Two additional steps, a cold bath followed by high temperature annealing, complete the casting process. During casting, the solvent is partially removed by evaporation. After the casting step, the membrane is immersed into a cold water bath which removes the remaining acetone and other leacheable compounds. Following the cold bath step, the membrane is annealed in a hot water bath at a temperature of 60–90°C. The annealing step improves the semipermeability of the cellulose acetate membrane resulting in a decrease of water trans-
port and a significant decrease of salt passage. After processing, the cellulose membrane has an asymmetric structure with a dense surface layer of about 1000–2000 angstrom (0.1–0.2 micron) which is responsible for the salt rejection property. The rest of the membrane film is spongy and porous and has high water permeability. Salt rejection and water flux of a cellulose acetate membrane can be controlled by variations in temperature and duration of the annealing step. Description of manufacturing process of cellulose acetate membranes and its properties can be found in number of publications (6, 8, 9, 10).

Cellulose acetate membrane polymer hydrolyzes rapidly at extreme pH. Therefore, the operating feed water pH range for CA membrane is 6–8. Accordingly, cellulose acetate membrane elements can be only cleaned in the narrow range of pH close to neutral (pH: 6–8). However, CA membrane polymer has sufficient tolerance to free chlorine that enables operation with chlorinated feed water and on line disinfection to control bacterial growth. For this reason cellulose acetate is still membrane of choice for applications where frequent disinfection of RO system with free chlorine is practiced, such as pharmaceutical industries and some food applications. Also, one of membrane manufacturers presently produces capillary RO membranes for seawater desalting using cellulose acetate polymer. Except for the previously mentioned applications, rest of the desalination market is dominated by the composite polyamide membranes in spiral wound configuration.

3.2. Composite polyamide membranes

The manufacturing process of composite polyamide membranes consists of two distinct steps (Fig. 3.7). First, a polysulfone support layer is cast onto a non-woven polyester fabric. The process of application of polysulfone and formation of ultrafiltration membrane is very similar to the process of manufacturing of the cellulose acetate membrane (Fig. 3.6). The polysulfone polymer solution is applied from a trough onto a moving polyester backing fabric. After polysulfone application and formation of UF membrane layer the fabric travels through water bath to remove solvent and is collected on a drum.

The polysulfone layer is very porous and is not semipermeable; i.e., does not have the ability to separate water from dissolved ions solution. However, it has high water permeability. In the next process step, the drum with polysulfone membrane is moved to the second machine where interfacial polarization takes place (Fig. 3.8). There, a semipermeable membrane skin is formed on the polysulfone
substrate by interfacial polymerization of two monomers, one: metaphenylene-diamine (MPD) containing amine groups and the other: trimesoyl chloride (TMC) provides carboxylic acid chloride functional groups. The polymerization reaction is very rapid and takes place on the surface of the polysulfone support forming a barrier, 1000–2000 angstrom thick. This barrier is responsible for the semipermeable property: passage of water and rejection of dissolved species. Following polymerization zone, membrane web enters a rinse bath. The membrane is rinsed to remove excess reagents and passed trough the oven to dry (4).

This manufacturing procedure enables independent optimization of the distinct properties of the membrane support and salt rejecting skin. The resulting composite membrane is characterized by higher specific water flux and lower salt passage than cellulose acetate membranes.

Polyamide composite membranes are stable over a wider pH range than the cellulose acetate membranes. However, polyamide membranes are susceptible to oxidative degradation by free chlorine, while cellulose acetate membranes can tolerate limited levels of exposure to free chlorine. Compared to a polyamide membrane, the surface of cellulose acetate membrane is smooth and...
has little surface charge (Figs. 3.9 and 3.10). Because of the neutral surface and tolerance to free chlorine, cellulose acetate membranes will usually have a more stable performance than polyamide membranes in applications where the feed water has a high fouling potential due to presence of dissolved organics such as with municipal effluent and surface water supplies.

The early composite membranes made of aliphatic polymers (9) were very sensitive to presence of oxidants and suffered from inadequate stability of performance in field conditions. The later generation of composite membranes, based on aromatic polyamide invented by Cadotte (4, 12) have some tolerance to free chlorine, good stability in wide range of feed pH (2–10) and shows excellent long term performance stability with majority of feed water types. This type of membrane material is used today almost exclusively to manufacture commercial RO membrane elements.

The variety of types of membranes made of composite aromatic polyamide includes seawater, brackish and nanofiltration membrane elements. Composite membranes are used in all areas of applications: seawater and brackish water desalting, potable water softening, wastewater reclamation, food processing and other industrial applications.
FIG. 3.9  SEM picture of surface of cellulose acetate membrane.

FIG. 3.10  SEM picture of surface of composite polyamide membrane.
Membrane module configurations

The reverse osmosis technology started with tubular and plate and frame configurations. Due to low packing density, these initial module configurations were gradually phased out of potable applications and at present are being very infrequently used in conventional reverse osmosis applications. However, new configurations of plate and frame modules are still being used in food processing applications and for treatment of waste streams including land fill leaches. In the past, the two major membrane module configurations used for reverse osmosis applications were hollow fiber and spiral wound. At present, the majority of RO membrane manufacturers offer elements in spiral wound configuration only.

4.1. Plate and frame configuration

The plate and frame configuration has been introduced at the early stages of development of reverse osmosis technology (15) and later on almost abandoned in favor of higher packing density spiral wound and hollow fiber configurations. Today the plate and frame modules are still used in applications where spiral wound and hollow fiber modules can not provide sufficient reliability or performance. These include treatment of streams with high concentration of suspended solids. One of such applications is reduction of volume of land fill leachate (63). In modern plate and frame configuration, the flow regime provides turbulent flow and short feed flow path. Therefore, the tendency for membrane scaling or fouling is significantly reduced. Due to the high cost of membrane modules, the plate and frame configuration is not used in commercial potable applications.
4.2. Hollow fine fiber membrane elements

The concept of hollow fine fiber (HFF) configuration module has been introduced by Mahon (13) in the early sixties. The HFF configuration utilizes semipermeable membrane in the form of hollow fibers which have been extruded from cellulosic or non-cellulosic materials (14). The fiber is asymmetric in structure and is as fine as a human hair, about 40–80 μ (0.0016–0.0030 in.) I.D. and 85–150 μ (0.0033–0.060 in.) O.D. (Fig. 4.1). Millions of these fibers are formed into a bundle and folded in half to a length of approximately 120 cm (4 ft).

A perforated plastic tube, serving as a feed water distributor is inserted in the center and extends the full length of the bundle. The bundle is wrapped and both ends are epoxy sealed to form a sheet-like permeate tube end and a terminal end which prevents the feed stream from bypassing to the brine outlet. The hollow fiber membrane bundle, 10–20 cm (4–8 in.) in diameter, is contained in a cylindrical housing or shell approximately 137 cm (54 in.) long and 15–30 cm (6–12 in.) in diameter. The assembly is called a permeator (Fig. 4.2). The pressurized feed water enters the permeator feed end through the center distributor tube, passes through the tube wall, and flows radially around the fiber bundle toward the outer permeator pressure shell. Water permeates through the outside wall of the fibers into the hollow core or fiber bore, through the bore to the tube

![Fig. 4.1](image-url)  
**FIG. 4.1** Configuration of the hollow fiber membrane and module.
sheet or product end of the fiber bundle, and exits through the product connection on the feed end of the permeator.

In a hollow fiber module, the permeate water flow per unit area of membrane is low, and therefore, the concentration polarization is not high at the membrane surface. The net result is that hollow fiber units operate in a non-turbulent or laminar flow regime. The HFF membrane must operate above a minimum reject flow to minimize concentration polarization and maintain even flow distribution through the fiber bundle. Typically, a single hollow fiber permeator can be operated at up to 50% recovery and still satisfy the minimum reject flow required. The hollow fiber module configuration allows a large membrane area per unit volume of permeator which results in compact systems.

Hollow fiber membrane modules are available for brackish and seawater applications. Membrane materials are cellulose acetate blends and aramid (a proprietary polyamide type material in an anisotropic form). Because of very close packed fibers and tortuous feed flow inside the module, hollow fiber modules require feed water of better quality (lower concentration of suspended solids) than the spiral wound module configuration. The hollow fiber modules are used mainly for desalting of seawater and treatment of good quality brackish water (well water). Due to fouling susceptibility of the conventional hollow fiber configuration, these module types are not used for desalting of municipal wastewater.
4.3. Spiral wound elements

The concept of the spiral wound membrane element device was introduced shortly after the invention of the hollow fiber configuration (15). In a spiral wound configuration two flat sheets of membrane are separated with a permeate collector channel material to form a leaf. The leaf assembly is sealed on three sides with the fourth side left open for permeate to exit (Fig. 4.3). A feed/brine spacer material sheet is added to the leaf assembly. A number of these assemblies or leaves are wound around a central plastic permeate tube. The permeate tube is perforated to collect the permeate from the multiple leaf assemblies (Fig. 4.3). During the element assembly process membrane leaves are rolled around the permeate tube in a spiral configuration (Fig. 4.4). The membrane leaves are kept in this form with a tape wrapped around the element and the outer shell, which is usually made of reinforced fiberglass.

The typical commercial spiral wound membrane elements are approximately 100 or 150 cm (40 or 60 in.) long and 10 or 20 cm (4 or 8 in.) in diameter (Fig. 4.6). The feed/brine flow through an element is in a straight axial path from the feed end to the opposite brine end, running parallel to the membrane surface. Fraction of the feed permeates through the membrane and flows through the permeate carrier fabrics to the central permeate tube. The remaining fraction of feed water continue to flow through the feed channel and becomes a concentrate (Fig. 4.4). The feed channel spacer is in the form of a two level (bi-planar) net. The strands in each level are parallel and crossing at about 90 de-
gree strands in the other level (Fig. 4.5). This two level net separates membranes from adjacent leaves and induces turbulence in the feed stream to reduce concentration polarization. The thickness of the feed channel is in the range of 0.7–0.9 mm (0.028–0.034”). However, the cross section of feed channel open to flow is much smaller, due to the presence of feed spacer.
Membrane manufacturers specify concentrate flow rate requirements to control concentration polarization by limiting recovery rate (or conversion) per element to 10–20%. Therefore, recovery rate is a function of the feed-brine path length. In order to operate at acceptable recoveries, spiral systems are usually staged with three to eight membrane elements connected in series in a pressure tube (Fig. 4.7). The concentrate stream from the first element becomes the feed to the following element, and so on for each element within the pressure tube.

Each element contains brine seal, which is in the form of flexible o-ring, usually position at the front end of element. The brine seal seals the space between the element outer wall and inner wall of the pressure tube. Brine seal prevents feed water to bypass the element, which would otherwise result in low flow through element and high recovery rate. Concentrate stream from the last element exits the pressure tube to the next processing stage or to waste. The permeate tubes of each element are connected to adjacent element through permeate interconnector, forming a common permeate tube. The first and the last element in the pressure vessel is connected through an adaptor to the pressure vessel permeate outlet (Fig. 4.8). Permeate from all elements in the pressure vessel exits the vessel as a common permeate stream. A single pressure vessel
with six to eight membrane elements connected in series can be operated at up to 50–70% recovery under normal design conditions.

The dimensions and geometry of spiral wound membrane elements is highly standardized. Spiral wound membrane elements produced by various manufacturers are of very similar configuration and outer dimensions. They can be operated in the same pressure vessels and are easily interchangeable. More recently a committee composed of representatives of major manufacturers of spiral wound elements has been evaluating feasibility of large scale commercial production of large diameter (>200 mm, 8”) elements (63). The conclusion of
the evaluation was that the optimum size of the future large elements should be
406 mm diameter by 1016 mm long (16" by 40" L). Such elements would have
about 4 times the membrane area (and the permeate flow) of the current, 200 mm
(8") diameter element and would provide some reduction of RO system capital
cost. One of the membrane manufactures (Koch Membrane Systems) provides al-
ready limited offering of 16–18" (406 mm–457 mm) diameter membrane ele-
ments, which are used in small to medium size RO units. It is expected that by
2008 most of the major manufacturers of membrane elements will include large
diameter membrane elements in their product offering.

4.3.1. Spiral wound element categories

Large desalination systems utilize membrane elements that are 1 m (40")
long and 200mm (8") in diameter. Smaller, 100 mm diameter elements are used
for small systems (light commercial, small potable) and pilot testing. The spiral
wound membranes are used commercially in three major application categories:
potable water softening (nanofiltration), brackish water desalting and seawater
desalting. They are categorized accordingly as nanofiltration, brackish and sea-
water elements. Although operated at different feed pressure range, they are of
almost identical configuration and utilize the same materials of construction.
One of the major difference between spiral elements used in low and high feed
pressure applications is feed permeate spacer. The type used in high pressure, sea-
water elements is of tighter type, reducing membrane embossing. Traditionally,
brackish water elements were manufactured with slightly higher (about 10%) mem-
brane area then seawater elements. At present, this difference is still maintained to
some extend but membrane area of elements in all categories has increased.

The representative properties and nominal performance of nanofiltration,
brackish and seawater elements are listed in Tables 1.1, 4.1, and 4.2. The nomi-
nal performances are measured during testing of a single element at nominal
test conditions. The nominal test conditions include feed salinity (as NaCl), feed
pressure, recovery rate and temperature (25°C). At field conditions, where op-
erating parameters are different then the conditions during the factory tests, the
elements are operating at a different performance level. In softening and brack-
ish applications 200 mm (8") diameter element will produce about 24 m³/d
(6400 gal/d). In seawater applications a single element of the same size will
produce about 12 m³/d (3200 gal/d). In brackish water RO systems the reduc-
tion of feed salinity will be about 97% i.e., permeate salinity will be about 3%
of the feed salinity. The corresponding values for seawater systems are about
99.0% reduction of feed salinity or producing permeate salinity of about 1.0% concentration of the feed salinity. In softening (nanofiltration) applications the actual system salt rejection will very much depend on type of elements selected. It can be as high as 90% for systems designed for salinity and hardness reduction and as low as practically insignificant for systems designed to remove color and dissolved organics only.
The nominal performances of membrane elements listed in Table 4.1 through 4.3 are difficult to compare directly. The one reason being is that the nominal test conditions are somewhat different for different model elements. The test conditions differ even for the elements listed in the same application category. The other reason is that performances in field conditions are very much different than the nominal performance values. Still, the nominal values are the basis for calculation of projected system performance. The projected field performances are calculated using membrane manufactures provided computer programs. The results of computer calculations are quite accurate for brackish and seawater applications. At present, the calculated results for soften-

<table>
<thead>
<tr>
<th>Element model</th>
<th>ESPA2+</th>
<th>ESPA4+</th>
<th>TMG20-430</th>
<th>BW30-LE440</th>
<th>BW30-LE-440</th>
</tr>
</thead>
<tbody>
<tr>
<td>Element dimensions</td>
<td>1m long, 200 mm diameter (40” × 8”)</td>
<td>1m long, 200 mm diameter (40” × 8”)</td>
<td>1m long, 200 mm diameter (40” × 8”)</td>
<td>1m long, 200 mm diameter (40” × 8”)</td>
<td>1m long, 200 mm diameter (40” × 8”)</td>
</tr>
<tr>
<td>Membrane area, m² (ft²)</td>
<td>39.5 (430)</td>
<td>39.5 (430)</td>
<td>39.5 (430)</td>
<td>40.5 (440)</td>
<td>40.5 (440)</td>
</tr>
<tr>
<td>Permeate flow, m³/d (gpd)</td>
<td>41.6 (11,000)</td>
<td>49.2 (13,000)</td>
<td>41.6 (11,000)</td>
<td>48.0 (12,700)</td>
<td>48.0 (12,700)</td>
</tr>
<tr>
<td>Salt rejection, %</td>
<td>99.60</td>
<td>99.60</td>
<td>99.50</td>
<td>99.0</td>
<td>99.3</td>
</tr>
<tr>
<td>Test feed pressure, bar (psi)</td>
<td>10.3 (150)</td>
<td>6.7 (100)</td>
<td>7.6 (110)</td>
<td>6.7 (100)</td>
<td>10.3 (150)</td>
</tr>
<tr>
<td>Test feed salinity, ppm NaCl</td>
<td>1500</td>
<td>500</td>
<td>500</td>
<td>500</td>
<td>2000</td>
</tr>
<tr>
<td>Test recovery rate, %</td>
<td>15</td>
<td>15</td>
<td>15</td>
<td>15</td>
<td>15</td>
</tr>
<tr>
<td>Test flux rate, l/m²/h (gfd)</td>
<td>43.5 (25.6)</td>
<td>51.4 (30.2)</td>
<td>43.5 (25.6)</td>
<td>49.1 (28.9)</td>
<td>49.1 (28.9)</td>
</tr>
<tr>
<td>Permeability, l/m²/h/bar (gfd/psi)</td>
<td>4.9 (0.20)</td>
<td>8.2 (0.33)</td>
<td>6.2 (0.25)</td>
<td>7.8 (0.31)</td>
<td>5.9 (0.24)</td>
</tr>
<tr>
<td>Relative salt transport value</td>
<td>17</td>
<td>21</td>
<td>22</td>
<td>49</td>
<td>34</td>
</tr>
</tbody>
</table>
ing applications are approximate and manual corrections based on field experience (usually pilot unit operation) have to be applied.

The membrane performance parameters that provide some insight into expected performance of membrane elements in field conditions are water permeability and relative salt transport value. The permeability is indicative of the required feed operating pressure. High permeability will results in low feed pressure required for a given flux rate during initial system operation. During the course of field operation the permeability may change due to fouling and/or membrane compaction. Relative salt transport value (RSTV), which is a product of

<table>
<thead>
<tr>
<th>TABLE 4.3</th>
<th>Representative nominal performance of seawater membrane elements.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Element model</td>
<td>SWC4+</td>
</tr>
<tr>
<td>Element dimensions</td>
<td>1m long, 200 mm diameter (40&quot; × 8&quot;)</td>
</tr>
<tr>
<td>Membrane area, m² (ft²)</td>
<td>36.8 (400)</td>
</tr>
<tr>
<td>Permeate flow, m³/d (gpd)</td>
<td>24.6 (6,500)</td>
</tr>
<tr>
<td>Salt rejection, %</td>
<td>99.8</td>
</tr>
<tr>
<td>Test feed pressure, bar (psi)</td>
<td>55.2 (800)</td>
</tr>
<tr>
<td>Test feed salinity, ppm NaCl</td>
<td>32,000</td>
</tr>
<tr>
<td>Test recovery rate, %</td>
<td>10</td>
</tr>
<tr>
<td>Test flux rate, l/m²/h (gfd)</td>
<td>27.6 (16.3)</td>
</tr>
<tr>
<td>Permeability, l/m²/h/bar (gfd/psi)</td>
<td>1.0 (0.04)</td>
</tr>
<tr>
<td>Relative salt transport value</td>
<td>6</td>
</tr>
</tbody>
</table>
multiplication of test flux rate by nominal salt passage, provides indication of
the salt passage. For the brackish and seawater elements low RSTV’s are asso-
ciated with low overall salt passage.

For nanofiltration membranes, due to high salt passage there is strong inter-
action of various ions during the transport process. In addition, some of the
nanofiltration membrane surfaces are strongly charged. Therefore, salt passage for
a given mixed ions composition could be significantly different than the nominal
salt passage determined using single salt solution. These issues will be dis-
cussed in more details in the chapters dealing with membrane performance cal-
culations and system design (see chapter 6 Calculation of system performance).
Performances of nanofiltration elements are discussed in details in chapter 16: Nanofiltration technology and applications.
5

RO system configuration

Configuration of RO system is affected to some extent by type of feed water being processed. Fig. 8.1 shows configuration diagram of RO system processing well water. This configuration is representative of brackish plant or seawater plants receiving feed water from wells. In such systems, feed water filtration is usually very limited, including only cartridge filtration. Systems processing surface water from open intake, brackish or seawater, require more extensive filtration of the raw water (Fig. 8.9). In such systems it may consist of single or two-stage media filtration, combined with flocculation and/or clarification. Wastewater reclamation systems utilize, almost universally, membrane pretreatment: ultrafiltration or microfiltration. Membrane pretreatment is being evaluated and extensively piloted for seawater desalination systems. However, no large capacity RO seawater plant with such pretreatment has been operated for a significant period of time so far but some systems are in an advanced design stage.

RO systems consist of the following basic equipment components:

• Feed water supply unit
• Pretreatment system
• High pressure pumping unit
• Membrane element assembly unit
• Instrumentation and control system
• Electric power supply system
• Permeate treatment and storage unit
• Cleaning unit
The configuration and operation of all components of RO system are designed to produce and maintain adequate quality of feed water to the membrane elements, maintain stable performance of all system components, produce design permeate flow and quality and maintain design process economics.

5.1. Membrane assembly unit

The membrane assembly unit (RO train) is the “heart” of the RO system. This is where the separation between water and dissolved species takes place. It consists of a stand supporting pressure vessels, interconnecting piping, and feed, permeate and concentrate manifolds. RO train also includes an instrumentation panel with local display of flow pressure and conductivities. In some systems permeate sampling panel is also included. This panel is a collecting point for permeate sampling tubing from individual pressure vessels.

Membrane elements are installed in the pressure vessels.

A pressure vessel has a permeate port on each end, located in center of the end plate, and feed and concentrate ports, located on the opposite ends of the vessel (Fig. 4.7). Each pressure vessel may contain from one to eight membrane elements connected in series. The permeate tube of the first and the last membrane element in the vessel is connected to the end plates of the pressure vessel (Fig. 4.8). Permeate tubes of all elements in the pressure vessel are connected to each other using interconnectors, practically forming one long permeate pipe inside the pressure vessel. On one end of each membrane element there is a brine seal (Fig. 4.7). The brine seal, in the form of V-shape o-ring, closes the passage between outside rim of the element and inside wall of the pressure vessel. This seal prevents feed water from bypassing the membrane module, and forces it to flow through the feed channels of the element. As feed water flows through each subsequent membrane element part of the feed volume passes through the membrane, and is removed as a permeate. The salt concentration of the remaining feed water increases along the pressure vessel (more extensive discussion on permeate salinity distribution is included in Chapter 5.5). Permeate tubes conduct the permeate from all connected elements. The collected permeate has the lowest salinity at the feed end of the pressure vessel, and increases gradually in the direction of the concentrate flow.

The RO system is divided into groups of pressure vessels, called concentrate stages. In each stage pressure vessels are connected in parallel, with respect to the direction of the feed/concentrate flow. The number of pressure vessels in each subsequent stage decreases in the direction of the feed flow. The
configuration is usually in the ratio of 2:1. Thus, one can visualize that the flow of feed water through the pressure vessels of a system resembles a pyramid structure: a high volume of feed water flows in at the base of pyramid, and a relatively small volume of concentrate leaves at the top. The decreasing number of parallel pressure vessels from stage to stage compensates for the decreasing volume of feed flow, which is continuously being partially converted to permeate. Permeate from all pressure vessels in each stage, is combined together into a common permeate manifold.

The objective of the taper configuration of pressure vessels is to maintain a similar feed/concentrate flow rate per vessel through the length of the system and to maintain feed/concentrate flow within the limits specified for a given type of membrane element. Very high flow through a pressure vessel will result in a high pressure drop and possible structural damage of the element. Very low flow will not provide sufficient turbulence, and may result in excessive salt concentration at the membrane surface. The limits of maximum feed flow and minimum concentrate flow are specified by membrane manufacturers for a given membrane element type depending mainly on combined height of the feed channels in the element and type of feed spacer net used.

Pressure vessels used in RO commercial systems are highly engineering products, especially design for RO applications. Configuration of pressure vessels manufactured to house spiral wound elements is highly standardized in respect of port-to-port dimensions, internal length and inside diameter. In this respect they are interchangeable, i.e., almost any commercial spiral wound element made by various manufacturers will fit into all commercial pressure vessel of any manufacturer. Materials of construction of pressure vessels are adapted to high salinity corrosive environment. Pressure vessel tube is made of epoxy based FRP materials. Feed and concentrate ports are made of corrosion resistant stainless steel. Pressure vessels are manufactured in accordance to ASME code. Maximum pressure rating of pressure vessels is according to application category. For brackish applications pressure vessels are rated up to 40 bar (600 psi). Pressure rating of pressure vessels for seawater application is up to 83 bar (1200 psi). For nanofiltration applications, a low pressure less expensive, products are also available.

5.2. Concentrate staging

A commercial RO unit usually consists of single pump and a multistage array of pressure vessels. A simplified block diagram of a two stage RO unit is shown in Fig. 5.1. The concentrate from the first stage becomes the feed to the
second stage; this is what is meant by the term “concentrate staging.” The flows and pressures in the multistage unit are controlled with the feed and concentrate valves. The feed valve, after the high pressure pump, controls feed flow to the unit. The concentrate valve, at the outlet of RO block, controls feed pressure inside the unit.

For a given RO unit, the number of concentrate stages will depend on the permeate recovery ratio and the number of membrane elements per pressure vessel. In order to avoid excessive concentration polarization at the membrane surface, the recovery rate per individual membrane element should not exceed 18%. It is common engineering practice to design brackish RO systems so that the average recovery rate per 40-inch-long membrane element will be about 6–8%. Accordingly, the number of concentrate stages for an RO unit having 6 elements per pressure vessel would be two stages for recovery rates over 60%, and three stages for recovery rates over 75%. With pressure vessels containing seven to eight elements, a two stage configuration would be sufficient for recovery rates up to about 85%.

Figure 5.2 shows drawing of an RO train in a single stage configuration. Feed, permeate and concentrate manifold are clearly indicated.

Figure 5.3 shows a picture of commercial, two stage, brackish train. The array is 32:14 pressure vessels with 7 elements per vessel. The picture shows two parallel feed manifold with 4 × 8 pressure vessel connected. Unit configuration is eight vessels high and six vessels wide. This translates into unit dimensions of 4.0 m high, 2.9 m wide and 8.0 m long (13.1’ × 9.5’ × 26’). The first
stage concentrate is collected from the 32 first stage pressure vessels and fed to 14 second stage vessels. The concentrate outlets from the second stage vessels are connected to the concentrate manifold. The concentrate throttling valve located on the concentrate pipe is shown as well. The train is equipped with permeate sampling panel, which enables sampling permeate conductivity from individual vessels and local display panel of flow, pressure, feed temperature and conductivities.

The unit contains total of 322 membrane elements, each 200 mm (8") diameter, 1000 mm (40") long, and is capable of about 8,000 m³/d (2.1 mgd) of permeate.
at recovery rate of 80%. The unit shown on Fig. 5.3 is equipped with pressure vessels that have one feed and one concentrate port. This leads to train configuration where two pressure vessels are connected to the feed or concentrate manifold at the given level, on each side. New version of pressure vessels, being introduced recently, are of multiport design. The “multiport” vessels have two large, entry-exit ports, at each end of pressure vessels. This configuration enables pressure vessel to be connected to feed or concentrate manifold and then to each other horizontally (or vertically), practically creating horizontal (or vertical) feed and concentrate lines (Fig. 5.4). Feed flow starts from feed manifold and then flows to the group of pressure vessels connected in parallel. The same situation, with reverse flow direction, takes place at the concentrate end of the train. Currently, the number of “multiport” pressure vessels that pressure vessel manufacturers recommend to connect together is five. Therefore number of pressure vessels connected to feed or concentrate manifold at given level could be increased to ten vessels (five vessels on each side of feed or concentrate pipe). This type of pressure vessels enables train design with lower requirement of high pressure piping, contributing to reduction of equipment cost.

5.3. Flow distribution

In some cases it is necessary to equilibrate permeate flow between stages, i.e., decrease permeate flow from the first stage and increase permeate flow from the last stage. This can be accomplished in one of two design configurations. One solution is to install a valve on the permeate line from the first stage. By
throttling this valve, permeate back pressure will increase, reducing net driving pressure and reducing permeate flux from the first stage (see Eq. 8). To compensate for lower permeate flow from the first stage the differential permeate flow is produced from the second stage by operating the RO unit at a higher feed pressure then would be required without permeate throttling. The other solution is to install a booster pump on the concentrate line between the first and the second stage (Fig. 5.5). The booster pump will increase feed pressure to the second stage. This configuration will result in lower permeate flow from the first stage and higher permeate flow from the second stage, i.e., more uniform permeate flux distribution. The advantage of the permeate throttling design is simplicity of the RO unit configuration and lower capital cost. However, this design results in additional power losses due to permeate throttling and higher power consumption. The interstage pump design requires modification of the interstage manifold and an additional pumping unit. The investment cost is higher than in the first design configuration, but the power consumption is lower.

5.4 Permeate staging (two pass systems)

For some applications, the single pass RO system may not be capable of producing permeate water of a required salinity. Such conditions could be encountered in two types of RO applications:

- Seawater RO systems, which operate on a very high salinity feed water, at high recovery ratio and/or at high feed water temperature.

- Brackish RO applications which require very low salinity permeate such as supply of makeup water for pressure boilers or production of rinse water for microelectronics applications.

![Fig. 5.5 Configuration of two stage RO unit with interstage booster pump.](image-url)
To achieve an additional reduction in permeate salinity, the permeate water produced in the first pass is processed again in a second RO system. This configuration is called a two pass design, or “permeate staging.” Fig. 5.6 contains schematic diagram of a two pass system.

Depending on permeate quality requirements, part or all of the first pass permeate volume is desalted again in the second pass unit. The system configuration is known as a complete or partial two pass system according to whether all of the 1st pass permeate or only some fraction is fed to the second pass unit. The first pass permeate is a very clean water. It contains very low concentrations of suspended particles and dissolved salts. Therefore, it does not require any significant pretreatment. The second pass RO unit can operate at a relatively high average permeate flux and high recovery rate without concerns of concentration polarization and scaling. The common design parameters for the second pass RO unit are average flux rate of 34 l/m²/h (20 gfd) and recovery rate of 85–90%. In a two pass system, the permeate from the first pass flows through a storage tank or is fed directly to the suction of the second pass high pressure pump. It is a common procedure in a two pass seawater RO systems to return concentrate from the second pass unit to the suction of the high pressure pump of the first pass unit. The dissolved salts concentration in the concentrate from the second pass is usually lower the concentration of the feed to the first pass unit. Therefore, blending feed water to the first pass with small flow rate of the second pass concentrate, reduces slightly the salinity of the feed to the first pass, and increases the overall utilization of the feed water (Fig. 5.7).

FIG. 5.6  Configuration of a two pass RO unit.
5.5. Partial two pass configuration

There are number of possible configuration of the two pass RO units. One configuration, which is a partial two pass system, is shown in Fig. 5.8. In this configuration the first pass permeate is split into two streams. One stream is processed by the second pass unit, and it is then combined with the unprocessed part of the permeate from the first pass. Provided that the partial two pass system can produce the required permeate quality, this configuration results in smaller second pass unit, therefore lower capital and operating costs, as well as
higher combined permeate recovery rate (utilization of the feed water), compared to a complete two pass system.

Another partial two pass configuration, which takes advantage of internal salinity distribution of RO permeate in the pressure vessel, is becoming increasingly popular in seawater systems. This unique concept of partial two pass system, designated as “split partial”, has been proposed in the past (51) but only lately is being implemented in large seawater systems (47, 52). In the RO system each subsequent element, in the direction of feed flow, produces water of increasingly higher salinity (Fig. 5.9). This is due to increasing feed salinity along the pressure vessel and decreasing permeate flux (Fig. 5.10) affected by the decrease of NDP. The NDP decrease results from increasing osmotic pressure and gradual decrease of feed pressure along the pressure vessel, due to friction losses. As the elements are connected together through the permeate tube, the permeate from the individual elements is mixed together and leaves pressure vessel through permeate port at composite salinity.

In a split partial configuration the first pass system is configured as a single stage unit. The permeate is collected from both ends of pressure vessels. The fraction collected from the feed end is of lower salinity and flows directly to permeate storage tank. The fraction collected at the concentrate end is processed with the second pass RO unit (Fig. 5.11). The salinity distribution between the two fractions of first pass permeate depends on their relative flow and

![Combined permeate salinity vs. element position in a pressure vessel.](image)
is shown in Fig. 5.12. As illustrated in Fig. 5.12, the salinity of both fractions increases, as the fraction taken from the low salinity end (feed end) increases. In other words, the salinity of the fraction to be processed by the second pass RO increases as the permeate flow rate taken from the concentrate side decreases. This is also the direction of increasing effectiveness of the split partial process in reduction of the overall system size and power consumption. The
The objective of the second pass unit is to remove salt (reduce TDS) from the first pass permeate stream. It is intuitively obvious that to achieve given salinity reduction, it will be more effective to process small flow of high salinity rather than the large flow of low salinity. Therefore, the “split partial” process takes advantage of internal separation of permeate salinity in the RO unit. In the conventional partial two pass configuration, after the permeate leaves the pressure vessel it has a uniform salinity and additional work has to be expended to reverse the mixing and achieve the separation.

Table 5.1 shows design parameters for seawater two pass system operating in a split partial configuration. The table lists the fraction of the high salinity permeate flow (fraction leaving pressure vessel at the concentrate side) that would be processed by the second pass RO and corresponding salinities of the low salinity, high salinity and blended stream.

For comparison the last column lists the fraction of second pass RO required to achieve the same blended flow salinity in conventional partial two
pass system. Two important observations could be concluded from the presented results:

1. Split partial configuration requires smaller second pass system capacity than the conventional partial two pass unit.

2. The capacity difference between this two configurations decreases with increased flow being processed by the second pass unit.

The flow rate processed by the second pass RO unit affects required capacity of the primary RO. With increased capacity of the secondary RO larger fraction of primary RO permeate is discharged as concentrate of the secondary RO. This has to be compensated by increased capacity of the primary RO. Therefore, it affects both the capital and operating cost (mainly power consumption but also some contribution from chemical consumption and membrane replacement cost). The benefits of split partial configuration have to be evaluated against somewhat higher cost resulting from additional permeate piping and more complex process control. Table 5.2 provides comparison of the membrane

<table>
<thead>
<tr>
<th>High salinity flow fraction (feed to 2nd pass RO), %</th>
<th>2nd pass capacity, %</th>
<th>Low salinity fraction, ppm TDS</th>
<th>High salinity fraction, ppm TDS</th>
<th>Blended flow salinity ppm TDS</th>
<th>2nd pass RO capacity required, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>90</td>
<td>81</td>
<td>108</td>
<td>351</td>
<td>18</td>
<td>87</td>
</tr>
<tr>
<td>80</td>
<td>72</td>
<td>108</td>
<td>383</td>
<td>31</td>
<td>84</td>
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<tr>
<td>70</td>
<td>63</td>
<td>113</td>
<td>422</td>
<td>43</td>
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<td>125</td>
<td>469</td>
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<td>50</td>
<td>45</td>
<td>134</td>
<td>532</td>
<td>78</td>
<td>72</td>
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<td>40</td>
<td>36</td>
<td>152</td>
<td>609</td>
<td>102</td>
<td>65</td>
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<td>173</td>
<td>717</td>
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<td>18</td>
<td>203</td>
<td>879</td>
<td>172</td>
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</tr>
<tr>
<td>10</td>
<td>9</td>
<td>249</td>
<td>1149</td>
<td>232</td>
<td>30</td>
</tr>
</tbody>
</table>
TABLE 5.2
Comparison of design parameters for split partial and conventional partial two pass system configurations

<table>
<thead>
<tr>
<th></th>
<th>First pass</th>
<th>Second pass</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Conventional design</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Permeate flow, m³/h</td>
<td>4,312.5</td>
<td>1,958.3</td>
</tr>
<tr>
<td>Processed by second pass, %</td>
<td>45.4</td>
<td></td>
</tr>
<tr>
<td>Recovery ratio, %</td>
<td>50.0</td>
<td>90.0</td>
</tr>
<tr>
<td>Feed pressure, bar</td>
<td>65.4</td>
<td>14.1</td>
</tr>
<tr>
<td>No. of pressure vessels</td>
<td>1,030</td>
<td>190</td>
</tr>
<tr>
<td>No. of elements</td>
<td>8,240</td>
<td>1,520</td>
</tr>
<tr>
<td>Power requirement, kWh</td>
<td>12,525</td>
<td>826</td>
</tr>
<tr>
<td>Combined power req., kWh</td>
<td>13,351</td>
<td></td>
</tr>
<tr>
<td><strong>Split partial design</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Permeate flow, m³/h</td>
<td>4,166</td>
<td>62.5</td>
</tr>
<tr>
<td>Processed by second pass, %</td>
<td>18.9</td>
<td></td>
</tr>
<tr>
<td>Recovery ratio, %</td>
<td>50.0</td>
<td>90.0</td>
</tr>
<tr>
<td>Feed pressure, bar</td>
<td>66.5</td>
<td>13.4</td>
</tr>
<tr>
<td>No. of pressure vessels</td>
<td>1,000</td>
<td>60</td>
</tr>
<tr>
<td>No. of elements</td>
<td>8,000</td>
<td>480</td>
</tr>
<tr>
<td>Power requirement, kWh</td>
<td>12,307</td>
<td>321</td>
</tr>
<tr>
<td>Combined power req., kWh</td>
<td>12,628</td>
<td></td>
</tr>
<tr>
<td>Difference of no. of elements, (%)</td>
<td>240 (2.9)</td>
<td>1,040 (68)</td>
</tr>
<tr>
<td>Power saving, kWh (%)</td>
<td>723 (5.4)</td>
<td></td>
</tr>
</tbody>
</table>

arrays of commercial RO seawater system of 100,000 m³/day (26.45 mgd) capacity, for split partial and conventional partial two pass configuration. For this particular design the split partial configuration results in approximately 13% savings of number of RO elements and pressure vessels and about 5% reduction of power consumption.
6

Calculation of system performance


The common approach to projecting performance of RO system is to calculate permeate flow according to the net driving pressure model and base calculations of permeate salinity on salinity gradient between feed and permeate as a driving force of the salt transport. The reference conditions are the nominal element performance, as tested at standard test conditions, defined by membrane manufactures. Single point calculations for basic system configuration can be conducted manually (as shown below). However repeated calculations, required for optimization of process design, are conducted using computer programs available from all major membrane manufacturers. The manual calculations process goes through the following steps:

1. According to the type of feed water source select membrane element type to be used and the value of system average permeate flux (APF).

2. Using nominal test conditions and nominal element performance calculate specific permeability of the selected membrane element (SP).

3. Using the above values of APF and SP calculate the required net driving pressure (Eq. 18).

4. Based on feed water composition, feed water type or project specifications select system recovery ratio and calculate average feed salinity (Eq. 6 or 7).

5. Calculate corresponding average feed osmotic pressure (Equation 2 or salinity–osmotic pressure relations).
6. Make assumptions regarding system array, pressure drop per stage and permeate back pressure.

7. Calculate required feed pressure (Eq. 8).

8. Calculate permeate salinity based on average feed salinity, average system permeate flux, nominal element salt passage and element permeate flux at nominal test conditions.

**Example 11**

Brackish two stage RO system. Feed salinity 2500 ppm TDS. Recovery rate: 85%. Average flux rate 27.2 l/m²/h (16 gfd). Feed water temperature 25°C.

Calculations of specific element performance

- Element type: ESPA2, membrane area: 36.8 m²
- Nominal element performance: 34.07 m³/d @ pressure 10.3 bar
- Salt rejection 99.6% @ flux rate 38.6 l/m²/h (22.7 gfd)
- Nominal test conditions: feed salinity 1500 ppm NaCl, recovery rate 15%
- Average feed salinity during nominal test: $1500 \times 0.5 \times (1 + 1/(1 – 0.15)) = 1632$ ppm NaCl
- Average feed osmotic pressure: 1.25 bar (18.1 psi)
- Nominal NDP: 10.3 – 1.25 = 9.05 bar (131.2 psi)
- Specific permeability: $38.6/9.05 = 4.26$ l/m²/h/bar (0.17 gfd/psi)

Calculation of system performance

- System NDP required: $27.2/4.26 = 6.4$ bar (93 psi)
- Friction pressure drop per stage 2 bar (29 psi), total for system 4 bar (58 psi).
- Permeate back pressure 0.5 bar (7.2 psi).
- Feed salinity 2500 ppm TDS, osmotic pressure 1.9 bar (28 psi)
- Average feed osmotic pressure: $1.9 \times 0.5 \times (1 + 1/(1 – 0.85)) = 7.3$ bar (105.8 psi)
- Required system feed pressure: $6.4 + 7.3 + 0.5 + 18.2 = 18.2$ bar (264 psi)

Permeate salinity

- Average feed salinity: $0.5 \times (2500 + 2500/(1 – 0.85)) = 9583$ ppm
- Permeate salinity: $9583 \times (1 – 99.6/100) \times (38.6/27.2) = 54$ ppm

Permeate salinity is function of average feed salinity and operating permeate flux rate as compared to the nominal flux (Eq. 13).

Additional corrections that should be applied to these calculations include correction for system configuration (flux distribution), temperature and element age. The above calculations are close approximation for two stage system treating brackish water. However, they were conducted treating a whole system as
an average element. Therefore, significant errors can be made when manually calculating RO unit performance at extreme values of feed salinity, temperature or permeate recovery. Computer programs have built in routine to provide adjustment for the above mentioned parameters. In seawater systems the apparent permeability is strongly affected by concentration polarization at high salinities. Therefore, additional “salinity” correction has to be applied in calculations of NDP and feed pressure.

6.2. Use of computer programs for projections of membrane performance

To enable large number of calculations required during design of RO systems, computer programs have been developed by membrane manufacturers. Various performance projection programs are quite similar in functionality, design of user interface, input values required and output format. The calculation procedure usually starts with input of feed water analysis (Fig. 6.1). After

![Computer projection program—feed water analysis entry screen.](image-url)
completing input to the water analysis screen the next screen includes entry of process and system design parameters (Fig. 6.2). These include:

- Feed temperature
- Feed water pH
- Membrane age
- Permeate flow
- Recovery rate
- Membrane model
- Membrane array
- Salt passage increase factor
- Flux decline coefficient

The element type is selected from elements look up table (Fig. 6.3).
The array and number of elements are manually adjusted targeting permeate flux to be within the range suitable for the feed water type being processed. Computer calculations start with calculation of specific flux for the selected element. This value is corrected for temperature and element age (or fouling factor). Similar corrections are applied to nominal salt rejection (salt passage). Next, the program conducts calculations of permeate flow and salinity for a single element, in a element by element mode, starting with the lead element and progressing toward the concentrate end of the system.

The calculations are based on projected feed salinity to a given element, specific permeability and value of feed pressure selected from some predefined range. The resulting permeate flow is used to calculate the recovery rate. The calculated recovery rate is used in turn to correct the average feed salinity, which again is used to calculate new permeate flow and permeate salinity. Then the subsequent results of two calculations of permeate flow are compared, and if the difference do not exceed preset value, the calculations are conducted for the next element. After completing calculations of permeate flow and permeate salinity for all elements in series, and combining them as the corresponding cumulative cumulative...

**FIG. 6.3** Computer projection program—element selection look up table.
values, the resulting permeate flow is compared with the input value. If difference is too large, new calculations are conducted with a new feed pressure range, adjusted accordingly to the difference between previous calculations and the input value of permeate flow. If the cumulative calculated value is sufficiently close to the input value, the calculations are completed and the results are displayed on the screen. The display (Fig. 6.4) includes major input values and results of calculations of pressures, flows, some more critical saturation values and composition of permeate.

At some site conditions it is necessary to design seawater system as partial two pass configuration. Most computer programs are flexible enough to enables calculations for multipass-multistage system configurations (Fig. 6.5). Additional configurations options may include permeate blending, permeate throttling, concentrate recirculation and interstage booster pump. Some programs may have auxiliary routines that enable calculation of power consumption associated with projected operating parameters (Fig. 6.6), calculation of operating cost or conduct simulation of chemical treatment of permeate.

FIG. 6.4 Computer projection program—performance calculation results display screen.
FIG. 6.5  Computer projection program—two pass unit design parameters entry screen.

FIG. 6.6  Computer projection program—calculation of power requirement screen.
The computer projection program produces summary of input and calculated parameters in a form of a printout (Fig. 6.7), which serves as a basis for RO process design.

The important information on the printout form includes basic process design information as summarized in Table 6.1. The information is divided into user input and calculated results. The first group of calculated results are of primary importance and may affect the design process directly. The results in the second category of values are of secondary importance, providing only general information about the process.

Operating parameters for number of representative system configurations, calculated using computer projection program, are listed in Appendix A.
### TABLE 6.1
Computer program information processing

<table>
<thead>
<tr>
<th>User input parameters</th>
<th>Water analysis, feed water temperature and pH, permeate flow, recovery ratio, flux decline coefficient, salt passage increase coefficient, membrane age, membrane element type, array.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Primary calculated results</td>
<td>Feed flow, feed pressure, concentrate pressure, permeate composition, feed pH, osmotic pressure, concentrate composition, pH, osmotic pressure, saturation values of scaling constituents, average permeate flux, concentration polarization factor, feed and concentrate flow per vessel.</td>
</tr>
<tr>
<td>Secondary calculated results</td>
<td>Recommended pump pressure, acid dosage, pH of permeate and concentrate, permeate flux, recovery rate and permeate salinity for individual elements.</td>
</tr>
</tbody>
</table>
Normalization of system performance

Performance of RO/NF system is a result of aggregate performance of individual membrane elements. Each subsequent element in a pressure vessel, connected in series, operates at different values of feed salinity and feed pressure. Along the system feed salinity increases and feed pressure decreases. The performance and operating conditions are significantly different than the corresponding nominal values. In addition, system performances are affected by fluctuation of operating parameters: feed salinity, temperature, recovery rate and feed pressure.

In order to identify intrinsic changes of membrane performance such as permeability or salt passage, at the early stages of membrane deterioration process, system operational data are recorded at frequency at least once per day and normalized performance are calculated. The generic method of RO performance normalization is described in the ASTM procedure (42). In actual commercial RO applications the following normalization approaches are adopted:

1. Normalization to the reference (initial) operating conditions of the plant.
2. Normalization to the nominal element(s) test conditions
3. Calculation of water transport and salt transport values for the membrane elements in operation.

In the normalization calculations process each set of plant (or desalting stage) flows, pressures and salinities data is initially reduced to the average values. These average values are assumed to be representative for an element positioned somewhere in the middle of the system, on the feed–concentrate cross section
line:, i.e., element that process an average feed salinity at an average applied
feed pressure and produces average permeate flow. The averages are calculated
based on feed–concentrate values. Then based on this data the water and salt
permeability are calculated. In normalization approach 1 every set of the per-
formance data of the system are being recalculated to the initial operating con-
ditions: temperature, average feed salinity and NDP.

Any of the above performance normalization method will provide good
presentation of membrane unit performance trend. Some advantage of the first
method is that, in addition to normalized permeate flow and salt passage, it usu-
ally also provides trend of the pressure drop. Pressure drop is an important indi-
cator of early stage of fouling, which results in blockage of the element feed
channels. In normalization approach 2 the performance of the system are calcu-
lated and presented as a performance of an average element, it would perform,
if tested at the nominal test conditions. Normalization approach 3 is very similar
to the first one. In this calculations performance of RO system is reduced to per-
formance of an average element. Then based on this data the water and salt per-
meability are calculated.

Normalization of salt passage and permeate flow is derived from the trans-
port relations: Eqs. 10, 11 and 12. According to these equations the salt passage
is function of salinity gradient and quantity of permeate available for dilution
(permeate flux rate). Therefore, salt passage at a given operating conditions
SP(1) is related to different operating conditions accordingly to the correspon-
ding average permeate flux rate (APF):

\[
SP(2) = SP(1) \frac{APF(2)}{APF(1)}
\]  

(22)

Permeate flow \(Q_p\) at condition 1 can be related to operating conditions 2
accordingly to corresponding net drawing pressures (ND) and temperature cor-
rection factors (TCF)

\[
Q_p(2) = Q_p(1) \frac{NDP(1) \text{TCF}(1)}{NDP(2) \text{TCF}(2)}
\]  

(23)

System information required and relations used for performance normaliza-
tion calculations are summarized in Table 7.1.

Example 12

Normalization of system performance to the initial operating conditions
(see Table 7.2)

Recovery rate – \(R_p\)

\[
R_p = \frac{Q_p}{(Q_p + Q_r)}
\]
TABLE 7.1
Summary of data and relations used in the normalization calculations

<table>
<thead>
<tr>
<th>Membranes information</th>
<th>System operating data</th>
<th>Normalized calculation results</th>
</tr>
</thead>
<tbody>
<tr>
<td>Total number of membrane elements, membrane area per element, nominal element flow and salt rejection.</td>
<td>Feed salinity or feed composition, feed temperature, permeate flow, concentrate flow, feed, concentrate and permeate pressure, feed, permeate and concentrate salinity</td>
<td>Parameter</td>
</tr>
<tr>
<td>System recovery rate</td>
<td>Equation 4</td>
<td>System average feed salinity</td>
</tr>
<tr>
<td>System average osmotic pressure</td>
<td>Equation 2</td>
<td>System NDP</td>
</tr>
<tr>
<td>Temperature correction factor</td>
<td>Equation 16</td>
<td>System average flux rate</td>
</tr>
<tr>
<td>System salt passage</td>
<td>Equation 14</td>
<td>Element nominal flux rate</td>
</tr>
<tr>
<td>System normalized salt passage</td>
<td>Equation 21</td>
<td>System normalized permeate flow</td>
</tr>
<tr>
<td>System normalized pressure drop</td>
<td>Equation 52</td>
<td>Element normalized salt passage</td>
</tr>
<tr>
<td>Element normalized permeate flow</td>
<td>Equation 22</td>
<td>Water transport coefficient</td>
</tr>
<tr>
<td>Salt transport coefficient</td>
<td>Equation 12</td>
<td></td>
</tr>
</tbody>
</table>

\[ R(1) = \frac{200}{200 + 50} = 0.80 \]
\[ R(2) = \frac{180}{180 + 60} = 0.75 \]

Concentration factor – \( CF \)
\[ CF = \ln\left(\frac{1}{1 - R}\right)/R \]
\[ CF(1) = \ln\left(\frac{1}{1 - 0.80}\right)/0.80 = 2.01 \]
\[ CF(2) = \ln\left(\frac{1}{1 - 0.75}\right)/0.75 = 1.85 \]

Average feed salinity, ppm – \( C_{favg} \)
\[ C_{favg} = C_f \cdot CF \]
\[ C_{favg}(1) = 2000 \times 2.01 = 4020 \]
\[ C_{favg}(2) = 2500 \times 1.85 = 4625 \]
Average osmotic pressure, bar (psi)–Posmavg
Posmavg = 0.77 \( \frac{C_{favg}}{1000} \)
Posmavg(1) = 0.77 \( \left( \frac{4020}{1000} \right) = 3.1 \) (45)
Posmavg(2) = 0.77 \( \left( \frac{4625}{1000} \right) = 3.5 \) (51)

Average permeate flux, l/m\(^2\)/h (gfd)–APF
APF = \( \frac{Q_p}{1000/(A \cdot EN)} \)
APF(1) = \( 200 \times \frac{1000}{(37 \times 210)} = 25.7 \) (15.1)
APF(2) = \( 180 \times \frac{1000}{(37 \times 210)} = 23.2 \) (13.6)

Temperature correction factor–TCF
TCF = \( \exp \left( 2700 \left( \frac{1}{273 + t} - \frac{1}{298} \right) \right) \)
TC F(1) = \( \exp \left( 2700 \left( \frac{1}{273 + 22} - \frac{1}{298} \right) \right) = 1.284 \)
TC F(1) = \( \exp \left( 2700 \left( \frac{1}{273 + 18} - \frac{1}{298} \right) \right) = 1.243 \)

Net driving pressure bar (psi)–NDP
NDP = \( P_f - 0.5 \left( P_f - P_c \right) - P_p - Posmavg \)

**TABLE 7.2**
System configuration and operating parameters

<table>
<thead>
<tr>
<th></th>
<th>Initial values of operating parameters, condition 1</th>
<th>Current values of operating parameters, condition 2</th>
</tr>
</thead>
<tbody>
<tr>
<td>Feed salinity</td>
<td>ppm TDS</td>
<td>2000</td>
</tr>
<tr>
<td>Permeate salinity</td>
<td>ppm TDS</td>
<td>30</td>
</tr>
<tr>
<td>Feed pressure</td>
<td>bar (psi)</td>
<td>14 (203)</td>
</tr>
<tr>
<td>Concentrate pressure</td>
<td>bar (psi)</td>
<td>10.5 (152)</td>
</tr>
<tr>
<td>Feed–concentrate</td>
<td>Bar (psi)</td>
<td>3.5 (51)</td>
</tr>
<tr>
<td>pressure drop</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Permeate pressure</td>
<td>bar (psi)</td>
<td>1.5 (22)</td>
</tr>
<tr>
<td>Feed temperature</td>
<td>C</td>
<td>17</td>
</tr>
<tr>
<td>Permeate flow</td>
<td>m(^3)/h (gpm)</td>
<td>200 (880)</td>
</tr>
<tr>
<td>Concentrate flow</td>
<td>m(^3)/h (gpm)</td>
<td>50 (220)</td>
</tr>
<tr>
<td>Element type</td>
<td>8040 ESPA2</td>
<td>8040 ESPA2</td>
</tr>
<tr>
<td>Number of elements</td>
<td>210</td>
<td>210</td>
</tr>
<tr>
<td>Membrane area per</td>
<td>m(^2) (ft(^2))</td>
<td>37 (400)</td>
</tr>
<tr>
<td>element</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>
NDP(1) = 14.0 – 0.5 (14.0 – 10.5) – 1.5 – 3.1 = 7.65 (111)
NDP(2) = 16.0 – 0.5 (16.0 – 11.0) – 1.5 – 3.5 = 8.5 (123)

Specific flux l/m²/h/bar (gfd/psi) – SF
SF = APF TCF/NDP
SF(1) = 25.7 × 1.284/7.65 = 4.31 l/m²/h/bar (0.172)
SF(2) = 23.2 × 1.243/8.50 = 3.39 l/m²/h/bar (0.136)

Salt passage, % – SP
SP = 100 \frac{C_p}{C_{favg}}
SP(1) = 100 \times \frac{30}{4020} = 0.74
SP(2) = 100 \times \frac{50}{4625} = 1.08

Normalized salt passage, % – NSP
NSP(2) = SP(2)APF(2)/APF(1)
NSP(2) = 1.08 \times 23.2/25.7 = 0.97%

Average feed flow m³/h (gpm)
Q_{fav} = \frac{Q_f + Q_c}{2}
Q_{fav}(1) = (200 + 50)/2 = 125 (550)
Q_{fav}(2) = (180 + 60)/2 = 120 (528)

Normalized pressure drop, bar (psi)
NPD(2) = DP1 (Q_{fav1}/Q_{fav2})^{1.4}
NPD(2) = 5 (125/120)^{1.4} = 5.29 (77)

**Table 7.3**
Summary operating parameters normalized to initial performances

<table>
<thead>
<tr>
<th>Normalized value</th>
<th>Initial values</th>
<th>Current values</th>
<th>Difference, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Specific flux, l/m²/h/bar (gfd/psi)</td>
<td>4.31 (0.172)</td>
<td>3.39 (0.136)</td>
<td>-21</td>
</tr>
<tr>
<td>Salt passage, %</td>
<td>0.74</td>
<td>0.97</td>
<td>+31</td>
</tr>
<tr>
<td>Pressure drop, bar (psi)</td>
<td>3.5 (51)</td>
<td>5.29 (77)</td>
<td>+51</td>
</tr>
</tbody>
</table>

**Example 13**
Normalization of system performance (Table 9, conditions 2) according to the nominal test conditions
Element type: 8040ESPA2
Membrane area: 37 m² (400 ft²)
Nominal permeate flow: 34 m³/d (9000 gpd)
Nominal salt rejection: 99.5%
Test pressure: 10.3 bar (150 psi)
Pressure drop 0.3 bar (4 psi)
Feed salinity 1500 ppm NaCl
Recovery rate 15%
Temperature 25°C

Permeate flux at nominal test conditions, l/m²/h (gfd)
P_FN = \frac{34 \times 1000}{(37 \times 24)} = 38.2 (22.5)

Concentration factor at nominal test conditions
C_FN = \ln \left( \frac{1}{1-0.15} \right) / 0.15 = 1.083

Average feed salinity at nominal test conditions
\bar{C}_f^{avgN} = 1500 \times 1.083 = 1624 ppm

Osmotic pressure at nominal test conditions, bar (psi)
\bar{P}_{osmN} = 0.77 \times \frac{1624}{1000} = 1.2 (17)

Net driving pressure at nominal test conditions, bar (psi)
DDPN = 10.3 – 0.5 \times 0.3 – 1.2 = 8.9 (129)

Average element permeate flux in the system (Table 9), l/m²/h (gfd)
Fluxavg(1) = \frac{200 \times 1000}{(210 \times 37)} = 25.74 (15.1)
Fluxavg(2) = \frac{180 \times 1000}{(210 \times 37)} = 23.16 (13.6)

Average element permeate flow in the system (Table 9), m³/d (gpd)
Q_{Pavg}(1) = \frac{200 \times 24}{210} = 22.85 (6036)
Q_{Pavg}(2) = \frac{180 \times 24}{210} = 20.57 (5434)

Average element permeate flow normalized to nominal test conditions
(NDP and TCF from Example 12), m³/d (gpd)
Q_{Pavg}^{N(1)} = 22.85 \times 1.284 \times \frac{8.9}{7.65} = 34.13 (9018)
Q_{Pavg}^{N(2)} = 20.57 \times 1.243 \times \frac{8.9}{8.50} = 26.77 (7072)

Permeate flow change between conditions 1 and 2: –22%

Average element salt passage normalized to nominal test conditions (SP(2))
from Example 12
SP_N(1) = 0.74 (25.74/38.2) = 0.50
SP_N(2) = 1.08 (23.16/38.2) = 0.65
Salt passage change: +30%
Average element salt rejection normalized to nominal test conditions, %
Rej (1) = 100 – 0.50 = 99.50
Rej (2) = 100 – 0.65 = 99.35

**Example 14**
Calculation of water transport and salt transport values for the membrane elements in operation.
Calculations of water transport value (A) are conducted according to Eq. 10 using values calculated in Example 13.

\[ Q_w = A S \overline{NDP} \]  \hspace{1cm} (10)
\[ A = \frac{\text{Flux}_{\text{avg}}}{\text{NDP}} \]  \hspace{1cm} (10a)

The units of \( A \) are s\(^{-1}\)

\[ A(1) = 25.70 \times \frac{1000}{10000 \times 3600}/(7.65 \times 1000) = 9.3E - 8 \]
\[ A(2) = 23.16 \times \frac{1000}{10000 \times 3600}/(8.50 \times 1000) = 7.6E - 8 \]

Difference of water transport coefficients –19%

Calculations of salt transport value (B) are conducted according to Eq. 12 using values calculated in Example 13.

\[ Q_s = B S \Delta C \]  \hspace{1cm} (13)
\[ B = \frac{\text{Flux}_{\text{avg}}}{\Delta C} \]  \hspace{1cm} (13a)

The units of \( B \) are g/cm\(^2\)-s

At low permeate salinity one can assume that \( \Delta C \) equals average feed salinity: \( C_{\text{favg}} \)

\[ B(1) = 25.70 \times 30 (1000/(10000 \times 3600))/4020 = 5.32E - 6 \]
\[ B(2) = 23.16 \times 50(1000/(10000 \times 3600))/4625 = 6.95E - 6 \]

Difference of salt transport coefficients +30%

The results listed in Examples 12–14 illustrate three different methods of performance normalization calculation. All calculations are based on the same principles that permeate flow depends on net driving pressure and temperature, and salt passage is function of salinity gradient. The calculation accuracy can be improved by incorporating the effect of temperature on salt transport and by more rigorous calculation of osmotic pressure.

The above methods of normalization are accurate enough for normalization of performance of brackish membrane elements. In normalization of performance
of low rejection nanofiltration elements, salinity of permeate should be included in calculations of NDP and salinity gradient. In system operating with high salinity feed (seawater), concentration polarization has significant effect on permeate flow. This effect should be accounted for, especially in calculations of normalized permeability. Otherwise, comparison of performance results obtained at different feed salinities will not provide meaningful results.
Feed water supply system and pretreatment

8.1. Well water

The configuration of the feed water supply system depends on the type of water source. The common sources of feed water for RO/NF plants are brackish wells, seawater intakes and treated municipal wastewater effluents. Less common sources are brackish surface waters, seawater beach wells or industrial wastewater effluents. Brackish wells providing water to an RO/NF system could be deep aquifer wells, reaching a depth of hundreds of meters, or shallow wells collecting surface water that infiltrated to shallow strata.

The composition of water from deep wells is in most cases very stable and of good quality. The fluctuations of water salinity and temperature are low. This is under conditions that the pumping rate is consistent with the permeability of the formation surrounding the well (production zone), and within the rate of aquifer replenishment.

The allowed pumping rate is determined through hydrologic evaluation of the water producing aquifer. This includes drilling pilot wells, test pumping and determination of step-drawdown for a range of water drawdown rates. In some aquifers, which are under influence of water of different salinities, the salinity could change with time as a result of pumping. In the majority of such cases the change of water quality is gradual. However, these changes should be considered at the system design stage. Shallow wells are more prone to quality fluctuations due to infiltration of seasonal surface run-off.

For RO applications it is important that the water supply well is constructed from materials that are compatible with a saline environment. The major concern is corrosion, therefore the well casing, pump and piping should be made of
a corrosion resistant alloy or polymeric materials. However, in the case of plastic materials, especially if epoxy based FRP components are used, it is important that the resin components of FRP material are completely cured so that they do not release even minute concentrations of organic compounds that could otherwise reduce membrane permeability.

Water originating from a properly designed well field has very low concentration of suspended solids. Therefore, in RO/NF systems treating well water, operating in the US, it is common practice not to include media filtration in the pretreatment process and use only cartridge filters on the feed line (Fig. 8.1). Cartridge filters for RO applications have a nominal porosity in the range of 5–15µ. The preferred porosity rating of filtration cartridges is 5µm. Feed water flow through cartridge filters should not exceed 1 m³/h (4.4 gpm) per 25 cm (10”) cartridge length. The schematic configuration of the cartridge filter housing is shown in Fig. 8.2. In this configuration the filtration cartridges are mounted in a vertical position. In large capacity plants horizontal cartridge filters (Fig. 8.3) are sometimes used. The important part of the cartridge filter housing is the baffle, which protects cartridges from direct impingement of suspended particles. Filtration cartridges are usually made of soft polymeric materials. Exposure to direct impingement of hard particles could result in abrasion of cartridges and eventually loss of integrity.

The role of cartridge filters is mainly to protect equipment located downstream (pumps and membrane elements) from the sudden appearance of particulate matter in the feed water. Such conditions could sometimes be experienced due to sudden sand or silt release from wells or from sand filters (in a system utilizing sand filters). Use of cartridge filters as a feed water filtration step for colloidal matter removal is usually prohibitively expensive in respect of the cartridge replacement cost.

---

**FIG. 8.1** Configuration of an RO unit (brackish or seawater) operating on a well water feed.
FIG. 8.2  Configuration of cartridge filters.

<table>
<thead>
<tr>
<th>Max flow, m³/h</th>
<th>Number of 25 cm cartridges</th>
<th>Diameter, cm</th>
<th>Height, cm</th>
<th>In-out diameter, cm</th>
<th>Weight, kg</th>
</tr>
</thead>
<tbody>
<tr>
<td>3.6</td>
<td>3</td>
<td>33</td>
<td>70</td>
<td>5.0</td>
<td>67</td>
</tr>
<tr>
<td>7.0</td>
<td>6</td>
<td>38</td>
<td>70</td>
<td>5.0</td>
<td>82</td>
</tr>
<tr>
<td>27.0</td>
<td>24</td>
<td>38</td>
<td>130</td>
<td>7.5</td>
<td>100</td>
</tr>
<tr>
<td>86.0</td>
<td>76</td>
<td>60</td>
<td>150</td>
<td>10.0</td>
<td>200</td>
</tr>
</tbody>
</table>
Extensive field experience shows that RO systems treating well water, with cartridge filtration as the only filtration step, operated successfully over the years. In some isolated cases, plants with this configuration experienced release of silt and/or sand from wells, preventing the cartridge filters from operating properly; the particulate matter eventually ended up reaching membrane elements. In almost all reported cases of such events, the solids intrusion and accumulation was limited to the lead elements only. This condition was eventually rectified by flushing of the lead elements (in the reverse flow direction, after turning them around and moving them to the end of the system) with some replacements.

One big advantage of a pretreatment filtration configuration limited to cartridge filters only, is the reduction of exposure of feed water to the outside environment, which is very convenient in the treatment of anaerobic water sources. Anaerobic water sources (for example water from a deep Floridian aquifer) may contain variable quantities of hydrogen sulfide and usually sulfate reducing bacteria are present as well. Had this water been exposed to air there would be a high probability of hydrogen sulfide being partially oxidized to elemental sulfur according to the following reaction:

\[
2\text{H}_2\text{S} + \text{O}_2 = 2\text{H}_2\text{O} + 2\text{S}
\] (24)

Elemental sulfur has very limited solubility in water or water based solutions and once deposited in the feed channel of RO elements, it cannot be removed. At the early stages of RO technology development attempts were made to oxidize hydrogen sulfide present in feed water with strong oxidants, prior to the RO. This process configuration almost always ended up in either fouling of the membrane elements with elemental sulfur or in oxidative damage of the membrane barrier. The design approach that provides stable system performance is to maintain anaerobic conditions of the water through the RO system. After the RO unit, hydrogen sulfide is removed from the permeate (and sometimes from the concentrate as well) either by aeration or oxidation. If the location of the RO system is close to an urban center the degasifiers can not vent off gasses to the air. The common solution in case of hydrogen sulfide aeration is to follow the degasifiers with an absorption system. In such a system hydrogen sulfide is absorbed on an iron based catalyst and eventually disposed as a solid waste.

For any feed water source, but especially for an anaerobic one, it is important to design a system that assures complete total exclusion of light from the feed water. Even slightly translucent materials used as a part of piping or storage tanks can transmit enough light to provide sufficient energy for bacteria to
grow. The problem of bio-growth due to some light transmission is quite common for unpainted FRP piping or water storage tanks made of plastic materials. If the feed water supply system consists of a number of wells, used as a combined source of feed water, it is important to evaluate compatibility of their mixture in respect of potential solids precipitation. According to what was described above about the nature of anaerobic water, water from an anaerobic source cannot be mixed together with water containing dissolved air due to presence of oxygen and possibility of hydrogen sulfide oxidation.

Seawater beach wells, sometimes used as a feed water source for seawater RO systems, are usually quite shallow. They can be built as a regular wells or Ranney wells or as a combination of both configurations (Fig. 8.4).

As is the case of brackish wells, seawater beach wells provide water with a low concentration of suspended solids. One of the major limitations of seawater wells is their limited output capacity, usually in the range of few thousand m³/day (few MGD). Because of the low recovery rate of seawater systems, e.g., 35–50%, beach wells can only support RO systems of a limited permeate capacity. Another problem with beach wells is in obtaining permits. The general public is quite sensitive about building any structure in the seashore area. At present, obtaining approvals for construction of a large number of beach wells necessary to support a large capacity desalination plant, can be a very difficult task.

As indicated in Fig. 8.1 the pretreatment for a well water based system is usually limited to pH adjustment and/or addition of a scale inhibitor together with cartridge filtration. For some feed water supply wells, which have a history

FIG. 8.4 Configuration of beach well (Courtesy of Collector Wells International).
of releasing sand, in line hydro-cyclones or sand screens are used. Water produced by shallow wells sometimes contains significant amount of suspended particles. The process designer may decide, based on pilot operation results, to use media filtration upstream of the cartridge filters. Inclusion of media filtration in treatment of questionable water quality could reduce the rate of RO membrane fouling and decrease the frequency (and cost) of cartridge filters replacement.

8.2. Surface water

8.2.1. Raw water intake and concentrate discharge

Brackish RO systems treating surface water are not very common. Large seawater RO systems, on the other hand, are based almost exclusively on surface water supply. The surface water is provided from an intake structure, which could be located off shore, or pumped from a lagoon or intake canal. The intake structure could be in the form of intake towers, shore intakes, siphon well intakes, floating intakes or submersible intakes. This last configuration is the prevailing one in large seawater RO plants. If possible, the submersible, off shore intake structure should be positioned at location of the most stable water quality. The location for the seawater intake is selected to ensure a 10–15 m (32–49 ft) water depth, even during low tide periods. Depending on the sea floor topography, it may require the intake inlet structure to be located hundreds of meters off shore. The submersible intake structure is usually designed and built to provide a water inlet at a level of 2–5 m (6–16 ft) above the sea floor to reduce the possibility of clogging by silt and sand (Fig. 8.5). The level of the inlet should be selected to consistently provide a good water quality, possibly free of algae and plankton. The screen size and a slow inlet flow rate (<0.3 m/sec, <1 ft/sec) should ensure low entrapment and impingement of water born organisms. The water flow in the intake conduit should be at much higher rate (1–1.5 m/s, 3–5 ft/s) to prevent deposition of sediments in the connecting piping. Very comprehensive descriptions of various intake configurations can be found in the book: “Water treatment plant design” (38), edited by AWWA.

In seawater plants about half of the feed flow is returned back to the ocean as a concentrate at a significantly higher concentration. The concentrate discharge structure should be positioned at sufficient distance from the intake to ensure that the water concentration at the intake is not affected by the concentrate salinity (Fig. 8.6). The example of an excess salinity decay profile during concentrate discharge is shown in Fig. 8.7 and discussed in reference 43.
Recently, some developers of large seawater projects have considered locating RO plants contiguous to some other larger seawater users. Convenient candidates are electric power generation plants with seawater cooled steam condensers. A schematic diagram of water flow in an RO system located adjacent to a power
plant is shown in Fig. 8.8. There are number of advantages and some potential problems with this option of a feed water supply configuration. Supply from a dedicated intake usually implies a dedicated outfall facility as well. Current regulations require careful design that will minimize any potential environmental
effect (43). The lengthy permitting process for the construction and operation of intake and outfall facilities makes location of an RO plant adjacent to an onshore located power plant a very convenient solution. In this process configuration the RO system utilizes the seawater discharged from the heat reject section of the power plant as a feed before it flows to the ocean. In a similar fashion, the RO concentrate is discharged to the same line, downstream of the feed uptake.

The temperature of seawater at the outlet from the power plant is usually higher by 3–10°C (5–18°F) than the water temperature at the intake. For locations with a low seawater temperature (<20°C, <68°F) the temperature increase due to operation of the power plant is beneficial since it increases membrane permeability. This allows the RO system to operate at lower feed pressure. However, at locations with high water temperatures, e.g., above 30°C (86°F), a further increase of the feed water temperature does not result in any significant decrease of feed pressure (Fig. 2.6). Depending on the feed salinity and recovery rate in the temperature range of 30–40°C (86–104°F), the effect of the higher rate of water permeability through the membrane at a higher temperature, is adversely compensated by increased osmotic pressure.

Higher feed water temperature also results in higher salt passage (shown in Fig. 2.6). If this increase of salt passage requires increased operation of the second pass, higher feed water temperature can actually result in higher power consumption of the RO plant. Location of an RO system, contiguous with a power plant may result in some feed water quality problems. It is common practice for a power plant to intermittently chlorinate the intake structure to reduce bio-growth. An additional periodic event, that may affect seawater quality, is cleaning of the heat transfer surfaces of the condenser. As a result of cleaning, small particle fragments are released to the cooling seawater and could end up in the RO feed. Both of these periodic events at power plants, i.e., intake chlorination and cleaning of the condenser heat exchange surfaces, should be addressed in the RO system design and operation to prevent potential membrane damage.

8.2.2. Conventional filtration pretreatment

Concentration of suspended solids in surface water sources is usually much higher than in well water and could fluctuate over a wide range. The fluctuations are due to seasonal weather patterns (rainy seasons) or variability of biological activity (algae blooming periods). Therefore, surface water pretreatment systems have to contain additional solid removal steps (Fig. 8.9), as compared to RO systems treating well water (Fig. 8.1). The treatment steps of surface seawater usually include coagulation, flocculation and media filtration. Sometimes sedimentation
and additional screening is included as an initial treatment step. As shown in Fig. 8.10 the feed water pretreatment system may include a number of treatment steps to improve water quality.

If disinfection is required, it is recommended to apply it intermittently to prevent bacterial after-growth and biofouling downstream. Disinfection with oxidants (free chlorine) requires subsequent injection of a reducing agent at some point before RO membranes to prevent membrane damage. The reducing agent most frequently used in RO systems, is sodium bisulfite. The stochiometric requirement is about 1.8 ppm of sodium bisulfite for 1 ppm of free chlorine. However, the usual dosing rate of sodium bisulfite in RO pretreatment systems is three times the concentration of free chlorine in the feed water. The reduction

![Fig. 8.9 Configuration of brackish RO unit operating on surface water feed.](image)

![Fig. 8.10 Configuration of two stage media filtration system in seawater RO plant.](image)
reaction between bisulfite and free chlorine is very fast and completed in matter of seconds.

After removing large size floating debris with traveling screens and settling, water is pumped to the next treatment step, which is usually in line coagulation and flocculation followed by direct filtration. In the so called direct filtration process, water is treated, as it flows through the delivery piping, with conditioning chemicals (coagulants) and filtered, without prior clarification, by media filters.

Coagulation and flocculation is a combined process of destabilization and conglomeration of colloidal particles to facilitate more effective removal in media filtration process. Colloidal particles in the water stream are negatively charged and electrostatic repulsion helps to maintain them in suspension. Coagulants are positively charged hydrolyzed metal salts that neutralize negative charges of suspended colloids and help to aggregate them into larger, heavier, more filterable solids. For coagulation ferric or alum salts are mainly used. During hydrolysis of these salts a complex polynuclear, positively charged species are formed in a matter of seconds. The solubility of these species is low and they form a dense, suspended flock. The action of coagulants is threefold: they adsorb colloidal particles on the flock surface, neutralize negative charges that surround colloidal particles and also enmesh suspended particles in the body of the flock formed. The effective quantity of coagulant required is specific to water composition, type of colloidal particles, water pH and temperature.

An excessive quantity of coagulant could have the undesirable effect of increasing the stability of colloidal particles. In addition, a high concentration of coagulant may increase dispersion of colloids, due to reversal of surface charges: formation of high density, positive charges on the colloids surface and mutual electrostatic repulsion. The initial estimation of the required dosing rate of the coagulant and the optimum pH range for the process is determined by conducting a “jar test” (described in chapter 12). The dosing rate of the inorganic coagulant is usually in the range of 1–30 ppm with pH in the range of 6–8. Following the results of the jar test the adjustment of the coagulation process parameters is conducted during the initial stages of commercial system operation.

For RO applications ferric salts are preferred over aluminum due to the lower tendency of forming deposits in the membrane elements. The solubility of hydrolyzed species of aluminum and iron depends on pH, which is at a minimum at about pH 6 for aluminum hydroxide and pH 8 for ferric hydroxide. The solubility of hydrolyzed ferric compounds is much lower than of corresponding aluminum species. Therefore, if coagulation is applied, the pretreatment system effluent, and subsequently RO feed water will have a lower concentration of
ferric ions than the potential concentration of aluminum ions at corresponding conditions. Accordingly, in case of ferric coagulant there will be a lower potential for precipitation as a result of feed water pH changes and/or due to the increase in concentration of dissolved species that occurs in the RO process.

As mentioned already coagulation is a very rapid process requiring just a few second to complete. However, effective coagulation requires intensive mixing to bring the coagulant in contact with a large number of colloidal particles. In RO applications this is usually achieved by incorporating static mixers positioned downstream of the coagulant injection point. Conversion of metal coagulants to the hydrolyzed form consumes alkalinity in the water, therefore the raw water pH is reduced (0.1–0.3 pH units), in proportion to coagulant dosing and alkalinity present, according to the following equations:

\[
\begin{align*}
FeCl_3 + 3HCO_3^- &= Fe(OH)_3 + 3Cl^- + 3CO_2 \\
AlCl_3 + 3HCO_3^- &= Al(OH)_3 + 3Cl^- + 3CO_2
\end{align*}
\]

Coagulation can be also conducted using long chain synthetic organic polymers, which could be of nonionic, anionic or cationic types. The nonionic and anionic polymers destabilize colloids by bridging particles together. The cationic type polymers have a dual action of bridging and neutralization the negative surface charges of the colloids. Cationic organic polymers can be used as primary coagulants. However, in RO pretreatment systems polymers are usually used as additives to enhance the effectiveness of metal based coagulants by binding flock particles together. In most cases, polymers are applied at low dosing rage, below 1 ppm, directly injected to the feed water downstream of the dosing point of the metal coagulant, at the location where hydrolyzed metal flock has been already formed. If polymers are used at a high dosing rate, and a carry over from the sand filters occurs, cationic polymers may react with anionic scale inhibitors and form a fouling layer on the membrane surface.

Flocculation, which follows coagulation, is a process of flock formation during gentle mixing. Flocculation is a slower process than coagulation and takes a number of minutes to complete. During flocculation, colloidal particles and some fraction of the dissolved organics are attached to the flock body, and are eventually retained on the filtration layer in granular media filters. In the granular media filtration process, suspended solids are removed through attachment to the filtration media particles and through blockage/capture by the filtration cake. The preferred process of filtration is capture of suspended solids with significant bed penetration as opposed to surface filtration, since the latter re-
sults in a faster increase of pressure loss and therefore shorter filter runs. In a single medium filtration bed, after number of backwash runs, fine size filtration media particles are aggregated at the top of the bed. This reduces penetration of suspended solids and therefore, mainly results in surface bed filtration. A graduation of the filtration bed from coarse to fine particles can be achieved in a dual media configuration by placing fine, high specific gravity, filtration media as the lower filtration layer and coarse, low specific gravity, filtration media as a top layer. Filtration media selection that provides a coarse to fine filtration bed configuration, includes anthracite (specific density 1.5–1.75 t/m³ (93–110 lb/ft³), effective size around 1 mm. as a top layer and silica sand (specific density: 2.55–2.65 t/m³ (159–165 lb/ft³), effective size around 0.5 mm, as a bottom filtration layer. Effective size means that size of 90% of filter media particles in the given lot is larger than the value indicated.

There is a variety of media filtration equipment configurations used in potable and waste water filtration. In RO applications the frequently used filter types are pressure or gravity filters in single or two stage configurations. Pressure filters are cylindrical pressure vessels filled with a layer(s) of filtration media. They could be configured for horizontal or vertical operation (Fig. 8.11 and 8.12). The diameter of pressure filters is limited by the transportation constraints.

FIG. 8.11 Dual media, vertical pressure filter configuration (Courtesy of Tonka Equipment Company).
Usually the diameter is limited to 3 m (10 ft). The above constraint limits the use of pressure filters mainly to smaller RO systems.

Some large RO seawater systems utilize pressure filters (Carboneras RO plant, Spain). However, they are of horizontal configuration, which enables larger filtration area per filter compared to the vertical type filters. When horizontal filters are used, the usual concern is to ensure good flow distribution and effectiveness of backwash of filter media. The common solution is to divide horizontal filters into compartments with independent feed and backwash.

Gravity filters have the configuration of rectangular tanks, usually made of concrete. An example of a gravity filter configuration is shown on Fig. 8.13. Influent water to the filter is supplied through a side channel. Filtrate leaves the filter through the effluent outlet located at the bottom of the filter. The same port serves as an entrance for the backwash water. Backwash operation is sometimes augmented by air. Backwash water is collected by the wash water troughs, located above the level that the media expands to during backwash. Some gravity filters utilize washing of the media surface using water jets. A typical dual media filter contains, in the direction of influent flow, 30–70 cm (12–28") layer of anthracite followed by 20–50 cm (8–20") of silica sand supported on a 10 cm (4") layer of gravel.

Dual media filters usually operate at a filtration rate of 7.5–15 m³/m²/h (3–6 gpm/ft²). As filtration progresses, the water passage between particles of the filtration media increases as the interstices are progressively blocked by the deposition of colloidal particles and coagulant flock, and pressure drop (head
loss) across filtration media layer increases. Pressure drop has to be maintained below a terminal value, usually below 2.5 m (8 ft), otherwise channeling or media breakthrough can develop.

At the end of a filtration run the filter bed is cleaned from deposits in a backwash step and the initial value of pressure loss restored. The length of filtration run depends on effluent water quality. The filtration run can last from 8 hours to a number of days. During the backwash step water passes through the filtration bed in the reverse direction causing bed expansion and removal of deposits. Backwash operation is conducted at a flow rate of 35–55 m$^3$/m$^2$/h (15–23 gpm/ft$^2$). Bed expansion during backwash is in the range of 20–50%. With increased water temperature, due to the lower buoyancy of the filtration media, the backwash flow rate has to be increased (about 2% per °C) to provide sufficient bed expansion. A special precaution should be applied for pressure filters to ensure that the backwash rate is not excessively high since it may result in removal of filtration media from the filter. In gravity filters a similar problem can be encountered. However, there the conditions of backwash and media level in the filter can be conveniently observed by plant operators.

Backwash is mainly conducted with filtered water. In seawater RO plants RO concentrate is sometimes used for filter backwash to reduce the filtrate requirement. For more effective bed expansion and the removal of deposits, backwash operation is sometimes combined with air scouring. Air mixed with
backwash water passes through the filtration bed at rate of 55–90 m/h (3–5 scf/ft²-min). Air scouring results in good bed expansion and effective mixing of the granular media in the filter. Therefore, for dual media filters, air scouring is followed by high rate water backwash to restore the original configuration of coarse to fine media segregation.

The typical backwash sequence includes: disconnecting the filter from the feed manifold, draining filter tank or vessel from water, backwash using an air and water mixture for 3–10 min followed by high rate backwash for 5–10 min, and reconnecting the filter to the feed manifold. Immediately after the backwash step, the filtrate quality is not as good as during normal operation, since it contains a high concentration of suspended particles. Filtrate quality gradually improves in a “ripening” process that could take up to 30 min. It is up to the plant operator to decide when to redirect filter effluent back to the membrane unit based on effluent quality (turbidity, SDI).

The filtrate volume required for a backwash operation is 2–3% of the system filtrate capacity per filtration stage. A backwash operation is usually conducted every 8–24 h of filter operation. It could be triggered by timer setting or pressure drop across filtration media.

At the majority of locations, the backwash effluent can not be discharged directly to any natural body of water, especially if the backwash water contains metal based coagulants. The current common procedure of backwash water disposal is on site treatment. It includes thickening, to increase solid concentration and dewatering in a filter press (see lower part of Fig. 8.10). Dewatered sludge is disposed as a solid waste and recovered water can be discharged or returned to the beginning of the pretreatment process.

For RO applications filtrate quality is determined through measurement of turbidity and SDI. There is no defined and consistent correlation between turbidity and SDI. However, field experience indicates, that to achieve an SDI below 3, the filtrate turbidity has to be below 0.1 NTU, preferably below 0.05 NTU. Except for relatively clean surface water sources, such low filtrate turbidity is difficult to achieve in a single stage filtration. Additional treatment steps may include clarification or diffused air flotation, prior to filtration.

A more frequent approach is to apply a two stage filtration treatment process to water with high and/or variable turbidity. It is common in a two stage filtration system that the first stage filters serve as roughing filters operating at high filtration rate. The second stage filters serve as a polishing filters operating at a lower filtration rate. However, filtration systems designed for the same filtration rate of both stages, or lower on the first and higher on the second stage, are also encountered. Coagulation and flocculation is usually conducted prior to
the first filtration stage. Some two stage filtration systems have provision for an interstage addition of coagulant and/or pH adjustment. The effluent from the media filters is collected in a clear well and subsequently pumped through micron rated cartridge filters and from there to the suction of high pressure pumps. If required for scale prevention, acid or scale inhibitor is added to the feed water prior to cartridge filters. Extensive information on the water treatment processes applied in the conventional RO pretreatment system can be found in number of water treatment handbooks, among them references 45, 46, and 123.

8.2.3 Membrane filtration pretreatment

Graeme Pearce*

Separation Mechanisms. The objective of pretreatment for an RO or NF system is to remove particles, reduce organics, and provide a feed that will not cause biofouling in the RO/NF elements. Conventional pretreatment technology relies on a combination of chemical treatment and media filtration to achieve conditioning of the feed to make it acceptable as an RO/NF feed.

The separation mechanism for UF and MF membranes differs from conventional treatment devices, such as granular or fibrous media filters. Media filters rely on a gravity removal mechanism. They have a nominal pore size considerably greater than the particles they are capturing. For a granular media filter, the grain size may be >100 micron, creating pores of a similar size. The absolute rating of such a filter will be of the same magnitude.

However, due to the depth of the media, and the tortuous path created for the feed as it moves through the media bed, relatively high removal rates can be achieved for particle sizes below the absolute rating. Media beds of 150 micron sand particles can routinely achieve removal efficiencies of 90-99% for particles down to 10–20 micron, and with coagulant addition, the removal rates are significantly better. The nominal rating of the depth filter means that the removal efficiency is variable, and is dependent on a host of environmental and operating parameters.

In contrast, UF and MF membranes operate by a surface removal mechanism, and resemble a fine screen or sieve. The pore size at the surface of a membrane is highly uniform, with a narrow pore size distribution. Particles larger than the size of the largest pore are rejected by the membrane surface,

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and remain on the feed or concentrate side. The bulk carrier fluid, and any particles finer than the largest pore, can pass through the membrane to the filtrate side.

UF can remove the finest particles commonly found in water supply, with the removal rating dependent upon the pore size of the active layer of the membrane; the typical removal capability of UF used for general water treatment is 0.01–0.02 micron. MF typically operates at a particle size an order of magnitude coarser than UF, e.g. approx 0.1–0.2 micron. UF therefore has the advantage over MF of providing a better disinfection barrier, since the pore size of UF will exclude viruses. However, in the application of RO/NF feed conditioning, both technologies are effective at reducing particle concentrations to a satisfactory level.

Although membranes provided a barrier to particles, colloids, and most microorganisms, dissolved components such as organics can still pass through the membrane. Organics may be a problem to the RO/NF, since they may cause fouling due to surface adsorption, or they may provide a food source to microorganisms. Sometimes therefore, chemical treatment with coagulants can still be beneficial prior to the UF/MF stage to adsorb organics, and be subsequently removed from the feed as fine particulates. If the concentration of dissolved organics is high, coagulant pretreatment would reduce both the fouling of the UF/MF stage, improve fluxes, and reduce cleaning frequencies of the RO/NF. However, coagulation is not needed in every case, and the need for coagulant should be established through pilot trails in the project development stage.

**Advantages of membrane pretreatment.** Since membranes provide a barrier to particulates, they provide significant benefits to the RO/NF, and to the overall system design. These benefits fall into two broad categories, namely reduced cost of the overall system, and improved on-stream time and security of supply.

The capital cost of membrane pretreatment normally exceeds that of conventional pretreatment by 20–50% depending on flow rate, feed quality, and local factors. However, the improved treated water quality can reduce the size of the RO by allowing a higher RO flux to be used. This advantage tends to be marginal for high salinity waters above 40,000 ppm TDS, since the RO flux at this salinity is limited by the osmotic pressure. However, for any feed below 30,000 ppm TDS, the RO flux could typically be increased by at least 25%, which would not only create a substantial saving for the RO system, but would mean that the pretreatment system would be smaller as well.

In addition to the savings due to the system size, UF/MF pretreatment has the advantage of requiring less space; typically, UF/MF provides a 33% saving in plant area, and this too can often translate into capital savings for the plant.
A further saving that results from the improved RO feed quality from UF/MF is that the replacement rate of the RO will be reduced, typically by 33%. Thus, if a 20% annual RO replacement rate is used for a conventional pretreatment system, the replacement rate following UF/MF may be reduced to 13%. Clearly, many factors influence replacement rate, both technical and commercial, but this gives an indication of the potential savings.

RO cleaning frequencies are typically reduced by UF/MF pretreatment, since SDI’s are lower, and the RO feed will have very low particle concentrations. This results in considerable lengthening of the interval between cleans, which not only reduces the cost of cleaning chemicals, and the generation of chemical waste, but improves on-stream time for the RO system. In a typical whole life cost comparison, the savings in RO dosing and cleaning chemicals will equate to the additional cost of UF/MF.

The UF/MF system itself may have higher running cost than conventional pretreatment. Although the chemical usage and disposal costs of conventional pretreatment will be significantly higher than UF/MF, the equipment and media of the conventional system are long lived, requiring low maintenance. For the UF/MF system, the membranes will need to be replaced from time to time. Membrane life is typically 5–10 years, so this requires a moderate running cost, but this should equate to the savings by using a lower RO replacement. Power cost for UF/MF is comparable to, or in some cases, slightly higher than conventional, but chemical usage in the pretreatment itself is lower.

Improved on-stream time for the RO with UF/MF pretreatment can be seen as both a cost issue and a security of supply issue. Not only does this benefit arise from reduced RO cleaning, but feed variability, occurring due to storms, can result in out of spec feed from conventional pretreatment. Operators then have the choice of shutting down, or operating outside RO warranty limits. It is not always appreciated that the effect of the excursion may be much longer lived than the excursion itself in terms of the effect on the RO membranes.

Cost study for UF/MF vs. conventional pretreatment. The factors listed above have been quantified in a study example, in which the costs are compared for UF/MF vs. conventional pretreatment for a nominal 50 mld feed to an RO system. The approximate overall cost for the complete RO system of this capacity with pretreatment, operating at 50% recovery, would be of the order of $20M. Pretreatment would be approximately 15–20% of the overall cost, as indicated in the example cost breakdown for UF/MF and conventional pretreatment systems shown in Table 8.1.
In this example, UF/MF pretreatment exceeds the cost of conventional pretreatment by approximately $1m.

In terms of running cost, the pretreatment power and chemical costs are very much lower than those of the RO. UF/MF has slightly higher power and somewhat lower chemical cost (due to reduced dosage of coagulant) than conventional pretreatment. UF/MF will save some of the RO chemicals cost, used in the process. Since this is significant, typically 5 cents/m³, a modest saving will potentially justify the additional UF/MF capex in a whole life evaluation. This would need to be confirmed in a real life case study, but the payback of the additional capital could well be in the region of 5–10 years.

There is a replacement cost for UF/MF membranes, but at approximately 1 cent/m³, it is almost exactly the same as the saving in RO replacement cost by having a 33% saving in RO replacement rate, e.g. 20% for conventional pretreatment being reduced to 13% for UF/MF, as explained above.

A saving not reflected in the capex comparison above is the smaller footprint of UF/MF. This will probably claw back some of the $1m cost differential,
but the impact will be determined by local conditions. The other factor which is difficult to evaluate, is the security of supply issue, since on-stream time will be improved by UF/MF, but the monetary value assigned to this benefit is hard to quantify.

The final factor which will have a major influence on the complete system is the influence of UF/MF pretreatment on RO flux. A modest increase, if allowed by the feed salinity as discussed in the previous section, would easily pay for the additional capex of UF/MF. This would need to be evaluated on a case by case basis for a given feed TDS, but it is likely that this factor alone would justify UF/MF pretreatment for any feed below 35,000 ppm.

**Membranes for UF/MF.** Commercial UF/MF membrane systems span the range from fully hydrophilic, e.g. Cellulose Acetate (CA), to fully hydrophobic, e.g. Polypropylene (PP). Between the two extremes, there is the Polysulfone (PS)/Polyethersulfone (PES) family, Polyacrylonitrile (PAN), Polyvinylidene Fluoride (PVDF), and a few other less common polymers. The most widely used polymers for UF/MF both in terms of the number of companies offering products, and the installed product base, are PES and PVDF.

Although both PES and PVDF are moderately hydrophobic in themselves, the membrane making process may utilize additional hydrophilic polymer components as pore formers or as modifiers, resulting in a final characteristic which is to some extent hydrophilic. The degree of hydrophilicity varies between products.

In water treatment, a hydrophilic membrane has some obvious advantages. Firstly, the membrane is easily wetted, and this results in high permeabilities relative to the pore size. Secondly, the organic fouling constituents often present in surface water sources would readily attach to a hydrophobic surface. A hydrophilic surface tends to resist absorptive attachment by organics, and such a surface is referred to as a low fouling surface.

In water treatment applications, both UF and MF membranes utilize a low energy process, often known as ‘direct flow,’ in which a dead end filtration cycle is followed by an intermittent backwash. The principle of direct flow is that particulates accumulated in the feed channel during the filtration cycle, are expelled by the periodic backwash. Either the backwash is carried out at a higher flux than the filtration cycle, e.g. by 2.5 times, or air is introduced into the backwash at a similar flux to the filtration cycle. The difference in energy between filtration and backwash cycles will tend to displace particles loosely lodged near the entrance of the pores. For a successful stable operation, it is important that the filtration cycle transmembrane pressure (TMP) does not increase to too high a value. A high TMP will result in particles penetrating too far
into the pores, i.e., plugging the pores, and the backwash cycle will then not be fully effective at recovering permeability. The critical pressure above which the lodging of particles in pores becomes plugging is dependent upon pore size, and varies significantly between UF and MF.

If MF is operated at a low TMP, of say <0.1 bar (< 1.5 psi), pore plugging can be avoided, since the forces are too low to force the fine particles into the depth of the pore structure. However, MF tends to be operated in typical water treatment application at significantly higher TMP’s than this, in an effort to reduce membrane area requirements. The preponderance of fine particles to plug the MF membrane pores results in a classic saw-tooth TMP profile, with TMP rising inexorably despite gas assisted backwash, necessitating aggressive chemical recovery cleans. The problem with this type of operation is that the feed pumping system has to be designed for the worst case TMP, leading to higher power costs.

MF tends to do better in applications where there are larger solids present in significant concentrations, especially if the solids can form a pre-coat membrane. Robust flocculated solids are particularly suitable. Under these conditions, stable performance may be possible at pressures well above the 0.1 bar (1.5 psi) quoted above. For example, wastewater applications operate well with MF membranes, giving stable performance up to at least 0.4 bar (6 psi).

Typical UF performance is characterized by stable permeability, with preventative strategies to prevent the rise of TMP as the filtration cycle proceeds. These strategies include backwashing and air or chemically enhanced backwash (AEB or CEB). Clean In Place (CIP) is available as a recovery strategy if the preventative strategies fall short, or cannot cope with an upset in feed quality or operation.

The ideal TMP for UF is normally <0.6 bar (<9 psi) to avoid fouling problems, with a top limit under normal conditions of <1.2 bar (17 psi). With TMP <0.6 bar (< 9psi), it should be possible to maintain stable performance just by use of a maintenance cleaning strategy with CEB’s. As pressures rise above 0.6 bar (9 psi), there will be a progressive need for recovery cleaning with CIP’s, or more frequent CEB’s, with the possibility of backflush TMP’s rising too high.

**Modules for UF/MF.** RO and NF product offerings are dominated by spiral wound elements. In contrast, UF/MF is dominated by hollow fiber/capillary modules. The term hollow fiber tends to be used for internal diameters <0.5 mm (<0.02"); capillary is used for internal diameters >1.0 mm (>0.04"). Both terms are used for the middle ground of 0.5–1.0 mm (0.02–0.04").
The main reasons that UF/MF has developed differently from RO/NF are as follows:

- Hydrodynamic efficiency
  The high permeability of UF/MF makes the spiral construction inefficient due to excessive pressure drop. Hollow fiber modules can avoid these limitations.

- Backwashability
  The fact that UF membranes are made from a single polymer or polymer solution, enables the membranes to be backwashed without the potential for delamination that would occur in a multi layer flat sheet membrane, such as those used for RO and NF. The hydrodynamics of the backwashing operation, which normally takes place at a higher flow than the filtration cycle flow, provides an important advantage to the hollow fiber configuration.

- Integrity
  The hollow fiber has higher seal integrity at the potting interface than the multi layer membrane used in a spiral. A small degree of leakage is inevitable in a multi layer membrane. This is due in part to the nature of the coating process used to make multi layer membranes, which gives some small defects. In addition, it is difficult to seal the ends of the substrate layer. This gives the spiral a lower degree of intrinsic integrity. More importantly, the spiral cannot be integrity tested in situ (other than by monitoring filtrate quality), or repaired. Thus for municipal water applications, where integrity has to be demonstrated both at the outset and in on-going operation, the spiral configuration is therefore unsuitable.

- Pretreatment
  Spirals need good quality feed in order to prevent the feed spacer from becoming clogged with particulates. Hollow fiber and capillary modules can tolerate higher feed solids loading, and require less pre-treatment.

  The feed to the hollow fiber module can either be introduced into the inside of the fiber lumen, with the permeate being withdrawn from the shell, or it can be fed to the shell, with permeate being taken from the fiber lumen. The first of these two options is the inside feed option, the second, the outside feed option.

  For many commercial fibers, the surface area of the outside surface is at least twice as great as the inside surface. This means that for the same flux, the outside feed system should produce at least twice the output of an inside feed system. However, in reality, there are other constraints, which significantly limit this advantage.
The inside feed configuration has feed entering the lumen during filtration, with the flow direction being reversed during backwash. The high flow of the backwash supply to the shell arrives along the entire length of the outside surface, the main constraint to efficiency being the removal of the backwash effluent from the fiber lumen. For system designs, which operate single modules in parallel, backwash effluent, can be removed from either end of the module, ensuring an efficient removal of accumulated debris. Multiple element systems remove debris from just one end, employing by-pass tubes to help regain the loss of efficiency.

The outside feed configuration cannot be operated with acceptable efficiency using a permeate backwash due to the high pressure drop along the length of the fiber at the high flow rates. In consequence, configurations based on outside feed utilize air during the backwashing process. Either the air passes back through the fiber in the reverse of the flow direction (air backwash), or air scour the outside of the fiber whilst a low flow rate permeate backwash passes through from the fiber lumen. A true air backwash requires a fully hydrophobic fiber. It provides an effective way of removing the debris accumulated during the filtration cycle, but is mechanically aggressive to the fiber itself, and can result in stress, fatigue, and, ultimately, fiber failure.

The outside air scour option is also aggressive, causing particular stress at the tubesheet interface. The tubesheet is the area in which the fiber is bonded to the shell or manifold, normally with an epoxy resin. A strong fiber is required to withstand the stress. Both air backwash and air scour utilize an appreciable quantity of air, resulting in a significant running cost, but reduce the cost and improve the recovery associated with the permeate backwash. Due to the aggression of the air assisted backwash process, outside feed systems employ fibers known for their strength and flexibility, such as PVDF, whilst the milder conditions of the inside feed configuration allows the use of the high permeability tighter cut off PES membranes.

All UF/MF systems utilizing an inside feed configuration use a shell or housing to contain the membranes. A pressurized feed is used to supply feed to the modules, hence this type of system is known as a pressurized system. The shell has a pressure rating, and allows the system to be operated against a back pressure if required. Some pressurized systems use an outside feed configuration, but these are mainly MF systems, in which gas is used to assist the backwash process.

The other type of UF/MF system is known as a submerged membrane system. The submerged system operates with the membranes unencapsulated, and uses out an outside feed configuration. Normally, the membranes would simply
be immersed in an open tank, and a vacuum applied to the permeate side to draw the filtrate through the membrane.

The main advantage of the submerged system is the saving on the cost of the pressure vessel. In addition, submerged systems can tolerate high solids without suffering the problem of lumen plugging that constrains inside feed pressurized systems. However, a major disadvantage is that the average concentration of feed near the membrane is higher than for pressurized systems, unless there is a full tank drain down at backwash or a zoned design. This is because the backwash alone does not completely remove the effluent as it does in the pressurized system. For a typical system operating with 95% recovery, the average feed concentration would be approximately three times as high for the submerged system than for the pressurized. Accordingly, submerged systems normally operate at lower flux, though the disadvantage is much less than three-fold, and in fact is more typically about half that. This is because at the lower fluxes and operating pressures of submerged systems, the higher solids form an open pre-coat at the membrane surface, enabling the system to tolerate the higher solids burden.

Another aspect of the comparison is that pressurized systems use chemical soaks in a chemically enhanced backwash (CEB), and recirculatory cleaning with a Clean In Place (CIP). The ability to use two completely different methods is an advantage in improving cleaning efficiency, and provides flexibility. The submerged system also uses a frequent CEB, but the CIP is different to the pressurized system. For the submerged CIP, chemicals are introduced to the feed side either by draining the tank, or removing the membranes from the feed tank and soaking in a separate tank. This is therefore a lengthy, and potentially inconvenient process, and does not allow the straightforward efficiency advantage of the recirculating chemicals in situ.

Operating pressure however is normally a factor that favors submerged systems, since pressures at the lower fluxes are often lower, and the vacuum pressure may be provided from a natural head difference at the site.

In the economic comparison between pressurized and submerged systems, several factors therefore need to be taken into account. The submerged system saves on pressure vessels, but requires a tank. Since operating fluxes are lower, more membrane area may be required. For larger plants, and particularly in a retrofit situation, submerged systems can gain a significant capex advantage. In addition, the operating cost advantages of submerged systems are more likely to be significant for larger plants. However, small to medium plants are likely to be more suitable for pressurized technology, especially if feed quality is good,
and for these plants operating costs are less important. Between these two limits, there is a large gray area, where local factors, and details of the application will determine which is the most suitable technology.

**Applications.** In 2004, the total UF/MF market for water and wastewater was approximately $450m, of which 40% was in RO pre-treatment. The use of UF/MF pre-treatment prior to RO is often characterized as an Integrated Membrane System (IMS). The drive towards IMS is partly due to the improved feed water quality to the RO, but also due to the reduced risk for the RO warranty resulting from employing a barrier process in the pre-treatment stage. The main IMS applications can be categorized as follows:

- Wastewater
- Boiler feed water
- Municipal surface water
- Seawater
- General process

The use of UF/MF and membrane bioreactors (MBR’s) in wastewater pre RO has already become well established, and is more widely used than conventional treatment. UF/MF simplifies the pre-treatment system, and produces better filtrate quality, since the membrane barrier achieves better removals of fine colloidal particles than several stages of a conventional system (20, 21, 22). Submerged technology is more widely used for this application than pressurized due to the high feed solids loading.

The second major application is boiler feed water treatment. Many IMS systems have already been installed in China in recent years, due to the rapid development of the power industry and a desire by the Chinese to apply the latest technology. Most of these installations, which are small to medium size, use pressurized technology even though some of the feed water quality is poor. Industrialization in other parts of the World is likely to create a similar market.

Municipal surface water applications form an important sector of IMS applications. Even if conventional treatment could be used for the RO, membranes are often favoured due to the integrity requirements of the overall treatment system. As discussed previously, it is difficult to monitor RO satisfactorily for integrity, and even if monitored, the individual elements at fault cannot be repaired.

The fourth application area is seawater. This is a newly emerging area, with the first large size IMS seawater system (26,500 m³/day, 7 MGD) currently...
being commissioned in Saudi Arabia (23), and some further large projects under contract. Both pressurized and submerged technologies are likely to feature in future installations dependent upon project size (24, 25).

**Submerged Membrane Process Design.** Figure 8.14 shows a schematic flow diagram of a seawater RO system with membrane pretreatment. In this diagram, the membrane pretreatment is represented by submerged vacuum driven technology, which is expected to be future technology of choice for large seawater desalination systems.

In the submerged system, fibers are potted together in the form of bundles or cassettes, collected together to form a module (Fig. 8.15). The module has two water collecting manifolds, enabling filtrate to be withdrawn from and backwash flow to be provided to both ends of the bundles. The dual manifold enables lower pressure drop inside the lumen during filtration and backwash.
steps. A number of modules connected in parallel form a stack or block (Fig. 8.16). The membrane blocks are immersed in feed water tanks and connected to the vacuum-backwash manifold (Fig. 8.17). Feed water is continuously fed to the tank with about 95% of being removed as filtrate. The remaining 5% is bled as a waste stream from the bottom, loaded with suspended solids. Additional piping provides air for scouring of fibers.

FIG. 8.16  Submersible capillary module assembly (Courtesy Zenon Corporation).

FIG. 8.17  Design concept of submersible system (Courtesy Zenon Corporation).
The operational sequence consists of filtration step followed by a filtrate backwash. The filtration step comprising suction of filtrate under vacuum of 0.07–0.5 bar (1–7 psi), lasts for 15–30 min. The filtration rates applied are 20–70 l/m²/h (12–41 gfd). During the filtration step, the surface of the fibers is scouried with air, released at the bottom of the stack to shake fibers and remove foulants. Air is provided in either a continuous or an intermittent mode. At the end of filtration interval, filtrate flow is reversed for a period of about 1 min. During the backwash the filtrate flows under pressure, applied from the lumen of the fiber at a flux rate somewhat higher than the filtration flux rate. This backwash flow of filtrate unblocks the pores and lifts deposits from membrane surface, restoring permeability.

Once to several times per day, cleaning and/or disinfecting chemicals are added to filtrate during the backwash. The duration of this chemical enhanced backwash (CEB) is longer than regular backwash as it usually includes a soaking period of 5–20 min. The chemicals added during CEB may include chlorine (or hydrogen peroxide), caustic and/or acid. Usually these chemicals are added separately. However, sometimes chlorine is added together with the caustic. If the water treated has high hardness, after CEB with hypochlorite and caustic, a backwash with acid is required to dissolve the scale formed. The CEB is applied in the same direction as backwash, from lumen (inside of the fiber) out to the tank. Periodically, when the TMP increases above 0.5 bar (7 psi) cleaning in place (CIP) is conducted. During CIP, the cleaning solution is applied from the same side as the feed water. The tank is drained and filled with cleaning solution prepared with soft water. The cleaning chemicals are the same as used during CEB, but potentially at higher concentration, and occasionally with addition of detergents. The CIP may include a few hours of soaking, preferably with warm (30–40°C) cleaning solution.

The sequence of operation of the submerged unit is listed in Table 8.2 and shown schematically in Fig. 8.18.

As indicated in Table 8.2, the submersible system has to be designed to produce a gross capacity of 10–25% higher than the nominal filtrate output. The additional capacity is necessary to compensate for off line time (backwash, cleaning and other maintenance) and the volume of filtrate required for backwash. The operating parameters that strongly affect the process economics are design filtrate flux rate, operating intervals between backwash, and frequency of cleanings. The design filtrate flux rate determines the membrane area (number of membrane modules) required to produce sufficient filtrate capacity. The length of operating intervals between backwash and the frequency of cleaning affects the on-line
**FIG. 8.18** Flow diagram of submersible capillary membrane plant.

**TABLE 8.2**
Operation sequence of submersible system

<table>
<thead>
<tr>
<th>Process step</th>
<th>Objective</th>
<th>Duration</th>
<th>Frequency</th>
</tr>
</thead>
<tbody>
<tr>
<td>Vacuum filtration</td>
<td>Filtrate production</td>
<td>15–60 min</td>
<td>Continuous</td>
</tr>
<tr>
<td>Pressure backwash</td>
<td>Foulants removal—maintaining permeability</td>
<td>30–60 s + 30 s for valves adjustment</td>
<td>Every 15–60 min</td>
</tr>
<tr>
<td>Chemical enhanced backwash (CEB)</td>
<td>Foulants removal—permeability restoration</td>
<td>1–20 min</td>
<td>Once—few times per day</td>
</tr>
<tr>
<td>Cleaning in place (CIP)</td>
<td>Foulants removal—permeability restoration</td>
<td>2–4 h</td>
<td>Every 1–6 months</td>
</tr>
<tr>
<td>Integrity test</td>
<td>Verification of capillary membranes integrity</td>
<td>20 min</td>
<td>Every 1–30 days</td>
</tr>
<tr>
<td>Off line time</td>
<td>60–180 min (4–12%)</td>
<td>Additional capacity required for off line time + backwash</td>
<td>10–22%</td>
</tr>
</tbody>
</table>
time factor of the plant. However, the major economic effect of these operating parameters is related to cost of chemicals, required for CEBs and CIPs.

Example 15
Vacuum driven capillary system
Filtrate output capacity: 37,850 m³/d (10 mgd)
Feed water turbidity: average 1 NTU, maximum 5 NTU
Feed water temperature: 16–28°C (61–82°F)

Design filtrate flux: 27 l/m²/h (16 gfd)
Backwash flux: 82 l/m²/h (48 gfd)
Backwash frequency: 30 min
Backwash duration: 1.0 min
Off line time due to backwash: 1.5 min
Integrity test frequency: every 7 d
Integrity test duration: 20 min
CEB frequency per day: 1
CEB duration: 20 min
Cleaning frequency: every 30 d
Cleaning duration: 4 h (8 min/d)
Membrane area per module: 200 m² (2170 ft²)
Number of modules per block: 10

Total backwash duration, min/day
1.5(1440 – 20/7 – 20)/30 + 1.5 = 1.5 × 45 = 68
On line time, min/day
1400 – 20/7 – 20 – 68 – 8 = 1341 (93%)
Filtrate capacity required
Backwash capacity
(45/1440)(82/27) = 0.10 (10%)
Total filtrate capacity
1/0.93 + 0.10 + 0.05 (contingency) = 1.22
3785 0 × 1.22 = 46,177 m³/d (12.2 mgd)
Membrane area required
46,177 × 1,000/(24 × 27) = 71,300 m²
12.2 × (1,000,000/16) = 762,500 ft²
Number of modules, each 200 m² (2170 ft²) = 360
Number of membrane blocks (10 modules) = 36
The pretreatment of feed water to a UF/MF submerged system consists of equipment that provides screening of particles that could otherwise damage polymeric fibers. This is usually accomplished with self cleaning strainer, rated at about 200 µ. In some systems, after the screening equipment, coagulant is added to the feed water. Then water flows into tanks where capillary modules are submerged. In large submerged systems, the tanks are of a similar configuration to gravity sand filters. However, the submerged membranes process produces a much higher output per tank unit volume compared to gravity sand filters. The footprint required for submersible membrane system depends very much on membrane module packing density, configuration and filtration flux. However, in most cases it is about 50–80% of conventional sand filtration system footprint for the same filtrate capacity (25).

The major equipment items in a submerged system include the filtrate pumps. The pumps draw water through the membranes and discharge it to the clear well. The filtrate pumps used are of a conventional type, that are able to operate under negative suction pressure of up to 1 bar (15 psi). During operation at negative suction pressure, some air dissolved in the water will be released. This air is accumulated in vertical tanks (air receivers) attached to the filtrate pumps and periodically purged from the piping manifold. Additional major equipment includes backwash pumps, sized for a pressure of about 3 bars (45 psi). The system also incorporates a significant number of automatic valves for isolation of membrane trains and providing water flow direction according to the operation sequences. System operation is managed through a central PLC that receives input from flow, pressure, temperature and water quality (turbidity and/or particle counters) sensors, and controls operation of pumps and valves.

Pressurized membrane technology. In pressure driven capillary membranes, feed pressure of up to 2 bar (29 psi) is applied to the feed water to create water flow through the MF/UF membranes. An example of one module configuration is shown in Figs. 8.19 and 8.20. The module diameter is about 225 mm (9") and the length is 1000 mm or 1500 mm. The membrane area of the 1000 mm long module is 30 m² (325 ft²) and 46 m² (500 ft²) for the 1500 mm long model. In field applications for RO pretreatment the module operates in a flux range typically between 60 – 100 l/m²/h (35–60 gfd), though higher fluxes are achievable for low solids feeds.

The module has three ports: feed, concentrate and filtrate. The mode of operation is very similar to the submerged system. There is alternate sequence of filtration followed by backwash steps. The backwash step is conducted at a pre-
determined frequency, sometimes with the presence of disinfectant. Periodically or whenever trans-membrane pressure (TMP) increases above the preset limit (usually 1.1 bar, 16 psi), a CIP is conducted. During the filtration step, which lasts for 15–60 min., the feed port is opened and the concentrate port closed, so that the module operates in a direct filtration mode (100% recovery). Feed flows under pressure to the lumen of capillaries, permeates through the capillary walls and is collected in central filtrate tube. The filtrate leaves the module through the permeate port. In the backwash step, the filtrate flow direction is reversed, flowing under pressure from the filtrate tank, through the filtrate port to inside
of the module and then through the capillary walls into the lumen. The backwash water leaves the module through either feed or concentrate port. The flow diagram illustrating unit operation is shown in Fig. 8.21.

The pressurized units are configured as a horizontal or vertical parallel array of modules, connected to a common manifold. An example of pressure driven ultrafiltration unit is shown in Fig. 8.22. At present, most pressurized systems are UF, with the main application being potable water production.
Membrane integrity verification. The unique premise of UF and MF technology is that it provides a membrane barrier and, potentially, an almost absolute removal of particles down to a submicron range. It is not surprising therefore that verification of membrane integrity is an important issue for these two technologies. Integrity verification methods for membrane systems are described in ASTM method (40). Some of integrity verification tests can be applied only when the system is off line, whilst others can be conducted during system operation.

1. Off line tests
   - Bubble point test
   - Pressure hold test
   - Diffusive air flow test
   - Vacuum hold test

2. Continuous (on line) tests
   - Particle passage counting/monitoring
   - Marked particle passage
   - Turbidity measurements
   - Acoustic sensing

The most frequently used integrity verification method is the pressure hold test. In this method water is purged from the system using clean, oil free air. Then air pressure at 0.3–1 bar (5–15 psi) is applied to the feed side of the membrane. After the feed and concentrate ports are closed, the decay of air pressure is monitored. The decay of applied air pressure is a result of air diffusion through the membrane and air flow through the defects. The integrity of the membrane filtration device is adequate if the pressure decay is typically less than 10% of the initial pressure within a period of 5 min (the decay is dependent upon membrane, module and pipework hold up). The ability to differentiate between diffusive flow and air flow through the broken fibers is a function of number of modules or membrane area in the system segment that is being tested for integrity. The number will vary for different membrane modules but the limit corresponds to a membrane area of about 1000–2000 m² (11,000–22,000 ft²), depending on the level of integrity being established. If the pressure decay is faster than expected for diffusive flow through the capillary walls, modules with defective fibers are located by pressurizing individual modules and looking for air bubbles. The pressure at which air will pass through membrane defects is related to diameter of the defects according to equation:

\[ P = \frac{4k\gamma\cos\theta}{d} \]  

(27)
where $P = \text{bubble point pressure}$

$k = \text{correction factor for the pore irregularity}$

$\lambda = \text{surface tension of the fluid (water)}$

$\theta = \text{contact angle}$

$d = \text{pore diameter}$

For hydrophilic membranes (contact angle = zero), round pores and a surface tension of water of 72 dynes/cm, the pore diameter ($d$) determination equation can be simplified to:

$$d = \frac{288}{\Delta P_{\text{max}}}$$  \hspace{1cm} (28)

In actual application, the bubble shows at much lower pressure than calculated by Eq. 27. This may be due to the irregular shape of membrane defects and incomplete wetting of pores (indicating a partial hydrophobic character of membrane material). Nevertheless, the pressure decay test and bubble test is probably the most effective method of membrane integrity determination. Since the diameter of capillary fibers is very low, a small number of broken fibers will have only moderate effect on membrane integrity, as shown in the example 16. This example shows calculations of water flow through a fiber that was cut in the middle. This flow is compared with the total filtrate flow. The number of capillary fibers in one module is quite large: 10,000–20,000 fibers. Therefore, the flow through one cut fiber is very low compared to overall filtrate flow from the module. In addition, because the internal diameter of the capillary is small, it is very likely that the cut fiber will be quickly plugged by particulate matter in the water. For this reason, a few membrane defects will have only a marginal effect on the quality of RO feed water produced in the membrane pretreatment system. Still, conducting an integrity test periodically is important since it provides an indication of the overall condition of the capillary membranes in the system.

**Example 16**

Integrity evaluation

Module membrane area: 46m$^2$, fiber length 1.25m, ID 0.8 mm
Number of fibers in one module: $46/(1.20 \times 3.14 \times 0.8 \times 10^{-3}) = 15,200$

Calculation of bypass flow through one broken fiber at TMP = 50kPa.

$Q = pd4\text{TMP}/(128Lm)$

$Q = 3.14(0.8 \times 10^{-3})4 \times 50/(128 \times 1.62 \times 10^{-3}) (1/0.65 + 1/0.65) \times 1000 \times 3600$
\[ Q = 3.43 \times 10^{-3} \text{ l/h} \]
Filtrate flow of element @ 85 l/m²h = 46m² × 85 l/m²h = 3910 l/h
Log of flow ratio = log(3910/3.43 \times 10^{-3}) = 6.06

### TABLE 8.3
Pore diameter and corresponding bubble point pressure.

<table>
<thead>
<tr>
<th>Pore diameter, micron</th>
<th>Bubble pressure, bar (psi)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.05</td>
<td>57.6 (835)</td>
</tr>
<tr>
<td>0.10</td>
<td>28.8 (418)</td>
</tr>
<tr>
<td>1.00</td>
<td>2.88 (42)</td>
</tr>
</tbody>
</table>

### TABLE 8.4
Effectiveness of particles rejection vs. number of broken fibers

<table>
<thead>
<tr>
<th>No. of broken fibers in element</th>
<th>1</th>
<th>2</th>
<th>5</th>
<th>10</th>
<th>100</th>
</tr>
</thead>
<tbody>
<tr>
<td>Log removal</td>
<td>6.1</td>
<td>5.8</td>
<td>5.4</td>
<td>5.0</td>
<td>4.0</td>
</tr>
</tbody>
</table>
Chemical stabilization of permeate

The permeate produced in brackish and seawater RO systems contains low concentration of alkalinity and hardness. Its buffering capacity is low and CO₂ presence results in decreased pH. Such water if sent directly, without treatment, to the distribution piping would have tendency to dissolve the protective calcium carbonate deposit from piping walls and would initiate corrosion. Therefore, RO permeate has to be stabilized by increasing concentration of alkalinity and hardness at the exit from the desalination plant. If RO permeate originates from anaerobic feed water source, then residual H₂S present in the permeate water has to be removed as well, prior to entrance of water to the distribution network. The common method of H₂S removal is aeration, conducted at permeate water pH lower than 6.5. Effectiveness of H₂S removal by air stripping is very high. Any residual H₂S after aeration can be converted to sulfuric acid by oxidation using chlorine gas or hypochlorite solution, according to reactions:

\[ \text{H}_2\text{S} + 4\text{Cl}_2 + 4\text{H}_2\text{O} = \text{H}_2\text{SO}_4 + 8\text{HCl} \]  \hspace{1cm} (29a)  
\[ \text{H}_2\text{S} + 4\text{NaOCl} = \text{H}_2\text{SO}_4 + 4\text{NaCl} \]  \hspace{1cm} (29b)  

Oxidation with hydrogen peroxide yields predominantly elemental sulfur according to equation:

\[ \text{H}_2\text{S} + \text{H}_2\text{O}_2 = \text{S} + 2\text{H}_2\text{O} \]  \hspace{1cm} (29c)  

In brackish RO systems, in absence of H₂S, the permeate stabilization process includes mainly removal or neutralization of CO₂. Removal of CO₂ is effectively accomplished by aeration using either forced draft or induced draft
degasifiers. Both degasifier types have capability to effectively reduce CO₂ concentration below 10 ppm.

If increase of alkalinity concentration is required for water stabilization, it could be achieved by addition of calcium hydroxide, calcium carbonate or sodium hydroxide to permeate water containing sufficient concentration of CO₂ (64).

$$2\text{CO}_2 + \text{Ca(OH)}_2 = \text{Ca(HCO}_3\text{)}_2$$  \hspace{1cm} (30)

$$\text{CO}_2 + \text{CaCO}_3 + \text{H}_2\text{O} = \text{Ca(HCO}_3\text{)}_2$$  \hspace{1cm} (31)

$$\text{CO}_2 + \text{NaOH} = \text{NaHCO}_3$$  \hspace{1cm} (32)

$$\text{CO}_2 + \text{Na}_2\text{CO}_3 + \text{H}_2\text{O} = 2\text{NaHCO}_3$$  \hspace{1cm} (33)

Seawater feed has low alkalinity (about 150 ppm). Therefore, permeate from seawater RO system has both low alkalinity and low CO₂ concentration. Due to very high rejection rate of seawater membranes, concentration of calcium in permeate stream is very low as well, in the range of 1–3 ppm. The objective of permeate treatment is to stabilize water composition by increasing concentration of calcium and alkalinity and increasing pH to achieve positive value of Langelier Saturation Index (LSI). In majority of seawater systems concentration of CO₂ in permeate is too low to form sufficient concentration of alkalinity to achieve positive LSI. Therefore, concentration of CO₂ has to be increased by injecting CO₂ gas into the permeate water prior to addition of caustic soda or lime. The chemicals added for permeate stabilization result in TDS concentration increase. Table 9.1 summarizes concentration changes due to addition of common chemicals used for permeate stabilization.

As shown in Table 9.1 the chemical system most effective in increasing hardness and alkalinity is one that involves reaction of CO₂ with CaCO₃. The decision which chemicals to use is mainly driven by the local availability and cost. In cases when only increase of hardness is required the simplest approach

<table>
<thead>
<tr>
<th>Reagent</th>
<th>Usage</th>
<th>Compound formed</th>
<th>Hardness increase</th>
<th>Alkalinity increase</th>
<th>TDS increase, ppm</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ca(OH)₂</td>
<td>0.84</td>
<td>Ca(HCO₃)₂</td>
<td>1.13</td>
<td>1.13</td>
<td>1.84</td>
</tr>
<tr>
<td>CaCO₃</td>
<td>2.27</td>
<td>Ca(HCO₃)₂</td>
<td>2.27</td>
<td>2.27</td>
<td>3.27</td>
</tr>
<tr>
<td>NaOH</td>
<td>0.91</td>
<td>NaHCO₃</td>
<td>0.0</td>
<td>1.13</td>
<td>1.91</td>
</tr>
<tr>
<td>Na₂CO₃</td>
<td>2.40</td>
<td>NaHCO₃</td>
<td>0.0</td>
<td>2.27</td>
<td>3.40</td>
</tr>
</tbody>
</table>
is to react calcium carbonate with sulfuric acid, produce CaSO₄ solution and add it to the permeate to achieve sufficient calcium concentration. During process design, the increase of permeate TDS due to permeate post-treatment, has to be taken into consideration in calculation of resulting final products salinity.

In majority of locations potable water in distribution systems has to contain residual of a disinfectant: free chlorine, chloramines or chlorine dioxide. The concentration level of disinfectant is determined by the Contact Time (CT), expressed in mg/l-min. For example CT = 100 could be concentration of free chlorine of 5 mg/l and residence time prior to point of use of 20 min. The CT required for a given deactivation credit for viruses and pathogens is significantly higher for chloramines then for free chlorine. The relations for CT and deactivation credits could be found in USEPA Guidance Manual (30) or other relevant local regulations.
RO/NF system design parameters

10.1. Feed water types

The composition and quality of water, considered for processing by reverse osmosis is influenced by its origin. For reverse osmosis application water of interest is the one with ions composition that exceeds potable water limits. In brackish water RO applications the ions that commonly exceed potable water limits are calcium, magnesium, sulfates and chlorides. Less common dissolved constituents of brackish water, that may require reduction of concentration, are fluoride and nitrate. Some waters may contain also excessive concentrations of iron, manganese, organic matter, color, hydrogen sulfite and sometimes even radioactive isotopes.

Potable water limits are specified by World Health Organization (26) and regional health authorities. Accordingly, one may define brackish water as any water of composition of soluble species exceeding potable water limits. The potable water limits, or acceptance of water composition for potable application can vary from country to country, according to local affordability of treatment methods. However, it is commonly accepted that water of salinity exceeding 1000 ppm is considered as brackish and requires treatment for salinity reduction.

The upper limit of water salinity that can be effectively treated with brackish RO membranes, in a single pass configuration, is about 10,000 ppm. On the low end of salinity spectrum there are some water sources that have salinity in the potable range but still require membrane treatment. This is usually due to presence of excessive concentration of hardness, iron, organics and/or color. The low salinity water sources are usually treated with loose RO membranes, commonly called as softening or nanofiltration membranes.
presence of pesticides at very low concentration in otherwise potable quality waters, is case of concern and could require membrane treatment. The ideal membrane for such application should have high rejection of organic pesticide and preferably very high passage of all dissolved ions. This also would be a nanofiltration application.

The composition of brackish water can vary widely. The composition is usually specific for the aquifer it originates from. If brackish water aquifer is very large and/or water is pumped at the rate it is being replenish by natural infiltration, then the composition remains stable. In case of excessive pumping the composition may change. In case of utilization of coastal aquifer there is possibility of seawater intrusion and salinity increase. For other locations there will be influence on composition from adjacent underground bodies of water due to hydrostatic pressure difference.

Seawater sources are characterized by high salinity, in the range of 30,000 ppm TDS to 47,000 ppm TDS. The ion composition includes mainly sodium and chloride, about 85% combined. The remaining fraction consists of sulfate (∼8%), magnesium (∼4%), calcium, potassium (∼1.2% each) and bicarbonate (∼0.6%). Boron is one of low concentration constituents. It is present in seawater at concentration of about 3–5.5 ppm. However, boron concentration is becoming increasingly important parameter of the process design as its concentration is being specified in RO permeate. Due to relatively low rejection by RO membranes of boron species existing in seawater, stringent boron specifications have significant effect on process design and product water cost. At majority of locations the ions composition of seawater is quite consistent and fluctuates in narrow range. Temperature of seawater usually reflects the seasonal fluctuations of ambient temperatures but could be affected by temperature of local water currents. At some locations, where rivers discharge or rain surface run off is present, the fluctuations of salinity and concentration of suspended matter could vary in a wide range and require careful consideration during the process design stage.

As a part of the design process of RO plant the feed water sources should be evaluated to determine the following:

1. How feed water ions composition and temperature will affect quality of permeate and the required feed pressure?
2. Does water source contain sparingly dissolved species at concentration that could result in membrane scaling at the design recovery rate?
3. Does water source contain particulate matter that could plug feed channels of membrane elements?
4. Does water source contain organic matter at concentration that could adsorb on membrane surfaces and result in significant permeability decline?

5. What is the level of biological activity? Can it result in biofouling of membranes?

Preliminary determination of the suitability of given water source for RO processing and requirements of the pretreatment process, is conducted based on results of analysis of water samples from the water source under consideration. A listing of typical water parameters being analyzed is provided in Fig. 6.1.

During the initial evaluation of water analysis it is important to check if the analysis report contains values of important water quality parameters and concentration of major ions. The primary group of water composition data includes: pH, temperature, turbidity, electric conductivity and concentrations of Ca, Mg, Na, K, HCO₃, SO₄, Cl and SiO₂. Additionally, concentration of any species that their maximum concentration had been defined in permeate, should be also determined in the feed water source, for example concentration of NO₃ or B. In some cases, mainly in specifications of nanofiltration projects, minimum concentrations of hardness in permeate is being listed as system performance requirements. The analysis should be balanced, i.e., sum of miliequivalents of positively charged ions (cations) should be similar to the sum of equivalents of negatively charged ions (anions). Data entry screens of majority of membrane performance projection computer programs converts concentration in ppm to miliequivalents and display the results automatically. If the difference of corresponding sums of positive and negative miliequivalents is within ±5%, the analysis is being considered as balanced. Some computer programs include option for automatic balance of negative or positive ions concentrations as necessary.

10.2. Feed water composition

Based on the feed water salinity range a suitable membrane element type is selected for performance calculations (i.e., NF, RO or seawater). The feed water composition and temperature, together with recovery rate and average flux rate, are used as an input parameters in calculations of projected permeate quality and feed pressure. At given operating parameters and membrane type the permeate concentration is direct function of feed salinity (Table 10.1 shows representative values of ion passage in RO systems treating low salinity and high salinity brackish feed and medium salinity range of seawater feed). However,
careful examination of experimental data for brackish composite polyamide membranes shows that there is anomaly of salt passage at very low and very high feed salinities (27). It is shown in Fig. 10.1 that for brackish RO membranes there is a minimum of salt passage at feed salinity of about 500 ppm. In the narrow range below this salinity, a steep increase of salt passage is observed. There is also increase of salt passage at higher feed salinities, however it is at a lower rate as compared to the very low salinity range. Eventually, at high feed salinities of about 5,500 TDS ppm the salt passage levels off. This effect of salt passage increase should be consider in calculations of permeate salinity. It is quite important in the low salinity range for ultrapure water applications such as production of high pressure boilers make up and process water for semiconductor industry.

The salt passage increase with increased salinity occurs in almost all RO brackish water systems. Even in systems operating with low feed salinity, due to the effect of recovery, the elements positioned close to the concentrate end, are exposed to high salinity water. For seawater membranes, in the common feed salinity range, the effect of salinity on salt passage is not as significant as for brackish systems and is usually neglected in permeate salinity calculations.

In brackish applications salt passage is also affected by the ion composition of the feed water. In the process of membrane elements manufacturing, salt rejection is determine using single salt solution: NaCl. In field application feed water is composed of variety of ions, which diffuse through the membrane at

![FIG. 10.1](image)  
**FIG. 10.1**  Effect of feed salinity on osmotic pressure and required feed pressure in RO unit.
different rates. In addition membrane surface of polyamide membranes is negatively charged which contributes to salt rejection through repulsion of negatively charged ions. When treating a solution of mixed ions composition the ion passage is affected by counter ions charge balance (Donnan effect). Therefore presence of highly mobile ions (like $\text{HCO}_3^-$) will increase passage of positively charged ions, contributing to increasing salt passage. Conversely, significant concentration of large, slowly diffusing ions (like $\text{SO}_4^{2-}$) will retard passage of associated cations and resulted in lower permeate salinity. In addition, high concentration of ions with high positive charges, such as calcium and magnesium, will tend to neutralize negative membrane surface charge, reducing ions rejection that is due to surface-anions charge repulsion. Due to above effects, salt passage of brackish membranes in field conditions could differ significantly from the nominal values, obtained during testing with solution of NaCl. The effect of feed water composition on salt passage is most pronounced in softening (nanofiltration) membranes. The salt rejection of nanofiltration membranes is low and most of them have high surface charges. Therefore, exposure to mixed ion solution with high fraction of concentration of divalent cations will result in significant salt passage increase. In seawater applications salt rejection of seawater

<table>
<thead>
<tr>
<th>Feed constituent</th>
<th>Low salinity brackish</th>
<th>Medium salinity seawater</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Feed, ppm</td>
<td>Permeate, ppm</td>
</tr>
<tr>
<td>Ca</td>
<td>83</td>
<td>0.6</td>
</tr>
<tr>
<td>Mg</td>
<td>66</td>
<td>0.5</td>
</tr>
<tr>
<td>Na</td>
<td>350</td>
<td>11.5</td>
</tr>
<tr>
<td>K</td>
<td>4</td>
<td>0.2</td>
</tr>
<tr>
<td>HCO$_3^-$</td>
<td>340</td>
<td>10.6</td>
</tr>
<tr>
<td>SO$_4^{2-}$</td>
<td>74</td>
<td>0.3</td>
</tr>
<tr>
<td>Cl</td>
<td>600</td>
<td>10.5</td>
</tr>
<tr>
<td>F</td>
<td>0.8</td>
<td>0.03</td>
</tr>
<tr>
<td>NO$_3^-$</td>
<td>46</td>
<td>5.6</td>
</tr>
<tr>
<td>B</td>
<td></td>
<td></td>
</tr>
<tr>
<td>SiO$_2$</td>
<td>29</td>
<td>0.5</td>
</tr>
<tr>
<td>TDS</td>
<td>1563.8</td>
<td>39.83</td>
</tr>
</tbody>
</table>

TABLE 10.1
Illustration of potential feed concentration reduction in an RO systems

(1) For the purpose of illustration of potential concentration reduction, passage is being calculated relative to the concentration in the feed stream. These values of ions passage are higher than the usual values calculated relatively to the average feed (average of feed and concentrate).
elements correlates well with the nominal values. This is most likely because the feed water is composed mainly of NaCl (~85%) and the membrane barrier of seawater membranes have high level of polyamide polymer crosslinking (lower density of surface charges compared to NF membranes). Feed salinity affects the required feed pressure through the value of average feed osmotic pressure (Eq. 8). In multistage brackish water systems increased osmotic pressure and pressure drop reduces NDP in each subsequent element. Consequently, permeate flux in elements in tail positions (see Figs. 84, 85) could be quite low. Therefore, feed pressure requirements for a given permeate capacity has to be adjusted according to the feed salinity. The effect of salinity on required feed pressure demonstrates itself both in brackish and seawater systems (Fig. 10.2).

10.3. Feed water temperature

Feed water temperature affects both permeate salinity and the required operating pressure. Both water and salt transport follow a similar trend (Eq. 16, Fig. 2.5). The commercial RO systems are designed to operate at constant permeate capacity (constant permeate flux). Therefore, increase of feed water temperature will result in increased permeate salinity (higher quantity of salt will pass the membrane and will be dissolved in a constant volume of permeate). This increase is about 3% per degree C. The effect is similar for brackish and seawater systems (Figs. 10.3 and 2.6). The changes of water permeability with temperature affects the net driving pressure required (Eq. 18). However, in par-
allel, temperature also affects osmotic pressure of the feed water (Eq. 2). With temperature changes net driving pressure and osmotic pressure have opposite effect on feed pressure (Eq. 8). An increase of feed water temperature at low temperature range (~up to about 30°C) enables production of a given permeate flow at reduced feed pressure both in seawater and brackish water systems. At higher temperatures the reduction of feed pressure in seawater systems levels off, mainly due to increase of osmotic pressure of the average feed (Figs, 10.4 and 10.5).

**FIG. 10.3** Effect of feed water temperature on required feed pressure in a brackish RO unit. Average permeate flux 28 l/m²/h.

**FIG. 10.4** Effect of feed water temperature on required feed pressure in a brackish RO unit. Average permeate flux 28 l/m²/h.
10.4. Sparingly soluble constituents

During the RO process concentration of all constituents increases due to reduction of the feed water volume. This increase of concentration is function of permeate recovery (Eqs. 6, 7). Some of the constituents presented in natural waters can precipitate if the concentration product of salt forming ions exceeds its solubility product: $K_{sp}$. For a given salt of composition $C_mA_n$ in equilibrium of solid phase salt (S) with dissolved ions, the $K_{sp}$ is defined as:

$$C_mA_n(S) = mC^{+n} + nA^{-m}$$  \hspace{1cm} (34)

$$K_{sp} = [C^{+n}]^m [A^{-m}]^n$$  \hspace{1cm} (35)

$$SI = [C]^m [A]^n / K_{sp}$$  \hspace{1cm} (36)

Where $C$ stands for cation and $A$ for anion, $m$ and $n$ are valency coefficients. Brackets [ ] indicate molar concentration of a given ion in solution. SI is the saturation index, indicating excess concentration of a given salt in comparison to its saturation value.

The $K_{sp}$ is determined through measurement of ions concentrations in solution at saturation conditions (in equilibrium with solid phase). $K_{sp}$ value is specific for a given salt and it is function of temperature and ionic strength of the solution. In brackish systems, treating natural waters, the salt of concern is mainly calcium carbonate. Less frequently calcium sulfate and silica are at concentrations that my result in scale formation. In very rare cases barium sulfate,
ferrous sulfide, and ferrous carbonate could be present at concentrations that may form scale at high recoveries. In RO systems treating municipal effluents calcium phosphate sometimes forms in the tail elements. Calcium carbonate is the most common scaling constituent in brackish waters but also the easiest to control either with pH adjustment or use of scale inhibitor. In solution, calcium ions are in equilibrium with bicarbonate and carbonate species as shown in the following equations:

\[
\begin{align*}
\text{OH}^- + \text{H}^+ &= \text{H}_2\text{O} \\
\text{H}_2\text{CO}_3 &= \text{H}^+ + \text{HCO}_3^- \\
\text{HCO}_3^- &= \text{H}^+ + \text{CO}_3^{2-} \\
\text{Ca}^{+2} + \text{CO}_3^{2-} &= \text{CaCO}_3
\end{align*}
\]

The calcium carbonate system is quite complex. Saturation conditions are not just function of concentrations of Ca, CO_3 and HCO_3 ions but also influenced by concentration of hydrogen ion (pH). Attempts to define relations for saturation conditions in potable water networks lead to development of number of saturation indexes. The calcium carbonate saturation index developed by Langelier (31) for potable water networks has been adopted by RO industry as an indicator of saturation conditions in concentrate stream of brackish water RO systems. The Langelier Saturation Index (LSI) is calculated according to relations:

\[
\text{LSI} = \text{pH} - \text{pH}_s
\]

Where pH is the actual pH of the water and pH_s is pH that corresponds to saturation concentrations of ions forming calcium carbonate.

\[
\begin{align*}
\text{K}_2 &= [\text{H}^+] [\text{CO}_3^{2-}]/[\text{HCO}_3^-] \\
\text{K}_{sp} &= [\text{Ca}^{+2}] [\text{CO}_3^{2-}] = [\text{Ca}^{+2}] [\text{HCO}_3^-] \text{K}_2/\text{H}_s^+ \\
\text{H}_s^+ &= [\text{Ca}^{+2}] [\text{HCO}_3^-] \text{K}_2/\text{K}_{sp} \\
-\log[\text{H}_s^+] &= \text{pH}_s = -\log[\text{Ca}^{+2}] - \log[\text{HCO}_3^-] + \log \left[ \text{K}_{sp}/\text{K}_2 \right] \\
\text{LSI} &= \text{pH} - \text{pH}_s = \text{pH} - \text{pCa} - \text{pAlk} + \text{pK}
\end{align*}
\]

Where K_2 is second dissociation constant to carbonic acid (H_2CO_3), K_{sp}—solubility constant of calcium carbonate at given pH and temperature.
Other parameters represent molar concentrations of relevant species in the solution.

Few years after introduction by Langelier the saturation index was modified by Larson and Buswell (32) to account for ionic strength in a form of salinity correction factor:

\[
pH_s = (9.3 + A + B) - (C + D) \tag{47}
\]

where

\[
A = (\log_{10} [\text{TDS}] - 1)/10 \tag{48}
\]

\[
B = -13.12 \times \log_{10}(^\circ \text{C} + 273) + 34.55 \tag{49}
\]

\[
C = \log_{10} [\text{Ca}^{2+} \text{ as } \text{CaCO}_3] - 0.4 \tag{50}
\]

\[
D = \log_{10} [\text{alkalinity as } \text{CaCO}_3] \tag{51}
\]

The parameter \(A\) is related to ionic strength of the solution. The value of \(A\) increases with increase of salinity. Parameter \(B\) reflects the changes of calcium carbonate solubility and changes of equilibrium of carbonic acid dissociation with temperature. The value of \(B\) decreases with temperature increase. In practical applications the LSI is either calculated using computer programs or monograms based on pH and composition of concentrate stream. Water solution has potential for CaCO\(_3\) scaling at LSI > 0 and it is assumed that saturation prediction using LSI is reliable up to salinity of about 5000 ppm TDS.

For high salinity and seawater applications the LSI was modified to account for increased ionic strength by Stiff and Davis (33). The Stiff and Davis saturation index (SDSI) introduces empirical constant \(K\) in calculations \(pH_s\) to account for high ionic strength of seawater concentrates. Stiff and Davis determined value of \(K\) experimentally in the high range of ionic strength, that covers salinities encountered in RO seawater applications.

\[
\text{SDSI} = pH - (9.3 + K - pCa - pAlk) \tag{52}
\]

where \(pH\) has the same meaning as in LSI equation and \(K\) is a constant found in monograms.

For high salinity solutions (seawater concentrates) the SDSI value is about 1–1.3 units lower than calculated according to LSI relations.

**Example 17**

Calculation of Langelier saturation index

Brackish water system is design to operate at recovery ratio of 80%.
Feed water feed has TDS = 1700 ppm, ionic strength 0.04 and the following concentrations of the relevant ions: 
Ca = 300 ppm, HCO₃ = 250, pH = 7.3, temp = 22°C.
After acidification and pH adjustment to 6.5, HCO₃ = 166 ppm, CO₂ = 78 ppm
Concentration factor = 1/(1 – \( R_\text{p} \)) = 5
Approximate concentrations in the concentrate:
TDS = 8500 ppm, ionic strength 0.18, Ca = 1500 ppm, HCO₃ = 830 ppm, CO₂ = 78 ppm

Calculation of concentrate pH.
\[
pH = \text{pK}_1 + \log\left(\frac{[\text{HCO}_3]}{[\text{CO}_2]}\right)
\]
\( \text{pK}_1 \) is the equilibrium constant of the carbonic acid dissociation reaction.
At 25°C its value is 4.2 \times 10^{-7}, \( \text{pK}_1 = 6.37 \) (neglecting effect of water ionic strength on \( \text{K}_1 \)).
Concentrate pH = 6.37 + log((830/78)(44/61)) = 6.37 + 0.89 = 7.26

Calculation of \( \text{pH} \text{s} \) and LSI
Following Eqs. 41 and 47–51:
\[
\text{pH} \text{s} \text{ (feed)} = 9.3 + 0.22 + 2.15 – 2.48 – 2.13 = 7.06
\]
\[
\text{LSI} \text{ (feed)} = 6.5 – 7.06 = -0.56
\]
\[
\text{pH} \text{s} \text{ (concentrate)} = 9.3 + 0.29 + 2.15 – 3.17 – 2.83 = 5.74
\]
\[
\text{LSI} \text{ (concentrate)} = 7.26 – 5.74 = 1.52
\]

Calculation of SDSI
K value for ionic strength of 0.04 and 0.18 and, from monogram (41) is 2.4 and 2.8 for feed and concentrate respectively.

According to Eq. 50
\[
\text{SDSI} \text{ (feed)} = 6.50 – (9.3 + 2.4 – 2.48 – 2.13) = -0.59
\]
\[
\text{SDSI} \text{ (concentrate)} = 7.26 – (9.3 + 2.8 – 3.17 – 2.83) = 1.16
\]

Example 18
Seawater water system is design to operate at recovery ratio of 50%.

Feed water feed has TDS = 40,000 ppm, ionic strength 0.69 and the following concentrations of the relevant ions:
Ca = 450 ppm, HCO₃ = 150, pH = 8.1, temp = 22°C.
After acidification and pH adjustment to 6.5, HCO₃ = 115 ppm, CO₂ = 42 ppm.
Concentration factor = 2
Approximate concentrations in the concentrate.
TDS = 80,000 ppm, ionic strength 1.39, Ca = 900 ppm, HCO₃ = 230, CO₂ = 42 ppm, pH = 7.2
Calculation of concentrate pH
Concentrate pH = 6.37 + log((900/42)(44/61)) = 6.37 + 0.6 = 6.97

Calculation of pHs and LSI
Following Eqs. 41 and 47–51:
\[ \text{pH}_s \text{ (feed)} = 9.3 + 0.36 + 2.15 - 2.65 - 1.97 = 7.19 \]
\[ \text{LSI (feed)} = 6.5 - 7.19 = -0.69 \]
\[ \text{pH}_s \text{ (concentrate)} = 9.3 + 0.39 + 2.15 - 2.98 - 2.28 = 6.58 \]
\[ \text{LSI (concentrate)} = 6.97 - 6.58 = 0.39 \]

Calculation of SDSI
K value for ionic strength of 0.69 and 1.39, from monogram (41) is 3.3 and 3.5 for feed and concentrate respectively.
According to Eq. 50
\[ \text{SDSI (feed)} = 6.50 - (9.3 + 3.3 - 2.65 - 1.97) = -1.48 \]
\[ \text{SDSI (concentrate)} = 6.97 - (9.3 + 3.5 - 2.98 - 2.28) = -0.57 \]

As indicated by results included in examples 16 and 17, LSI and SDSI have similar values at low ionic strength. At high ionic strength of solution, the values of SDSI are becoming lower then LSI, due to the difference of corresponding values of K (Stiff and Davis) and A and B (Langelier). So far field results from seawater RO systems confirm lower tendency of calcium carbonate scaling as compared to brackish RO units.

The saturation relation of calcium sulfate is simpler to calculate than the one for calcium carbonate as its solubility depends only on concentrations of calcium and sulfate ions, temperature and ionic strength. The effect of ionic strength on solubility of calcium sulfate is quite significant as shown in the following example of calculations of saturation indexes (SI) for brackish and seawater RO systems:

**Example 19**
For CaSO\(_4\); \( K_{sp} = (\text{temp}/25)^{0.152} \times 1.8 \times 10^{-3} \times \text{IS}^{0.75} \)

**Brackish water system**
\( R = 75\%, \text{ TDS of concentrate} = 4000 \text{ ppm}, \text{IS} = 0.07 \)
\( \text{Ca} = 900 \text{ ppm}, \text{SO}_4 = 2400 \text{ (concentrations in concentrate)} \)
\( K_{sp} = 2.5 \times 10^{-4} \)
\( \text{SI} = ([900/40000][2400/96000])/2.5 \times 10^{-4} = 2.25 \)

**Seawater system**
\( R = 50\%, \text{ TDS of concentrate} = 80,000 \text{ ppm}, \text{IS} = 1.60 \)
Ca = 900 ppm, SO₄ = 6000 ppm (concentrations in concentrate)

\[ K_{sp} = 2.6 \times 10^{-3} \]

\[ SI = \frac{([900/40000] [6000/96000])}{2.6 \times 10^{-3}} = 0.54 \]

In the example above the product of calcium and sulfate ions concentrations in solution is much higher in concentrate stream of the seawater system as compared to the brackish system. However, due to the differences of ionic strength, the calcium sulfate is above saturation level in the brackish system and below saturation in the seawater system.

Another constituent that sometimes can present problem in RO applications is silica. In seawater sources the silica concentration is very low, just few ppm. However, in brackish water silica sometimes is present in significant concentrations. Silica can be present both in colloidal and reactive (soluble) form (61). In the past, the safe limit of silica in concentrate was considered as being about 140–170 ppm (as SiO₂). In the last decade a new scale inhibitors were introduced that are effective in maintaining much higher concentration of silica in solution. Some suppliers of these specialty inhibitors claim safe limits for silica concentration as high as 300 ppm. When treating brackish sources with significant silica concentration an extreme caution should be exercised with maintaining the designed recovery rate as silica scale is very difficult to remove.

Another potential scaling constituents, barium, although has very low solubility as barium sulfate, is much less common then silica. Events of barium sulfate scaling in RO systems were seldom reported in the past. Another scaling component of concern is calcium phosphate. Its presence is almost uniquely associated with municipal or industrial effluents at various concentrations. In majority of RO system treating municipal effluents concentration of phosphates is low and does not result in membrane scaling. At locations where phosphates concentration is high it still can be managed by using scale inhibitors or acidification. In extreme cases high concentration of phosphates can negatively affect economic feasibility of wastewater reclamation process due to high cost of acid required for pH adjustment.

The \( K_{sp} \) values (at 25°C) of common salts that could form scale in brackish RO systems are listed in Table 10.2 according to reference 58.

The potential for formation of calcium sulfate scale and blocking of feed channels in the spiral wound element is demonstrated in the following example:

**Example 20**

\[ [Ca] = 1000 \text{ ppm} = 0.025 \text{ mol} \]
\[ [SO₄] = 2400 \text{ ppm} = 0.025 \text{ mol} \]

\[ K_{sp} = 2.25 \times 10^{-4} \]
Saturation \([\text{Ca}] = [\text{SO}_4] = (2.25 \times 10^{-4})^{1/2} = 0.015 \text{ mol}\)

Excess \([\text{Ca}] = [\text{SO}_4] = 0.010 \text{ mol}\)

MW \(\text{CaSO}_4 = 40,000 + 96,000 = 136,000 \text{ mg/l}\)

Excess \([\text{CaSO}_4] = 0.010 \text{ mol} = 0.010 \times 136,000 = 1,360 \text{ mg/l}\)

RO system operates at 80% recovery rate. At average flux of 26 l/m²/h,

7 elements per vessel the concentrate flow in the last element is: \(26 \times 36.8 \times 7 \times (1 - 0.80)/0.80 = 1674 \text{ l/h}\)

Potential \(\text{CaSO}_4\) deposit = \(1674 \times 1.36 = 2276 \text{ g/h} = 2.28 \text{ kg/h}\)

Assuming specific density of \(\text{CaSO}_4\) = 4 g/cm³, volume of excess \(\text{CaSO}_4\)

that could precipitate is RO element is \((2.28 \times 1000)/4 = 570 \text{ cm}^3/\text{h}\).

Free volume of feed channels in SW element is:

100 cm × 100 cm × 0.075 cm × 20 = 15000 cm³ (about 50%)

The above calculations shows that at the saturation conditions scale could

lead in a short time to complete blockage of tail elements.

Potential for scale formation is extremely important issue in brackish applications. In some locations it may determine the maximum recovery rate for brackish RO systems. Due to variability of water compositions in brackish applications and limited level of understanding of the relevant salt solutions systems at saturation in RO conditions, it is quite difficult to make accurate predictions about scaling. RO industry adopted limits for individual salts based literature data and some field experience (Table 10.2). Due to lack of accurate analytical models, developed for RO applications, these limits include significant safety margins.

Manufacturers and suppliers of scale inhibitors are continuously introducing new scale inhibitors that enable operation at higher levels of concentrations then those initially proposed by membrane manufacturers. So far the experience

<table>
<thead>
<tr>
<th>Compound</th>
<th>Formula</th>
<th>(K_{sp}) (ppm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Calcium sulfate</td>
<td>(\text{CaSO}_4)</td>
<td>(2.5 \times 10^{-5})</td>
</tr>
<tr>
<td>Barium sulfate</td>
<td>(\text{BaSO}_4)</td>
<td>(2.0 \times 10^{-10})</td>
</tr>
<tr>
<td>Reactive silica</td>
<td>(\text{H}_4\text{SiO}_4)</td>
<td>((120 – 160))</td>
</tr>
<tr>
<td>Calcium carbonate</td>
<td>(\text{CaCO}_3)</td>
<td>LSI &lt; 0</td>
</tr>
<tr>
<td></td>
<td></td>
<td>S&amp;DSI &lt; 0</td>
</tr>
<tr>
<td>Calcium phosphate</td>
<td>(\text{Ca}_3\text{PO}_4)</td>
<td>(see comments in Chapter 17)</td>
</tr>
</tbody>
</table>
with the subsequent products introduced over the years was quite positive. Seldom any problems of scale formation could be related to malfunction of scale inhibitor if applied according to manufacturer specifications. Scale inhibitors prevent scale formation by retardation of nucleation process of scale forming crystals. The mechanism of prevention of crystal grow is either through threshold effect, crystal structure distortion, dispersion or sequestration. Dosing rate of scale inhibitor is determined by supplier of chemicals based on feed water composition and recovery rate. Feed water analysis should include information on concentration of iron. High concentration of iron containing compounds in concentrate may reduce effectiveness of some scale inhibitors. Required concentration of scale inhibitor seldom exceeds 10 ppm of active ingredient in the concentrate stream.

In seawater applications, due to typical seawater compositions and high ionic strength, potential for scaling is not a recovery limiting issue. In seawater systems recovery rate is limited by required operating pressure as discussed in the following chapter.

10.5. Particulate matter

To achieve high membrane packing density the RO membrane elements are configured with very narrow feed channels. In spiral wound elements the height of the feed channels is less than 0.8 mm (0.03”), filled with feed spacer net, which separates adjacent membrane surfaces and promotes turbulence. To avoid blockage of such narrow internal passage, RO feed water entering membrane elements, should have low concentration of particulate matter. The commonly accepted quality indicators of RO feed water in this respect include:

1. Turbidity
2. Suspended solids concentration
3. Silt density index (SDI)

Turbidity determination, usually expressed as nefelometric turbidity units (NTU) is determined through measurements of intensity of light scattered by suspended particles in water samples. Suspended solids concentration is determined by filtration of measured volume of water sample and weighting of dry residue on the filter. The SDI is determined through measuring the rate of filtration of water sample through the filter (Fig. 10.6). Determination of all three
above indicators is described in ASTM procedures (55, 54, 17). Out of above three indicators only turbidity can be measured continuously on line. The other two are conducted as discrete measurements on water samples taken periodically.

Another water quality parameter that is being used to monitor operation of pretreatment system is number of particles measured with particle counters. This could be applied as on line, semi-continuous measurement. Application of particle counters for RO application is still at very early stages. So far no relations had been reported as being established between particle counter measurements and fouling rate or performance stability of RO membranes.

The feed water indicator, most relied on in RO applications, is SDI. It is based on measurement of rate of declining flow, at a constant pressure, of a water sample through a porous filter membrane of nominal 0.45 µ porosity. The filter is placed in a simple circular holder (Fig. 10.6) and connected to feed water line at applied pressure of 2 bar (30 psi). The time measured for filtration of a constant volume (500 ml) at the beginning of the test \( t_1 \) and after 15 minutes \( t_2 \) is used to calculate the SDI according to the following equation:

\[
\text{SDI} = 100\% \left(1 - \frac{t_0}{t_2}\right) \left(\frac{1}{15}\right)
\]  

\[ (53) \]

If the filter plugs to fast for meaningful determination of the filtration time, the volume of filtrate being collected or time between measurements can be decreased. As shown in Fig. 10.6 (last entry in the table) it is possible to have long filtration times and still calculate low SDI values. If the reading for filtration time \( t_1 \) significant exceeds 30 seconds then most likely there is a problem with

\[
\begin{array}{|c|c|c|}
\hline
 t_0 & t_15 & \text{SDI (15 min)} \\ \hline
 18 & 22 & 1.2 \\ \hline
 20 & 33 & 2.6 \\ \hline
 26 & 73 & 4.3 \\ \hline
 46 & 50 & 0.5 (?) \\ \hline
\end{array}
\]
equipment or test conditions. The SDI method is very sensitive to concentration of foulants but it is not very accurate. No meaningful correlation has been established between values of SDI and turbidity.

Attempts to improve accuracy of this method led to introduction of Modified Fouling Index (MFI). The test for MFI is based on measurement of pressure increase required for maintaining of constant filtration rate through well defined membrane filter (28). The MFI results are more reproducible than SDI but the test is difficult to perform manually and automatic equipment is necessary (18) for a routine determination in the field conditions.

According to the Eq. 53, the maximum value of SDI (for 15 min measuring interval) can be only 6.67. The majority of membrane manufacturers specify upper limit for feed water SDI as 5. However, field results show that for stable, long term performance of RO elements, the SDI of feed water should be consistently below the value of 4. Some limited research works (34) demonstrated that in respect of solids concentration the SDI scale is a geometric one. Therefore, for water having SDI = 3 and another water having SDI = 5 the corresponding suspended solids concentration difference is about four times higher.

The following Figs. shows the SEM pictures of clean filter (Fig. 10.7) and two filters (Figs. 10.8, 10.9) after being used for SDI determination of seawater feed. Fig. 10.6 shows SEM picture of filter pad after SDI determination that resulted in SDI value of 2.2. Fig. 10.9 shows corresponding SEM picture for SDI value of 4.8. The tick deposit on the filter, which was used to filtrate water sample with higher SDI, is clearly visible.

Field results have demonstrated that in majority of cases water from deep wells has very low SDI, usually less than 1. RO systems, operating with good quality well water feed, practically do not show any pressure drop increase across the membranes or flux decline. Surface water, after a conventional pretreatment, usually has SDI in the 2–4 range. RO system processing feed water with SDI in the 2–3 range shows stable membrane performance. Membrane cleaning frequency for such feed water does not exceed 1–2 per year. RO systems processing feed water of higher SDI, in the 3–4 range, usually suffer from some degree of membrane fouling and somewhat higher membrane cleaning frequency may be required. Long term operation of RO system with feed water having SDI above 4 is not recommended. As mentioned already, past attempts to correlate turbidity with SDI values were not successful. These two feed water quality indicators correlates to the number and size of suspended particles in a different way. However, usually the feed water with SDI in the 2–3 range has corresponding turbidity below 0.1 NTU, usually at 0.05 NTU range.
FIG. 10.7 SEM picture of a clean SDI filter pad. Magnification $\times 2000$.

FIG. 10.8 SEM picture of a filter pad after SDI determination. SDI = 2.2. Magnification $\times 2000$. 

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10.6. Organic matter

Organic matter in RO feed is customary expressed as a total organic carbon (TOC). Surface water, water from shallow wells and municipal secondary effluent usually contains some concentration of dissolved organics. In surface waters the organic matter originates from decomposition of humic substances. In shallow well water the presence of organics could be result of water infiltration through strata containing natural organic matter. In both cases the TOC concentration is usually below 3 ppm. This low concentration of TOC in the fed water has little effect on membrane permeability. In RO feed originated from secondary effluent, the TOC concentration fluctuates in the wide range between 2–10 ppm. The presence of high concentration of organic matter results in flux decline due to adsorption on the membrane surface. The initial rate of flux decline, due to organic adsorption is rapid, about 10–30% decrease from the initial permeability value. Then the permeability stabilizes, and the decline levels off. However, if colloidal matter is also present in the feed water, the permeability decline is usually more severe, due to formation of thick, mixed foulants layer of low water permeability.
Some potential water sources for RO processing are contaminated by oil and grease. Such conditions could exist in industrial wastewater streams or in seawater in the areas of heavy maritime traffic. Oil and grease have to be completely removed from the feed water prior entering the RO membrane elements. Low concentration of oil and grease will not result in structural damage of the membrane element but will cause severe decline of water permeability. Conventional method of removing low concentration of oil and grease includes air flotation and/or media filtration with flocculation using metal based flocculants.

10.7. Biofouling

The assumption is that every body of water contains microorganisms in equilibrium with the local nutrients supply. The water born microorganisms easily attached themselves to the surfaces in the RO systems and form colonies. The attachment to the surfaces is through excretion of extra cellular polymeric substance (EPS), composed mainly of polysaccharides. On wetted surfaces, the microbiological cells and surrounding EPS form biofilm, that can grow at rapid rate if sufficient nutrients and energy are available. The structures of biofilms are not uniform, depending on type of microorganisms and environmental conditions: pH, temperature, flow velocity, age and variety of other parameters. The general knowledge on biofilms mechanism of formation and their control is still developing. More recent research approach, under umbrella of “biofilm ecology” utilizes genetic tests to monitor biofilms growth and study microorganisms diversity (37).

The phenomena of biofilms formation in RO systems (biofouling) has been recognized early and is well documented. The methods of effective prevention are continuously being developed. Majority of RO systems processing well water experience little or no biofouling. Mitigation of biofilm formations can be achieved through proper system configuration and operating conditions:

1. Elimination of stagnant areas in the pipe-network.
2. Direct hydraulic connection between well head(s) and RO system, without feed water storage tank.
3. Total prevention of exposure of feed water to light. All FRP piping and other plastic components of the system have to be completely opaque.
4. Avoiding addition of water treatment chemicals that can be metabolized by microorganisms.
5. Avoiding continuous chlorination of the feed water.
RO systems treating surface water (mainly seawater), are more susceptible to biofouling than brackish systems. In the past it was common procedure to chlorinate raw seawater through the pretreatment system to control bacterial population, and reduce free chlorine (using sodium bisulfite) prior to RO membranes. More recently it was established (35, 36) that chlorination/dechlorination approach actually has a potential of increasing biofouling in the membrane unit. The current understanding is that large size molecules of humic substances, that are present in the surface waters are not digestible by microorganisms. However, during chlorination of the feed water, free chlorine breaks them into smaller fragments of assimilable organic carbon (AOC), that can be metabolized by microorganisms. The microorganisms that survived chlorination will grow rapidly in subsequent sections of the system, past the dechlorination point. As the polyamide membrane elements can not be exposed to free chlorine, past the point of addition of chlorine reducing agent, no biocide is present in the feed water. In systems where chlorination/dechlorination is practiced, membrane surfaces provide good conditions for development of biofilm.

Following the above understanding of conditions leading biofilm formation, the current approach is to avoid all together chlorination of the feed water or practice intermittent chlorination/dechlorination only (dosing rate range: less than 1–5 ppm, duration 0.5–few hours per week). This approach restores the balance between the microorganisms population and availability of nutrients and usually results in low level of biological activity. One of important conditions, contributing to lower fouling rate is selection of intake location that will provide clean water, with low concentration of foulants and low level of biological activity. Once biofilm has been formed in the RO system it is extremely difficult to restore it to clean conditions.

The general approach to mitigate microorganisms grow in RO system is to periodically change drastically the environmental conditions (extreme pH, wide change of salinity) and conduct extensive cleaning procedures. Biofouling demonstrates itself through increase of pressure drop in the RO unit, followed by general decline of membrane permeability (need to apply higher feed pressure). In initial stages the biofouling it is localized in the feed piping and lead elements.

RO treatment of municipal effluents represents unique situation of high biofouling potential of RO feed water and development of effective way of controlling microorganisms population. In the past, municipal effluents were processed using cellulose acetate membranes. Therefore, it was a common practice to chlorinate RO feed water to prevent microorganisms grow. Due to presence of ammonia and high concentration of organics in the effluent the free chlorine
was converted to chloramines. Today, almost all RO systems treating municipal effluents use composite polyamide membranes. Based on field experience it was established that in wastewater applications salt rejection of composite membranes, in presence of chloramines concentration of 2–4 ppm, is sufficiently stable to provide 3–5 years effective membrane life. The above range of chloramines concentration is also sufficient to effectively mitigate membrane biofouling. The specifics of design and operating conditions of membrane based wastewater reclamation systems are discussed in Chapter 17, Wastewater treatment and reclamation by RO and NF process.

10.8. Permeate recovery ratio

Permeate recovery ratio, which is the rate of conversion of feed water to permeate affects both the economics and process of the RO/NF systems. The volume of raw water required for a given permeate capacity of the RO system is directly determined by the design recovery ratio. Therefore, the size of the raw water supply system, capacity of the pretreatment system, size of the high pressure pump, feed and concentrate manifold are all functions of the recovery ratio as well. On the operating side, permeate recovery rate affects average osmotic pressure and pressure drop. Therefore, recovery rate affects feed pressure and permeate salinity. Higher recovery results in lower feed and concentrate flow rates. For this reason with increased recovery the pressure drop decreases (Fig. 10.10) according to Eq. 54.
\[ \Delta P = A \cdot Q_{fb}^B \]  

(54)

Where \( \Delta P \) is pressure drop, \( A \) and \( B \) are constants specific for element configuration, \( Q_{fb} \) is average of feed–concentrate flow.

Because seawater RO systems operate in lower recovery range than brackish plants, and in seawater systems overall pressure drop is lower due to lower flow rate per vessel, the effect of recovery on pressure drop is more pronounced in brackish units. Recovery rate affects directly the increase of feed concentration along the RO unit, which in turn affects the average feed osmotic pressure (Fig. 2.4). Therefore, higher recovery will result in need of higher operating pressure. In brackish systems, depending on feed salinity, there could be a minimum of feed pressure with increased recovery (Fig. 10.11). The minimum of feed pressure value is result of system pressure drop decrease, which, at sufficiently low feed salinity could be more pronounced that the parallel increase of the osmotic pressure. However, as recovery increases, eventually the osmotic pressure increase will be more significant then the decrease of pressure drop.

In seawater systems the conditions are somewhat different. There, due to high feed salinity, increase of recovery rate is translated directly to increase of required feed pressure (Fig. 10.12). The increase of average feed salinity with recovery rate affects directly permeate salinity both in brackish and seawater systems (Figs. 10.13, 10.14). Due to process economics and limited availability of brackish feed, there is tendency to design brackish water systems for a highest possible recovery rate. Concentrate disposal is usually less expensive at higher recovery. However, in some brackish RO systems, where limitations on concentrate salinity are imposed, the recovery rate has to be adjusted accordingly.

**FIG. 10.11** Required feed pressure vs. recovery rate in a brackish RO unit.
FIG. 10.12  Required feed pressure in vs. recovery rate in a seawater RO unit. Average permeate flux 14 l/m²/h.

FIG. 10.13  Permeate salinity vs. recovery rate in a brackish RO unit. Average permeate flux 28 l/m²/h.

FIG. 10.14  Permeate salinity vs. recovery rate in a seawater RO unit. Average permeate flux 14 l/m²/h.
In majority of cases, in brackish water systems the maximum recovery is limited by solubility of sparingly soluble salts. The scaling constituents that most frequently affect the recovery limits are calcium sulfate and silica. Calcium carbonate, although has also limited solubility, its precipitation can be easily controlled by pH adjustment and/or use of scale inhibitors. Common range of brackish system recovery rate is 75–85%. RO systems operating with very low salinity feed are sometimes designed to operate at 90% recovery.

In seawater RO systems the recovery rate is determined by considerations of maximum feed pressure, operating cost and required permeate salinity. The common range of recovery rate is 35% for feed water of high salinity (≥45,000 ppm TDS) and up to 50% recovery at lower feed salinities (~35,000 ppm TDS). The general tendency is to design RO seawater systems for recovery rate that require feed pressure would not exceed 70 bar (~1000 psi). In the recent efforts to reduce operating cost through minimizing power consumption the recovery rate is being optimized according to feed salinity and temperature. The optimization of operating parameters will be described in the process design section (Chapter 11, RO/NF system design).

The array of pressure vessels in RO unit is affected by the total recovery rate. In brackish RO units the recovery rate is almost always above 60%, most likely in the 70–90% range. The common configuration of brackish unit is of two stages system for the system recovery rate up to 85%. Above 85% recovery rate a three stage system configurations are more likely. To correct the uneven permeate flux distribution along the three stage systems, inclusion of a feed booster pump between stage two and three is quite common. As mentioned already in chapter 5, it is customary to design RO units for a 2:1 ration of pressure vessels in subsequent stages in order to maintain similar average feed flow rate per vessel. As in seawater systems maximum recovery rate seldom exceeds 50%, therefore seawater units at present are mostly design in a single stage configuration. The number of elements per vessel, in a single stage design, is being recently increased from 7 to 8. The 8 elements per vessel configuration has an advantage of lower capital cost of RO unit, smaller footprint and better mem-brane performance due to lower concentration polarization (153, 154).

10.9. Permeate flux rate

Permeate flux represents the water flow rate through a specified area of membrane surface. That is, part of the feed water passes the membrane leaving
on the feed side rejected dissolved constituents and water born particles. The value of the average flux rate (the flux rates averaged over the entire system) is a very important consideration in designing a system. Depending on the feed water source type and how “clean” the water is, if the selected flux rate is too high, membrane surface fouling from the retained constituents becomes more likely. The design average permeate flux rate of the RO system uniquely defines number of membrane elements and pressure vessels required for a given system capacity. Value of the average flux rate also affects the operating parameters. Higher permeate flux rate results in lower permeate salinity and requires higher feed pressure.

Relations between permeate flux, permeate salinity and feed pressure are defined by Eqs. 2–6. As mentioned above, as water flows through the system, rejected dissolved constituents and water borne particles are retained in the concentrate stream. The concentration of the retained constituents near the membrane surface is higher than in the bulk of the feed stream. With increased concentration, formation of a fouling layer on the membrane surface may occur. This excess of concentration at the membrane surface depends on the flux rate and the concentration of the rejected constituents in the feed water.

The constituents, which are of special concern in respect of potential membrane fouling, are particles, biofragments and dissolved organics. The recommended range of average flux rate for an RO system is defined according to the quality of the feed water. Feed water quality is defined in terms of the Silt Density Index, turbidity, TOC and suspended solids. Because in most cases, prior to construction of the desalination system feed water quality parameters are not available, feed water is qualified in terms of water source. From the conventional surface sources, one can expect that the raw water could have a high concentration of potential foulants and that their concentration may experience significant seasonal variations.

Feed water from deep wells is usually of good and consistent quality. It can be assumed that RO permeate, used as a feed in two pass systems, does not have any significant fouling tendency. The corresponding design range of the average flux rate for low salinity applications is 17–23 l/m²/h (10–14 GFD) for surface water, 23–30 l/m²/h (14–18 GFD) for well water and 31–43 l/m²/h (18–25 GFD) for treating RO permeate. When designing second pass systems, treating RO permeate, for operation at a very high flux rate, attention should be paid not to exceed maximum feed flow recommended for a given element type. Feed water to seawater systems originates either from an open intake or shallow beach wells. The range of design flux rate for seawater systems is 11.0–15.0 l/m²/h (6.5–9.0 GFD)
for feed water from an open intake and 13.6–19.0 l/m²/h (8–11 gfd) for a beach well source. In applications of RO membranes treating municipal effluents following membrane pretreatment, stable performance can be maintained at flux rate of 17–21 l/m²/h (10–14 gfd).

The above values represent an average permeate flux rates, used for RO system design. In the actual system the permeate flux varies considerably along the system. The permeate flux rate is high in the lead elements and decreases gradually in the direction of feed-concentrate flow. This is due to increase of osmotic pressure and pressure drop and corresponding decrease of NDP (see Eq. 8). An example of feed and osmotic pressures profile in a two stage brackish RO unit is shown in Fig. 10.15. The values of pressures for the feed and concentrate points were calculated for an RO unit treating feed water of 2000 ppm TDS salinity at recovery rate of 85%. For illustration simplicity the change of osmotic pressure and decrease of feed pressure is depicted as straight lines. As shown in Fig. 10.15, feed pressure at the entrance to the RO membranes has to be sufficiently high to compensate for osmotic pressure of the concentrate and pressure drop along the RO unit. Approximately, only about 50% of the applied pressure is available as a driving force of the desalination process.

The specific shape of pressures profile in the RO unit will depend on feed salinity, temperature and recovery rate and permeate flux rate. Corresponding conditions for a single stage seawater unit, processing 40,000 ppm TDS feed at 50% recovery, are shown in Fig. 10.16. It is evident that elements in the lead position, being exposed to higher NDP, will operate at higher flux rate as compared to elements that follows. The difference of the average permeate flux,

![Fig. 10.15](image_url)  
**Fig. 10.15** Approximate osmotic pressures and feed pressure distribution along a two stage brackish RO unit. Feed salinity 2000 ppm TDS, 85% recovery rate.
from the first and last element, could be quite significant. This flux rate difference will be higher for membrane elements with higher specific permeability. Common solutions to equilibrate flux rate along the RO unit are utilization of interstage booster (Fig. 5.3) or combination of low and high permeability membrane elements on subsequent concentrate stages (see information on hybrid configuration in chapter 16).

Table 10.3 lists nominal performance parameters of three common types of commercial brackish RO membranes: regular, high permeability and ultra high permeability. As examples, representative performance of CPA3, ESPA2 and ESPA4 are used. The above elements performance were used to calculate feed pressure, flux distribution and power consumption for a RO unit treating 2000 ppm TDS feed at recovery rate of 85%. The results are summarized in Table

<table>
<thead>
<tr>
<th>Type</th>
<th>Model</th>
<th>Nominal permeate flow, m$^3$/d (gpd)</th>
<th>Nominal salt rejection, %</th>
<th>Element membrane area, m$^2$ (ft$^2$)</th>
<th>Permeability, l/m$^2$/h-bar (gfd/psi)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Regular</td>
<td>CPA3</td>
<td>41.6 (11000)</td>
<td>99.7</td>
<td>36.8 (400)</td>
<td>3.31 (0.133)</td>
</tr>
<tr>
<td>High permeability</td>
<td>ESPA2</td>
<td>9000</td>
<td>99.6</td>
<td>36.8 (400)</td>
<td>4.47 (0.18)</td>
</tr>
<tr>
<td>Ultra high permeability</td>
<td>ESPA4</td>
<td>12000</td>
<td>99.2</td>
<td>36.8 (400)</td>
<td>8.28 (0.33)</td>
</tr>
</tbody>
</table>
10.4. Comparison of results, calculated for system configurations with single type of membranes of increased permeability, indicate that use of higher permeability membranes will result in lower operating pressure and power requirements. However, at the same time the flux disparity between membrane elements in lead and tail positions will increase and the average flux of elements in the lead position will be increasingly higher.

High water permeability of membrane is usually associated with high rate of salt transport. Example of such conditions is shown in example of ESPA4/ESPA4 configuration.

Use of interstage booster pump, as shown in the ESPA4/ESPA4 + (B) example, results in more even flux distribution and lower permeate salinity. The interesting aspect of this particular configuration is that use of interstage booster with very high permeability membranes results in energy consumption somewhat lower than for configuration with a single process pump. For the cases listed in Table 10.4 of a very high permeability ESPA4 membranes this difference is 0.11 kWhr/m³ (0.42 kWhr/kgal) or about 7%. For system utilizing lower permeability membranes and therefore requiring higher feed pressure this small

<table>
<thead>
<tr>
<th>Element type: 1st/2nd stage</th>
<th>Feed pressure, bar (psi)</th>
<th>1st stage concentrate, bar (psi)</th>
<th>2nd stage concentrate, bar (psi)</th>
<th>Permeate salinity ppm TDS</th>
<th>Lead element flux, l/m²/h (gfd)</th>
<th>Tail element flux, l/m²/h (gfd)</th>
<th>Energy consumption, kWhr/m³ (kwhr/kgal)</th>
</tr>
</thead>
<tbody>
<tr>
<td>CPA3/</td>
<td>16.1</td>
<td>14.2</td>
<td>12.7</td>
<td>56</td>
<td>38.1</td>
<td>9.5</td>
<td>0.67</td>
</tr>
<tr>
<td>CPA3</td>
<td>(233)</td>
<td>(206)</td>
<td>(184)</td>
<td>(22.4)</td>
<td>(22.4)</td>
<td>(5.6)</td>
<td>(2.54)</td>
</tr>
<tr>
<td>ESPA2/</td>
<td>13.9</td>
<td>12.1</td>
<td>10.8</td>
<td>63</td>
<td>41.6</td>
<td>5.9</td>
<td>0.58</td>
</tr>
<tr>
<td>ESPA2</td>
<td>(202)</td>
<td>(175)</td>
<td>(157)</td>
<td>(24.5)</td>
<td>(24.5)</td>
<td>(3.5)</td>
<td>(2.19)</td>
</tr>
<tr>
<td>ESPA4/</td>
<td>10.6</td>
<td>9.1</td>
<td>8.1</td>
<td>277</td>
<td>56.2</td>
<td>0.3</td>
<td>0.44</td>
</tr>
<tr>
<td>ESPA4</td>
<td>(154)</td>
<td>(132)</td>
<td>(117)</td>
<td>(33.1)</td>
<td>(33.1)</td>
<td>(0.2)</td>
<td>(1.66)</td>
</tr>
<tr>
<td>CPA3/</td>
<td>15.4</td>
<td>13.5</td>
<td>12.0</td>
<td>55</td>
<td>36.5</td>
<td>10.0</td>
<td>0.64</td>
</tr>
<tr>
<td>CPA3</td>
<td>(223)</td>
<td>(196)</td>
<td>(174)</td>
<td>(21.5)</td>
<td>(21.5)</td>
<td>(5.9)</td>
<td>(2.42)</td>
</tr>
<tr>
<td>ESPA2/</td>
<td>13.9</td>
<td>11.8</td>
<td>10.0</td>
<td>130</td>
<td>32.3</td>
<td>9.2</td>
<td>0.58</td>
</tr>
<tr>
<td>ESPA4</td>
<td>(202)</td>
<td>(171)</td>
<td>(145)</td>
<td>(19.0)</td>
<td>(19.0)</td>
<td>(5.4)</td>
<td>(2.19)</td>
</tr>
<tr>
<td>ESPA2</td>
<td>12.8</td>
<td>10.9</td>
<td>9.4</td>
<td>148</td>
<td>37.9</td>
<td>9.4</td>
<td>0.53</td>
</tr>
<tr>
<td>ESPA4</td>
<td>(186)</td>
<td>(158)</td>
<td>(136)</td>
<td>(27.3)</td>
<td>(27.3)</td>
<td>(5.5)</td>
<td>(2.00)</td>
</tr>
<tr>
<td>ESPA4+(B)</td>
<td>8.1</td>
<td>6.2</td>
<td>9.6</td>
<td>188</td>
<td>40.4</td>
<td>7.5</td>
<td>0.41</td>
</tr>
</tbody>
</table>

(B)–Interstage booster pump–5 bar (73 psi)
energy benefit associated with interstage booster in a multistage system configuration declines. In seawater systems the single stage configuration is always more energy efficient than corresponding configuration of two stages or two stages with interstage booster. This is due to a lower pressure drop across the single stage as compared to two stage system operating at the same average permeate flux rate.

Results included in Table 10.4 shows that hybrid design, combination of two membrane types of different permeability, provides more even flux distribution, good permeate salinity and intermediate power consumption requirement. Availability of high permeability membranes with acceptable salt rejection makes this solution a convenient and cost effective design alterative.

Apart from considerations of energy requirement and permeate salinity, it is not clear at this point if conditions of large difference of permeate flux between load and tail elements and high permeate flux of elements in lead position are detrimental to long term membrane performance stability. It has been documented in field operation that RO system designed at high permeate flux rate have high fouling rates. This is especially the case for system treating water containing high concentration of dissolved organics. However, if RO system is designed for an average flux rate, considered adequate for a given type of feed water, excessive flux of the lead element, may not necessary result in accelerated membrane fouling. The lead elements that operate at high permeate flux rate, treat water with relatively low concentration of fouling constituents compared to the elements in tail position. Low concentration of foulant and high feed flow rate in lead elements could mitigate fouling process even at high flux rates. It can be also assumed that the RO system is self regulating in respect of surface absorption fouling. If as a result of fouling the permeability of elements in lead position declines, the system is converting itself to a hybrid configuration of low permeability elements followed by higher permeability and flux distribution should improve, effectively reducing fouling rate. So far little information from field operation is available to confirm the above assumptions and probably will be safer for a designer to follow the current approach and configure brackish RO system for a more uniform permeate flux distribution.

10.10. Membrane age

Performance of membrane elements changes with operating time. Usually permeability declines and salty passage increases. The membrane barrier, made
of aromatic polyamide is very robust. However, formation of fouling layer on the membrane surface, abrasive effect of particles in the feed water and exposure to extreme pH cleaning chemicals will eventually change property of membrane surface and result in performance deterioration. These expected changes of performance are accounted for in calculation of projected performance of RO system by assuming annual increase of salt passage and decrease of permeability. The approach varies among different membrane manufacturers but it is generally assumed that salt passage could increase up to 10% per year and permeability irreversibly decline by 7%. Fig. 10.17 shows changes of membrane performance according to Eqs. 55 and 56. The assumptions are that the salt passage increase is linear and permeability decline is a compound function of operating time.

\[
SP_y = SP_0 (1 + SPI \times Y) \quad (55)
\]
\[
Perm_y = Perm_0 (1 – Decl)^Y \quad (56)
\]

where \(SP_y\) is salt passage at the year \(Y\), \(SP_0\) is the initial salt passage and \(SPI\) is salt passage increase factor, expressed as a decimal fraction. For the permeability (Perm) the same designation applies.

Accordingly for 3rd year of operation at 10% and 7% increase:

\[
SP_3 = SP_0 (1 + 0.1 \times 3) = SP_0 \times 1.30
\]
\[
Perm_3 = Perm_0 (1 – 0.07)^3 = Perm_0 \times 0.80
\]

The useful membrane life is determined by changes of membrane performance and required quality of product water and operating feed pressure. The

![Image](https://via.placeholder.com/150)

**FIG. 10.17** Projected changes of membrane performance (salt passage and water permeability) with operating time.
usual approach is to budget in the operating cost an annual replacement of some fraction of the number of membrane elements in the system. However, for practical reasons the actual replacement is conducted by replacing all membrane elements in a train or desalting stage when system performance deteriorates below the acceptable level in respect of permeate salinity or feed pressure. Membrane replacement is usually conducted after attempts to restore membrane performance through cleaning did not produce satisfactory results. In order to reduce membrane replacement component of the operating cost, the old elements removed from the system are tested individually. Those with satisfactory performance are used in the next replacement operation.

For projections of long term performance an estimation of average membrane life is necessary. The estimation is based on assumed (or warranted) membrane replacement rate. An example of calculation of average membrane age is given in Table 10.5.

In the above table the average age is calculated based on setting the membrane replacement event at the beginning of the given year of plant operation. If, for example, the replacement would be conducted at the year end, starting at the end of first year, then the average membrane age at the end of 5th year would be higher, about 3.75 years. As will be discussed in the following chapters, the average membrane age is used as a parameter for calculation of membrane performance, which eventually is the basis for the terms of system performance warranty.

If the attempts to restore system performance by applying membrane cleaning are not sufficiently successful, the only other alternative is to correct performance through membrane replacement. The fraction of membrane elements

<table>
<thead>
<tr>
<th>Year of operation</th>
<th>0</th>
<th>1</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>Average age</th>
</tr>
</thead>
<tbody>
<tr>
<td>0</td>
<td>100%</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0</td>
</tr>
<tr>
<td>1</td>
<td>15%</td>
<td>85%</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>0.85</td>
</tr>
<tr>
<td>2</td>
<td>15%</td>
<td>15%</td>
<td>70%</td>
<td></td>
<td></td>
<td></td>
<td>1.55</td>
</tr>
<tr>
<td>3</td>
<td>15%</td>
<td>15%</td>
<td>15%</td>
<td>55%</td>
<td></td>
<td></td>
<td>2.10</td>
</tr>
<tr>
<td>4</td>
<td>15%</td>
<td>15%</td>
<td>15%</td>
<td>15%</td>
<td>40%</td>
<td></td>
<td>2.50</td>
</tr>
<tr>
<td>5</td>
<td>15%</td>
<td>15%</td>
<td>15%</td>
<td>15%</td>
<td>25%</td>
<td>25%</td>
<td>2.75</td>
</tr>
</tbody>
</table>
that have to be replaced will depend on membrane conditions and required level of system performance.

Table 10.6 demonstrates effect of element replacement on permeate salinity. In this example the system with new elements produced initially permeate salinity of 240 ppm TDS. The upper limit of allowed permeate salinity is 400 ppm TDS and at the current conditions of the elements in the system the permeate salinity is 500 ppm TDS. The fraction of the elements to be replaced depends on the target permeate salinity after replacement. Under assumption that new elements will have the same nominal salt rejection as the original load, to restore system to the ability to produce permeate of the original salinity (240 ppm TDS) all elements in the system will have to be replaced. To bring system in compliance with required permeate salinity of 400 ppm TDS a 37% of elements will have to be replaced.

Table 10.6 demonstrates two options for productivity restoration. In this example the system was designed to maintain rated capacity at 80% of nominal elements productivity. Due to excessive fouling the current productivity is only 70%, example A (or 60%, example B). To recover lost productivity two alternatives are possible: some of the old elements can be replaced with new ones or new elements can be added to the system.

The results listed in Table 10.7 demonstrate that it is more cost effective, for productivity increase, to add new elements to the system rather than replace some of the old elements. However, with the addition of elements, the mem-

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**Table 10.6**

Membrane replacement for permeate salinity correction.

<table>
<thead>
<tr>
<th>Target permeate salinity, ppm TDS</th>
<th>Actual salt passage, %</th>
<th>Nominal salt passage, %</th>
<th>Elements to be replaced, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>240</td>
<td>0.43</td>
<td>0.30</td>
<td>100</td>
</tr>
<tr>
<td>300</td>
<td>0.52</td>
<td>0.38</td>
<td>80</td>
</tr>
<tr>
<td>350</td>
<td>0.62</td>
<td>0.44</td>
<td>58</td>
</tr>
<tr>
<td>400</td>
<td>0.71</td>
<td>0.50</td>
<td>37</td>
</tr>
<tr>
<td>500</td>
<td>0.88</td>
<td>0.63</td>
<td>—</td>
</tr>
</tbody>
</table>
brane area in operation will increase. This will result in increased permeate salinity. The salinity increase will be in proportion to the increase of membrane area. Therefore, the option of adding elements can only be implemented if a sufficient permeate quality margin exists.

TABLE 10.7
Permeate flow restoration through membrane replacement or addition

<table>
<thead>
<tr>
<th>Flux restoration mode</th>
<th>Current capacity, %, examples A &amp; B</th>
<th>Target capacity, %</th>
<th>Elements to be replaced or added, %</th>
</tr>
</thead>
<tbody>
<tr>
<td>Elements replacement (A)</td>
<td>70</td>
<td>80</td>
<td>33</td>
</tr>
<tr>
<td>Elements replacement (B)</td>
<td>60</td>
<td>80</td>
<td>50</td>
</tr>
<tr>
<td>Elements addition (A)</td>
<td>70</td>
<td>80</td>
<td>10</td>
</tr>
<tr>
<td>Elements addition (B)</td>
<td>60</td>
<td>80</td>
<td>20</td>
</tr>
</tbody>
</table>