

4

Reverse Osmosis

Richard G. Sudak

INTRODUCTION

The first large industrial application of reverse osmosis occurred in 1970 when a 100,000 gallon per day (GPD) system was placed in operation at Texas Instruments' electronics manufacturing facility in Dallas, Texas. In this application, the reverse osmosis plant was used to pretreat Dallas municipal water which was being converted to ultrapure process water by ion exchange resins. It was reported by the facility engineer that the use of reverse osmosis increased the yield in the manufacturing operation to such an extent that the resultant savings paid for the reverse osmosis plant in about two weeks. Now, virtually every electronics plant in the United States uses reverse osmosis as pretreatment in the preparation of ultrapure water.

The worldwide total reverse osmosis operating capacity by the end of 1970 was 880,000 GPD. By the end of 1976, reverse osmosis operating capacity had grown to 167,000,000 GPD¹ and, by the middle of 1980, this capacity had more than doubled to 390,500,000 GPD.² Reverse osmosis continued to expand and it was reported³ that the operating capacity by the end of 1984 was 524,000,000 GPD.

Figure 4.1 shows that the rate of growth in reverse osmosis capacity was slow between 1970 and 1973. From 1974 through 1980, there was an accelerated rate of growth during which time the reverse osmosis capacity increased from 62,500,000 GPD to 412,500,000 GPD or an increase of 560% in the 7-year period. Between 1981 and 1984 (4 years), the total operating capacity increased from 412,500,000 to 524,000,000 GPD, an increase of 27%.

Reverse osmosis finds applications predominantly in the following areas:

- Industrial - Used to prepare industrial process water or to process wastes.

- Irrigation - Used to upgrade waters for agricultural purposes.
- Military - Used for military purposes.
- Municipal - Used to upgrade waters to municipal drinking water levels.
- Power - Used to prepare process water in electric power stations.
- Miscellaneous - Used for various purposes.

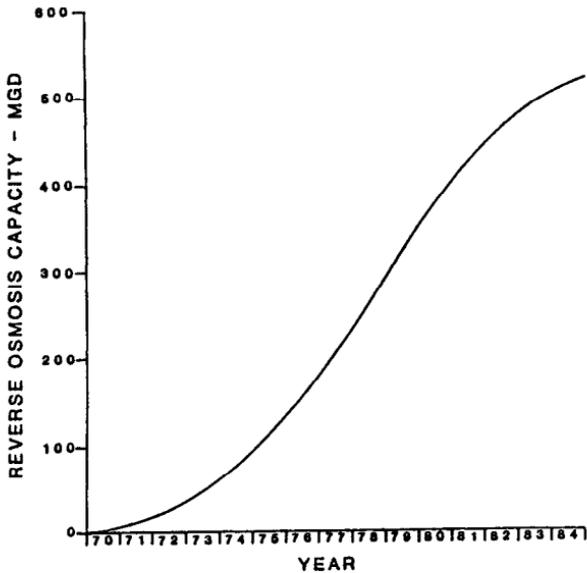


Figure 4.1: Reverse osmosis capacity vs. year.

Table 4.1³ below shows the uses of reverse osmosis as described above.

Table 4.1: Reverse Osmosis Product Water Usage

<u>Product Water Use</u>	<u>Capacity, MGD*</u>	<u>Percent of Total</u>
Industrial	158	31.5
Irrigation	2	0.4
Military	13	2.6
Municipal	191	38.0
Miscellaneous	10	2.0
Power	55	11.0
Yuma Desalting Plant	<u>73</u>	<u>14.5</u>
Total	502	100.0

* MGD = One million gallons per day

Reverse osmosis finds much of its application in the production of potable water for municipal water supplies. However, the use of reverse osmosis to prepare industrial process waters and to treat industrial waste is in a close second place. Agricultural irrigation accounts for only a very small percentage of total capacity because current sources of agricultural water are much cheaper than water from reverse osmosis. The military use of the reverse osmosis process was small as of the end of 1984 but the U.S. military services have discovered reverse osmosis to be a process ideally suited to providing drinking water in the field. The process not only is suitable for converting brackish and seawaters into potable supplies but it will also remove toxic and biological agents which result from germ, chemical and nuclear warfare. Consequently, military use of reverse osmosis should increase in the future. Power plants have found reverse osmosis very useful in preparing boiler feedwater. In Yuma, Arizona, a desalting plant is being constructed by the U.S. Government to improve the quality of the Colorado River prior to its flowing into Mexico. That plant alone accounts for 14.5% of the worldwide capacity of the reverse osmosis process.

Reverse osmosis is a process that transforms an unusable water supply into a usable resource. It is capable of renovating a broad spectrum of feedwaters from municipal water supplies that need polishing for industrial purposes to seawater that is refined into a potable water supply. Table 4.2³ shows the different types of feedwater being processed by reverse osmosis units. Seawater is considered to have a nominal total dissolved solids (TDS) content of 35,000 mg/l. Wastewater is from industrial or municipal sources and the TDS is variable. Brackish water is defined for the purposes of Table 4.2 only as a water that may have a TDS from that of municipal water supplies up to 10,000 mg/l.

Table 4.2: Source of Reverse Osmosis Feedwater

	<u>Capacity, MGD</u>	<u>Percent of Total</u>
Sea Water	67.9	13.0
Waste Water	26.5	5.0
Brackish Water	<u>429.6</u>	<u>82.0</u>
Total	524.0	100.0

As of the end of 1984, the desalination of brackish water accounted for 82% of capacity. This is due to the fact that early reverse osmosis membranes were incapable of single stage seawater desalination and, thus, they were limited to brackish water desalination. Within the last 10 years, significant advances have been made in both the flux and rejection capability of membranes and reverse osmosis is technically able to desalt seawater in a single stage. In the recent past, it has been an effective competitor to the distillation process in seawater desalination. In fact, reverse osmosis is now beginning to replace existing distillation capacity in the Middle East.⁴

Although reverse osmosis is a relatively new technology, there is sufficient operating capacity in a number of varied applications to warrant confidence in the process. From a technical and economic point of view the process is ca-

pable of desalting a broad range of feedwaters from municipal water supplies to seawater. It has economic viability in a large number of industrial applications.

BASIC PROCESS CONSIDERATIONS

An industrial reverse osmosis plant usually will consist of three separate sections which are shown in Figure 4.2. The first section is the pretreatment section in which the feedwater is treated to meet the requirements of reverse osmosis element manufacturers and the dictates of good engineering practice. Following pretreatment, the feedwater is introduced into the reverse osmosis section where the feedwater is pressurized and routed to the reverse osmosis elements which are in pressure vessels. The feedwater flows across the membrane surface where product water permeates through the membrane and a predetermined amount remains behind as reject. The reject is discharged to waste while the product water is routed to the posttreatment section. The third or post-treatment section treats the product water to remove carbon dioxide and adds chemicals as required for industrial use of the product water.

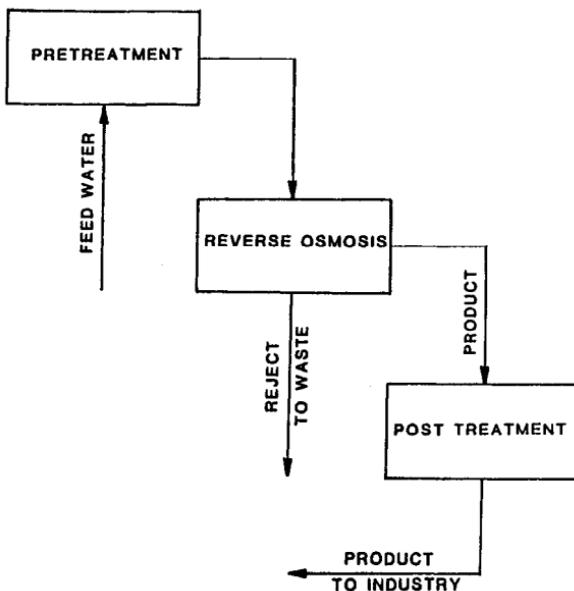


Figure 4.2: Industrial reverse osmosis.

Although operating reverse osmosis plants are commonplace, the subject of salt and water transport is still controversial. Several transport models have been proposed and these will be discussed briefly below. A more thorough presentation is made in a paper by Soltanieh and Gill.⁵

The sieve mechanism is the simplest and easiest model to understand. This model proposes that salt and water are separated due to physical size differences by a membrane with a pore size that lies between the sizes of salt and water. While most laymen prefer this concept, it is unfortunate that for solutions, such as sodium chloride in water, the sizes of the salt and water molecules are almost the same. This fact would seem to rule out the sieve mechanism model.

Another model proposed is the wetted surface mechanism or the water clustering concept of transport. It is generally recognized that reverse osmosis membrane material is wettable and that water tends to be absorbed on the membrane by hydrogen bonding. In this concept, it is theorized that the water film at the surface of the membrane obstructs the pores and prevents salt from entering the pores. The water passes through the membrane by passing from one absorbed site to the next until it reaches the other side of the membrane barrier layer. The energy requirements for solvent migration are much less than the energy requirements for salt migration and, thus, the separation of salt from water takes place.

Another concept of water and salt transport in reverse osmosis is the preferential sorption-capillary flow mechanism. In this model, the surface of a membrane is microporous and heterogeneous at all levels of solute separation. It is hypothesized that, due to the chemical nature of the membrane skin layer in contact with the aqueous solution, a preferential sorption for the water causes a sorbed water layer to be formed at the skin. This layer of purified water is then forced through the capillary pores by pressure.

The solution-diffusion model of transport assumes a nonporous, homogeneous membrane surface layer. Each component in a pressurized solution dissolves in the membrane and then diffuses through the membrane. The flow of water and salt through the membrane is uncoupled, i.e., they are independent of each other, and the water transports through the membrane at a more rapid rate than the salt.

The product water flow through the membrane is defined as follows:

$$F_w = A^*(\Delta P - \Delta \pi)$$

- where F_w = water flux through the membrane
 A = water transport coefficient
 ΔP = pressure differential across the membrane
 $\Delta \pi$ = osmotic pressure differential across the membrane

The flow of water through a reverse osmosis membrane is primarily dependent on the applied pressure differential and the osmotic pressure differential across the membrane. The osmotic pressure is directly dependent on the salt concentration of the process stream. As a rule of thumb, each 100 mg/l of dissolved solids is roughly equivalent to one psi of osmotic pressure. Since the product stream usually has a very low salt content, the osmotic pressure of that stream is negligible. In addition, the product stream normally leaves the reverse osmosis pressure vessels at near atmospheric pressure so that the applied pressure differential is the feed pressure. Consequently, the term "net applied pressure" has come to mean the applied pressure minus the feed osmotic pressure.

The flow of salt or dissolved solids across the membrane is dependent on the following equation:

$$F_s = B \cdot \Delta C$$

- where F_s = salt flux
 B = salt transport coefficient
 ΔC = salt concentration gradient across the membrane

The salt transport is primarily dependent upon the concentration of dissolved solids on each side of the membrane.

The solution-diffusion model seems to represent the performance of a reverse osmosis membrane. Figure 4.3 shows the salt rejection and flux of a low pressure polyamide membrane as a function of applied pressure. The membrane was operated on a 5,000 mg/l aqueous solution of sodium chloride at 25°C. As can be seen, there was no product water flow until the applied pressure exceeded the osmotic pressure (50 psi). After this, the flux increased linearly as would be predicted by the above water flux equation. Rejection is poor at lower pressures and increases rapidly until it reaches an asymptote at an applied pressure of about 150 psig. This can be attributed to a near constant flow of salt with a rapidly increasing product water flow which results in a more dilute product or in increased rejection. These data tend to substantiate the assertion of the solution-diffusion model that flow is uncoupled.

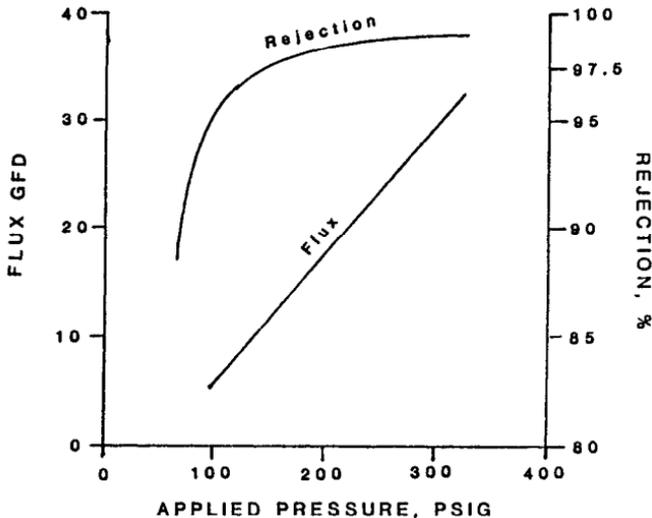


Figure 4.3: Membrane flux and rejection vs. applied pressure.

It is noted that the data shown in Figure 4.3 was derived in membrane test cells with near perfect small areas of membrane. In a practical system, there would be a number of imperfections in the membrane and the salt flow through these capillaries would contribute to the total salt flow. Therefore, a practical

salt transport model must also take into account the contribution of the membrane imperfections to salt flow.

The water transport coefficient (A) is not a constant in that it varies with temperature. The product flow as a function of temperature may be estimated⁶ by using the following equation:

$$Q_{25} / Q_t = e^x$$

- where Q_{25} = flow at 25°C or 77°F
 Q_t = flow at temperature T, °C
 e = 2.71828
 x = $U [1/(T + 273) - 1/298]$
 T = Temperature in °C
 U = 2723 (for cellulose acetate membranes).

Figure 4.4 graphically shows the ratio of water flux at 25°C to water flux at other temperatures.

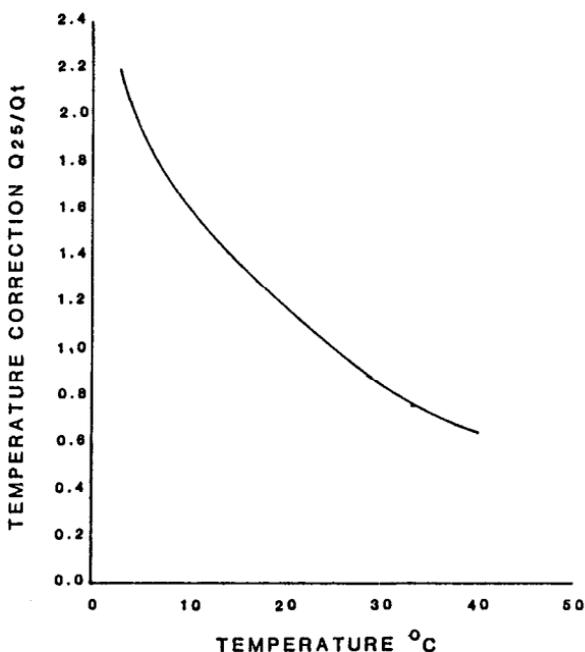


Figure 4.4: Temperature correction factor.

As a rule of thumb, the product water flow with constant net applied pressure will increase about 3% for each degree centigrade increase in feedwater temperature. Salt flux through the membrane is also directly proportional to temperature and the ratio of salt flux to water flux is essentially constant at different temperatures. This results in little or no change in rejection as a function of

temperature. For some of the newer composite membranes, the water and salt permeation coefficients also vary as a function of pH.

Reverse osmosis is a cross-flow membrane separation process which separates a feed stream into a product stream and a reject stream. The recovery of a reverse osmosis plant is defined as a percentage of feedwater that is recovered as product water. As all of the feedwater must be pretreated and pressurized, it is economically prudent to maximize the recovery in order to minimize power consumption and the size of the pretreatment equipment. Since most of the salts remain in the reject stream, the concentration of salts increases in that stream with increased recovery. For instance, at 50% recovery, the salt concentration in the reject is about double that of the feed and at 90% recovery, the salt concentration in the reject is nearly 10 times that of the feed. In cases of sparingly soluble salts, such as calcium sulfate, the solubility limits may be exceeded at a high recovery. This could result in precipitation of the salt on the membrane surface resulting in decreased flux and/or increased salt passage. In addition, an increase in recovery will increase the average salt concentration in the feed/reject stream and this produces a product water with increased salt content. Consequently, the recovery of a reverse osmosis plant is established after careful consideration of the desired product quality, the solubility limits of the feed constituents, feedwater availability and reject disposal requirements.

While the above equations are helpful in describing the reverse osmosis process, the water and salt transport coefficients seldom are used to describe membrane performance. Most manufacturers test reverse osmosis membrane with standard solutions as described below.

- (1) *Brackish Water Membrane*—The flux and rejection of the membrane is determined when the membrane is tested on a feedwater of an aqueous sodium chloride solution with a concentration of 2,000 mg/l at an applied pressure of 420 psig (net applied pressure = 400 psig) with a temperature of 25°C and a feed pH of from 5.0 to 6.0 for 30 minutes prior to data collection.
- (2) *Seawater Membrane*—The seawater test is similar to the brackish water test except that the feedwater is an aqueous sodium chloride solution with a concentration of 35,000 mg/l and the test pressure is 800 psig.

The results are reported as gallons per day per square foot of membrane area (membrane flux) and as rejection of sodium chloride. Rejection is calculated as follows:

$$R = (1 - CP/CF) 100$$

where R = Rejection, percent
 CP = Product water concentration
 CF = Feedwater concentration

The sodium chloride rejection differs from that of other inorganic and organic dissolved solids, and membrane manufacturers will provide information and rejection data that are available for their specific membrane. Table 4.3

shows typical results for a composite membrane when tested on a multicomponent solution. The rejection of the divalent ions such as calcium and sulfate is much better than the rejection of the monovalent ions such as sodium and chloride. If salt passage is defined as product concentration divided by the feed concentration, or one minus rejection, then it can be seen that the salt passage for the divalent ions is about one-fifth of the salt passage for the monovalent ions.

Table 4.3: Ionic Rejections

<u>Ion</u>	<u>Feed, mg/l</u>	<u>Product, mg/l</u>	<u>Rejection, %</u>
Calcium	61	0.2	99.6
Sodium	150	3.0	98.0
Potassium	12	0.3	97.4
Bicarbonate	19	0.7	96.2
Sulfate	189	0.4	99.8
Chloride	162	2.9	98.2
Nitrate	97	3.5	96.4
TDS	693	11.0	98.4

The abovedescribed tests are conducted at a low recovery rate to minimize the effects of concentration polarization which is described below. For example, membrane tests above are conducted at less than 1% recovery and tests with spiral wound elements are conducted at recoveries from 5 to 10%.

Reverse osmosis is a cross-flow process and, as in any dynamic hydraulic process, the fluid adjacent to the membrane moves slower than the main stream. While the main stream flow may be turbulent, the layer next to the membrane surface is laminar. This thin, laminar flow film is called the boundary layer. When water permeates through the membrane, nearly all of the salt remains behind in the boundary layer next to the membrane. The salt must then diffuse across the boundary layer and back into the bulk stream. This results in a boundary layer with a salt concentration which is more concentrated than the bulk stream. The effect has been termed concentration polarization, and it is defined by the following equation:

$$\beta = CB/CM$$

- where β = Concentration polarization
 CM = Concentration in the main stream
 CB = Concentration in the boundary layer

Concentration polarization increases the salt concentration at the membrane surface, and this results in an increased osmotic pressure at that surface. The increased osmotic pressure causes a drop in water flow as shown in the following equation:

$$F_w = A * (P - \beta \pi)$$

where F_w	=	the water flow
A	=	the water transport coefficient
P	=	applied pressure
β	=	concentration polarization
π	=	osmotic pressure of main stream

The increased salt concentration at the membrane surface will also increase the tendency of sparingly soluble salts to precipitate on the membrane.

The flow of salt also increases and this can be simply shown in the following equation:

$$F_s = B * (\beta C_1 - C_2)$$

where F_s	=	salt flux
B	=	salt transport coefficient
β	=	concentration polarization
C_1	=	main stream salt concentration
C_2	=	product water salt concentration

In the imperfect membrane with a small number of pores, the increased salt concentration at the membrane surface would also result in increased salt passage through the pores which would be directly proportional to concentration polarization.

The information shown in Table 4.4 below assumes a membrane operating on various feedwaters with a total dissolved solids (TDS) content of 2,000, 5,000 and 35,000 mg/l. It is assumed that the membrane will deliver 20 gallons per square foot per day at a 400 psi "net applied pressure" and have a rejection of 99% when there is no concentration polarization, i.e., $\beta = 1.0$.

Table 4.4: Effects of Concentration Polarization

<u>Feed TDS, mg/l</u>	<u>2,000</u>	<u>5,000</u>	<u>35,000</u>
Flux, gfd			
$\beta = 1.0$	20.0	20.0	20.0
$\beta = 1.1$	19.9	19.8	18.3
$\beta = 1.5$	19.5	18.8	11.3
$\beta = 2.0$	19.0	17.5	2.5
Rejection, %			
$\beta = 1.0$	99.0	99.0	99.0
$\beta = 1.1$	98.9	98.9	98.8
$\beta = 1.5$	98.5	98.4	97.3
$\beta = 2.0$	97.9	97.7	84.0

The penalty of a high concentration polarization is not as severe with a low TDS feedwater as it is with a high TDS feedwater. The recommendations as to minimum flows or maximum recoveries which are specified by the reverse osmosis element manufacturer should be followed at all times, especially when the application is the desalination of high TDS waters.

Concentration polarization cannot be eliminated, but it can be minimized by decreasing boundary layer thickness. This is done by increasing the flow rate across the membrane surface or introducing turbulence promoters into the feed/reject stream. In order to achieve optimum performance, most membrane manufacturers will recommend a minimum feed rate to or from their elements and a maximum recovery in order to minimize the effects of concentration polarization.

REVERSE OSMOSIS MEMBRANES

Osmotic phenomena have been observed since the middle of the eighteenth century. The first experiments were conducted with animal membranes and it wasn't until 1867 that artificial membranes were employed. In the early 1950's, research workers at the University of Florida demonstrated, with thick films, that cellulose acetate possessed unique salt and water transport properties which made it potentially attractive as a reverse osmosis desalination membrane. During the 1960's, Loeb and others at the University of California at Los Angeles developed techniques to prepare cellulose acetate membranes with an economical water flux and salt rejection at moderate driving pressures. With this development, reverse osmosis became a practical possibility.

The early cellulose acetate membranes were only suitable for brackish water desalination as they were not capable of operating at the high pressures required for seawater desalination (see below). In addition, the membrane rejection was insufficient to desalt seawater in one stage. The first membranes were fabricated from cellulose diacetate which is subject to hydrolysis at either a low or high pH, biological attack and compaction at temperatures in excess of 85°F. Cellulose acetate also compacts (densifies) at pressures above 400 psig and temperatures below 85°F. Suitable feedwater pretreatment lessens the adverse effects of temperature, pH and biological organisms to the point where the membrane can be used in practical reverse osmosis plants.

A considerable amount of research has been done to develop a membrane of cellulose triacetate as this material is more stable to extremes of temperature and pH and it will better withstand chemical and biological attack. While the cellulose triacetate membranes operated quite well on a short term basis, they were prone to compact at an operating pressure of 400 psig, with the resultant loss of flux to an impractical level, in a short period of time.

It has been demonstrated that a blend of cellulose diacetate and cellulose triacetate provides an improved membrane in that:

- (1) it is more stable than the cellulose diacetate membranes;
- (2) it has a better flux and rejection than the cellulose diacetate; and
- (3) it is more resistant to compaction than either the diacetate or triacetate.

Incidentally, all of the cellulose acetate membranes will tolerate a limited amount of residual chlorine which allows chlorine to be used for feedwater disinfection.

The cellulose acetate membranes are asymmetric and fabricated from a single polymer. The use of electron microscopy in the 1960's demonstrated that the cellulose acetate membranes consisted of a relatively thin dense layer and a thicker porous layer of the same material. The membrane thickness is usually about 100 micrometers with the dense layer accounting for about 0.2% of the thickness and the remainder being an open cell porous matrix (see Figure 4.5).

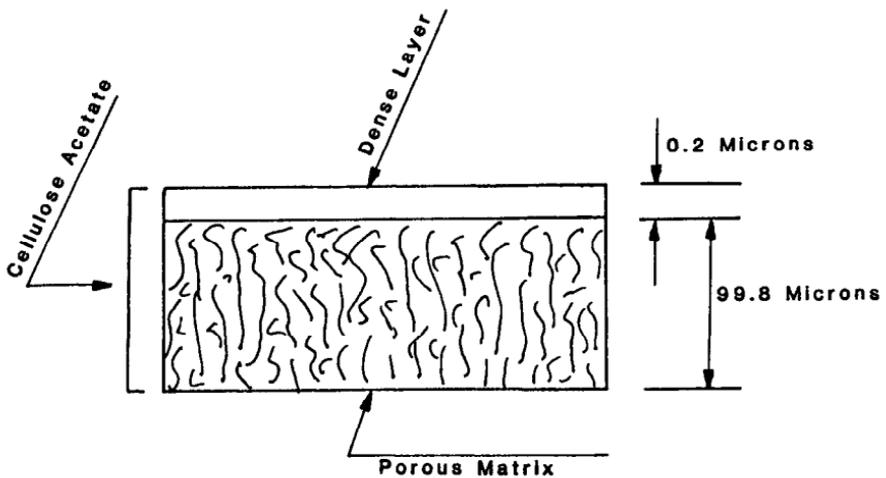


Figure 4.5: Single component asymmetric membrane.

During the 1960's, the DuPont Company screened numerous polymers to determine the suitability of materials other than cellulose acetate for use in reverse osmosis desalination. The results of this work indicated that aromatic polyamides were the "choice as the best polymer type for use in the DuPont commercial permeators".⁷ The company was most successful in developing an asymmetric aromatic polyamide reverse osmosis membrane in a hollow fine fiber configuration which successfully competed with cellulose acetate in the market place.

Shortly after the concept of an asymmetric membrane was established, composite membrane research was initiated. A composite membrane is also asymmetric but it consists of two polymer layers which are the membrane barrier layer and the porous support layer (Figure 4.6). The porous support is formed separately, by conventional membrane casting techniques, from one polymer. The porous support has a thickness of between 75 and 100 micrometers and its porosity is due to numerous small perforations through the support. The membrane barrier layer is a dense thin film of another polymer that is formed or deposited in a subsequent operation on the porous support. The membrane barrier layer varies in thickness from 400 to 1,000 angstroms.

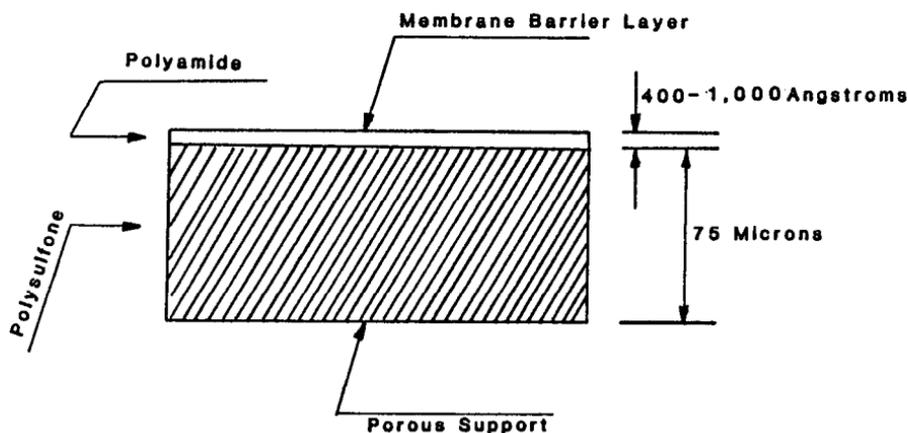


Figure 4.6: Two component composite membrane.

Several polymers have been used as porous supports. One of the earliest composite membrane systems was a porous support formed from cellulose nitrate-cellulose acetate with a membrane barrier layer of cellulose triacetate. While this membrane successfully desalted seawater, it was fragile and expensive. To a large degree, present day commercially available composite membranes use a polysulfone porous support.

Membrane barrier layers have been formed on porous supports in the following manner:⁸

- (1) solution,
- (2) thin film polymerization and
- (3) interfacial polycondensation.

The solution coating technique was used in the preparation of the cellulose triacetate membrane discussed above. A solution of cellulose triacetate in chloroform was deposited on the porous support and the solvent was then evaporated leaving a thin film on the porous support. Thin film polymerization was used to prepare a polyfuran membrane barrier layer on polysulfone. In this case, the monomer furfuryl alcohol is polymerized in situ by adjustment of pH and temperature. This membrane proved to be highly susceptible to oxidizing agents and is of limited value. By far the most valuable technique in the formation of membrane barrier layers is interfacial polycondensation. In this method, a polymer is formed on the porous support surface at the interface of organic and aqueous phases by reaction of specific molecules dissolved in each phase. It is by this method that a number of polyamides and polyurea membrane barrier layers have been formed on polysulfone. Elements containing these membranes are available commercially.

There has been very little progress in the last five years in improving the performance of single polymer asymmetric membranes. Meanwhile, the composite membranes have been improved and they exhibit a higher flux and better re-

jection at lower operating pressures than is available with the single component asymmetric membranes. The polyamide and polyurea composite membranes can withstand higher temperatures and larger pH variations than the cellulose acetate based membranes and they are immune to biological attack. On the other hand, as noted above, the cellulose acetate membranes can tolerate a limited concentration of residual chlorine while the polyamide and polyurea membrane barrier layers are subject to disintegration by residual chlorine and other oxidizing agents. Consequently, alternate methods of water disinfection or chlorination followed by dechlorination are employed in reverse osmosis systems using these membranes.

While the cellulose acetate membranes are compacted at the moderate pressures required for brackish water desalination, the thin film composite membranes exhibit no compaction at these pressures and a very low compaction rate at pressures of 1,000 psig.

The superior flux and rejection capabilities of the thin film composite membrane has been demonstrated at the municipal wastewater reclamation facility of the Orange County Water District in California. Both asymmetric cellulose acetate and thin film composite membranes were tested on lime clarified secondary effluent. The pilot plants were operated at 85% recovery and the rejections reported in Table 4.5 are the percent rejection of the constituents in the feed-water and not the rejection of the average concentration of the specific constituents in the feed/reject stream. Use of the average concentration would give a higher rejection in both cases.

Table 4.5: Rejection Comparison

<u>Constituent</u>	<u>CA Membrane(9)</u>	<u>TFC Polyamide Membrane(10)</u>
Na	93.8	99.0
NH ₄	89.0	94.8
SO ₄	99.3	99.8
Cl	93.3	95.2
NO ₃	71.0	97.7
COD	93.6	>94.8
TDS	94.4	97.5

The thin film composite membrane exhibited superior overall rejection performance in these tests, with ammonia and nitrate rejection showing an outstanding improvement. It has also been reported that silica rejection by the thin film composite membranes is superior to that of cellulose acetate. While the above data indicates a marginal improvement in the rejection of chemical oxygen demand (COD), which is an indication of organic content, other tests conducted by membrane manufacturers show that the polyurea and polyamide membrane barrier layers exhibit an organic rejection that is clearly superior to that of cellulose acetate. Reverse osmosis element manufacturers should be contacted for rejection data on specific organic compounds.

MEMBRANE PACKAGING

Reverse osmosis membrane is produced in sheet form—up to 60 inches wide and lengths up to 1,500 feet—and as a hollow fine fiber. The asymmetric cellulose acetate was originally produced as a sheet and later as a hollow fine fiber. The asymmetric aromatic polyamide was originally produced as a hollow fine fiber and later in sheet form. The composite membranes with polyamide or polyurea membrane barrier layers are produced in sheet form as of the end of 1987, but research has been and will continue to be done to produce the composite reverse osmosis membranes as a hollow fine fiber.

The Office of Saline Water, U.S. Department of Interior, sponsored much of the development of membrane packaging configurations for reverse osmosis membrane. The configurations which have been developed and evaluated are as follows:

- (1) plate and frame,
- (2) spiralwound,
- (3) tubular and
- (4) hollow fine fiber.

Plate and Frame

The plate and frame configuration is much like the conventional plate and frame filtration concept except that, in reverse osmosis, a typically higher fluid operating pressure is used. Figure 4.7 is a schematic of the plate and frame approach. The membrane package is installed in a pressure vessel designed and fabricated to withstand operating pressures from 400 to 1,000 psig. A stack of parallel porous plates are used to support the membrane on each side of the porous plate. Feedwater under pressure enters the top of the pressure vessel and flows between the parallel stacks of membrane/porous plates. Product water passes through the membrane and into the porous plate to be routed to the product water collection system and then out of the pressure vessel. As the feed/reject stream passes across the membrane, it becomes concentrated and eventually leaves the pressure vessel as concentrate or reject. More advanced designs utilize external tie bolts which hold together sufficiently thick end plates which contain the pressurized feed/reject stream. The plate and frame packaging configuration has an advantage in that only the membrane must be replaced when a membrane becomes defective. It has disadvantages of complex flow patterns and high costs. The plate and frame configuration can achieve a packaging density of up to 150 square feet of active membrane area per cubic foot of pressure vessel. Overall, the plate and frame concept has not been economically competitive and there are no remaining manufacturers of this type of equipment in the United States.

Spiral Wound

The spiral wound membrane packaging configuration is shown in Figure 4.8. Basically, the spiral wound element consists of two sheets of membrane separated by a grooved, polymer reinforced fabric material. This fabric both supports the membrane against the operating pressure and provides a flow path

for egress of the product water. The membrane envelope is sealed with an adhesive on three sides to prevent contamination of the product water. The fourth side is attached to a product water tube which has perforations within the edge seal so that product water can be removed from the porous product water carrier material. The membrane envelope is rolled up around central product water tube, with a plastic mesh spacer between the facing membrane surfaces, in a spiral. The mesh spacer not only serves to separate membrane surfaces but it provides a flow path for and turbulence in the feed/reject stream of each element. The elements have an outer wrap to contain the feed/reject stream in the mesh passageway and brine seal to insure that the feed/reject stream goes through the element and not around it. Spiral wound elements are available in lengths from 12 to 60 inches and diameters from 2 to 12 inches. Packaging densities of up to 300 square feet of membrane to 1 cubic foot of pressure vessel have been attained. Reverse osmosis plants with a capacity from 100,000 to 5,000,000 gpd of product would normally use elements with an 8-inch diameter by a 40-inch or 60-inch length.

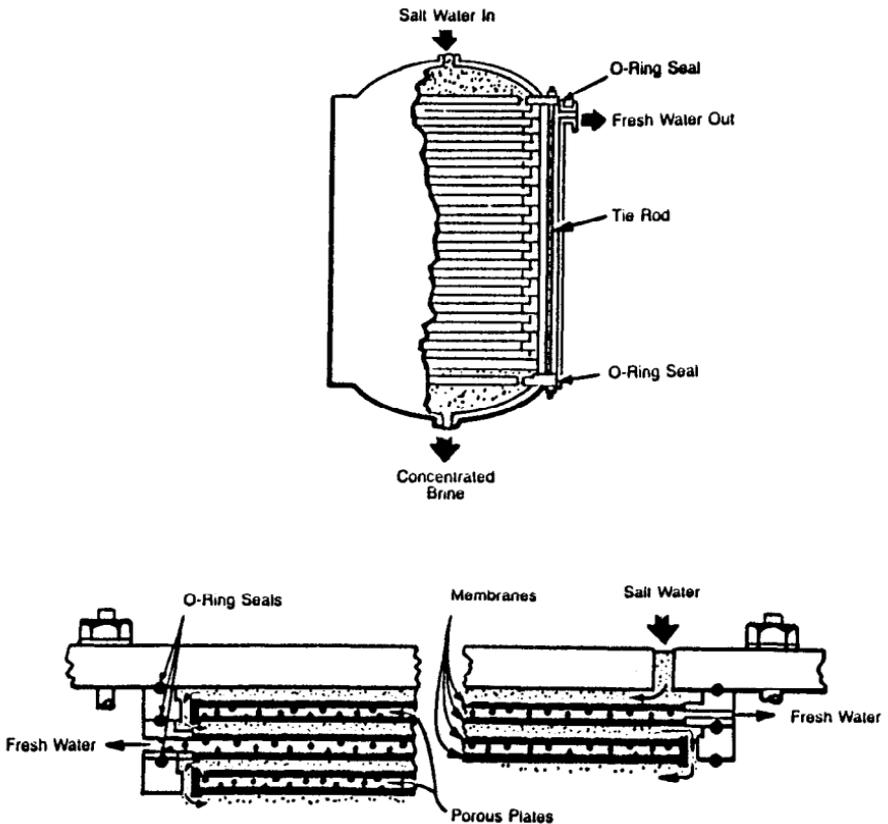


Figure 4.7: Plate and frame packaging.

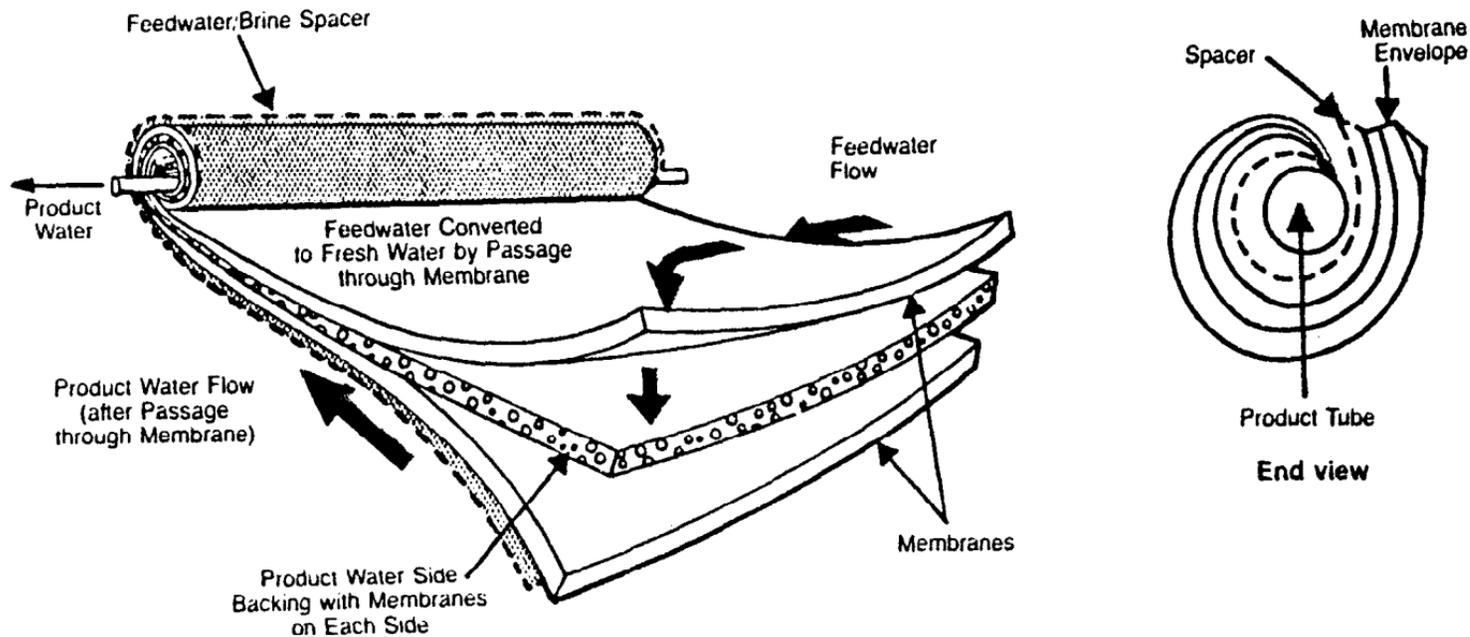


Figure 4.8: Spiral wound packaging.

Spiral wound elements are installed in a pressure vessel which is usually fabricated from fiberglass reinforced plastic. The pressure vessel inside diameter is sized to match the outside diameter of the element brine seal. The pressure vessels are designed and fabricated to accommodate from 1 to 6 elements and operating pressures from 50 to 1,000 psig. Figure 4.9 shows a pressure vessel with 6 elements installed. Feedwater enters one end of the pressure vessel and flows through the first element in which about 10% of the feed permeates through the membrane and flows through the product water carrier material into the product water tube. The reject from the first element flows to and through the second element and the reject from this element becomes the feed to the next element, etc. The reject from the last element is routed from the pressure vessel to the high pressure reject manifold. The first and sixth element product water tubes are sealed to the pressure vessel end caps by O-ring devices to prevent contamination by the high pressure feed or reject to the purified product water stream. The element product water tubes in the pressure vessel are connected to each other with interconnectors which again are O-ring devices whose seals prevent product water contamination. The product water can exit the pressure vessel, usually at near ambient pressures, from either end of the pressure vessel. In a single pressure vessel with six elements, between 40 and 60% of the feedwater to the pressure vessel is recovered as product water from a brackish water feed and from 25 to 35% is recovered from a seawater feed.

The advantages of the spiral wound elements are the high packing density and high flux which makes it one of the most cost effective elements. The disadvantage of the element is that a moderate amount of pretreatment is required for some feedwaters to prevent fouling of the mesh brine spacers.

Tubular

The tubular reverse osmosis device is shown in Figure 4.10. The tube serves as the pressure vessel and the membrane is installed inside the tube. Tubes with inside diameters of $\frac{1}{2}$ and 1 inch have been used. Uniformly porous fiberglass reinforced plastic tubes have been used and nonporous but perforated copper, stainless steel and fiberglass tubes have also been successfully used. The membrane can be bonded to the tube in which case it is cast in situ or the membrane can be loose. The loose membrane is cast in sheet form and a cylindrical section is formed and placed in the tube. Packing densities for the $\frac{1}{2}$ -inch diameter tube are about 100 square feet per cubic foot and about 50 square feet per cubic foot for the 1-inch diameter tube.

Pressurized feedwater enters the tube through an end fitting which seals the membrane to the tube and prevents cross contamination of the product water. The feed water flows down the length of the tube and product water permeates through the membrane and weeps through the tubular pressure vessel into a collection basin. The reject flows through an end fitting and is routed to additional tubes in series or to waste.

The major advantages of the tubular reverse osmosis configuration are the ability to tolerate high suspended solids concentrations in the feed and the possibility of mechanical membrane cleaning. The disadvantages are the excessive number of tube end fittings in proportion to the active membrane area in each pressure vessel, the bulkiness of the reverse osmosis plant and the high cost.

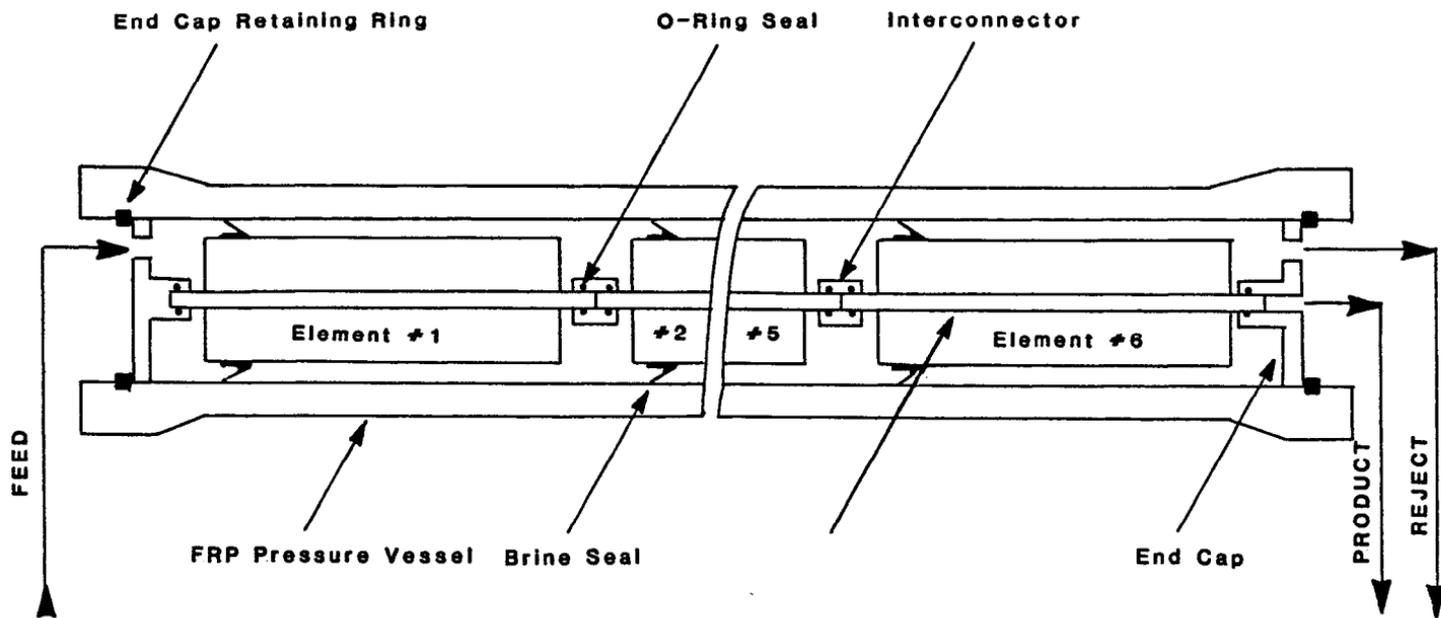


Figure 4.9: Spiral wound element pressure vessel assembly.

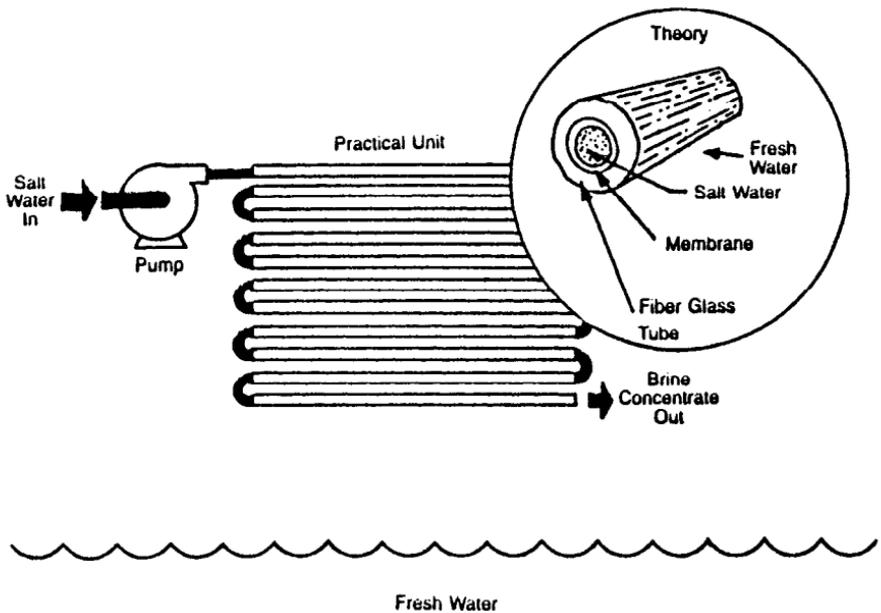


Figure 4.10: Tubular packaging.

There have been a number of attempts to commercialize tubular reverse osmosis systems in industrial applications. As of the end of 1987, there were no large scale tubular reverse osmosis manufacturers in the United States, although there are some in Europe and Japan.

Hollow Fine Fiber

Hollow fine fiber synthetic filaments have been prepared for use in the textile industry for a long time. This well-known technology was adapted to prepare polymers, which are suitable for reverse osmosis desalination, into hollow fine fibers. The most readily available polymers in hollow fine fiber elements are aromatic polyamide, cellulose diacetate and cellulose triacetate. The fibers are indeed very fine in that they approach the diameter of a single human hair. The aromatic polyamide hollow fine fiber has an outside diameter of 85 micrometers for the brackish water fiber and an outside diameter of 95 micrometers for the seawater fiber. Both of the fibers have an internal diameter of 42 micrometers.¹¹ The ratio of outside to inside diameter exceeds two. The fibers are thick walled cylinders that have the compressive strength necessary to withstand the operating pressures. This concept is capable of achieving a high packaging density of 3,000 square feet per cubic foot but the flux is considerably lower than with sheet membrane.

Figure 4.11 is a schematic of a typical hollow fine fiber element. A continuous fiber is looped into a bundle around a central feed tube. The central feed tube is sealed at the product water end and is perforated within the active membrane area. Both ends of the fiber bundle are potted in an adhesive. When cured, the adhesive and protruding fibers in the product water end are machined

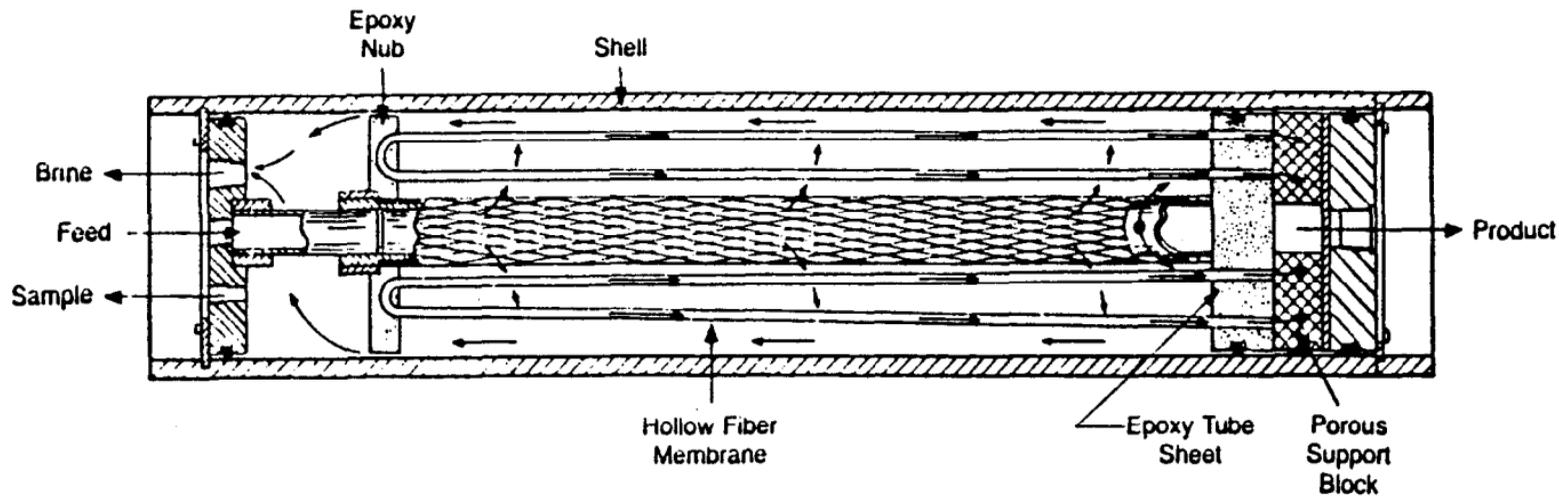


Figure 4.11: Hollow fine fiber packaging.

to open the fibers at that end. The fiber bundle is then installed in a pressure vessel and an O-ring in the product water end tube sheet seals against the pressure vessel wall to prevent the high pressure reject stream from contaminating the low pressure product stream. The element resembles a tube and shell heat exchanger. The pressure vessels are fiberglass reinforced plastic and they are designed and fabricated to accommodate operating pressures from 250 to 1,000 psig. Hollow fine fiber elements are available with nominal diameters of 4, 6 and 8 inches.

Pressurized feedwater enters the element through the central feed tube which is connected to the high pressure feed manifold. The feed flows through perforations in the tube and then radially between the spaces in the fiber toward the outer diameter of the bundle. Longitudinal feed flow is minimized by the reject end being potted. Product water permeates the membrane, enters the capillaries of the fiber bundle and flows to the tube sheet where it is discharged into the product water plenum and then into the product water manifold. The reject flows from the outer diameter of the fiber bundles, through the end cap and to waste via the high pressure reject manifold. Normally, each hollow fine fiber element is capable of recovering from 40 to 75% of the feed water from brackish water and from 25 to 35% from seawater.

The advantages of the hollow fine fiber element are the high packaging density and the elimination of membrane support materials. The prime disadvantage is the need for an efficient feedwater pretreatment to remove suspended and colloidal solids.

Dynamic Membranes

Dynamic membranes are formed by flowing a feed solution containing 50 to 100 mg/l of membrane-forming material tangential to a clean porous surface at velocities from 5 to 50 feet per second under pressures from 500 to 1,200 psi.¹² The dynamic membrane is formed within minutes and the optimum performance is usually achieved within an hour. The membrane-forming materials have been hydrous thorium, natural humic and fulvic acids, pulp mill wastes, municipal wastes, synthetic polyelectrolytes, zirconium oxide, and polyacrylic acid. One of the most successful membrane-forming materials has been a layer of polyacrylic acid deposited on top of a layer of zirconium oxide. The porous surfaces used have been carbon, ceramic and stainless steel tubes. Use of a stainless steel tube which had been pretreated with a filter aid and coated with a layer of polyacrylic acid over a layer of zirconium oxide achieved a sodium chloride rejection of 90% with a flux of about 60 gfd. Recently, several plants using dynamic membranes have been installed and there has been a renewed interest for such applications as orange juice concentration. These membrane promise a high flux with a lower rejection and a low membrane cost. However, the capital and operating costs have not always been commercially competitive.

PLANT DESIGN

The first step in the design of an industrial reverse osmosis plant is to determine the amount of water to be treated, peak demand, product water quality,

source of feedwater and reject discharge requirements. The next step is to obtain an accurate and representative analysis of the feedwater. This analysis should include determination of the concentration of the feedwater constituents shown in Table 4.6. Knowledge of the feedwater constituents in Table 4.6 will provide sufficient information for an experienced design engineer to successfully design a reverse osmosis plant in the majority of applications.

Table 4.6: Feedwater Analysis

<u>Cations</u>	<u>Anions</u>	<u>Other</u>
Calcium	Carbonate	Silica
Magnesium	Bicarbonate	Total Dissolved Solids
Sodium	Sulfate	Total Suspended Solids
Potassium	Chloride	Turbidity
Iron	Nitrate	Silt Density Index
Manganese	Phosphate	Temperature Range
Ammonia	Fluoride	Total Organic Carbon
		pH

The above determinations with the exception of the silt density index should be made in accordance with *Standard Methods for the Examination of Water and Wastewater*.¹³ A schematic diagram of the silt density index apparatus is shown in Figure 4.12 and the silt density index (SDI) is determined as follows:

- (1) Measure the amount of time required for 500 ml of feedwater to flow through a 0.45 micrometer Millipore filter (47 mm in diameter) at a pressure of 30 psig.
- (2) Allow the feedwater to continue flowing at 30 psig applied pressure and measure the time required for 500 ml to flow through the filter after 5, 10 and 15 minutes.
- (3) After completion of the test, calculate the SDI by using the equation below.

$$SDI = \frac{100 (1 - T_i/T_f)}{T_t}$$

where SDI = Silt Density Index

T_t = Total elapsed test time (either 5, 10 or 15 minutes)

T_i = Initial time in seconds required to collect the 500 ml sample

T_f = Time in seconds required to collect the second 500 ml sample after test time T_t (normally after 15 minutes).¹¹

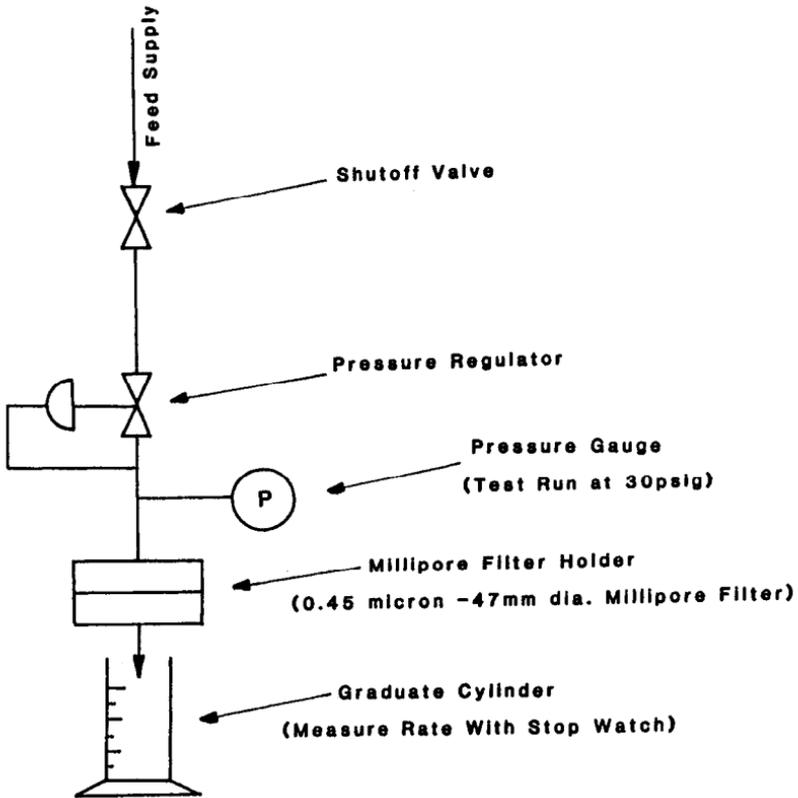


Figure 4.12: Silt density index apparatus.

Manufacturers of hollow fine fiber elements usually require that the pretreated feedwater have an SDI of 3.0 or below in order for the element warranty to be effective. One manufacturer of spiral wound element requires that the pretreated feedwater have a turbidity of less than 1.0 turbidity unit to maintain the element warranty. As a general rule, if hollow fine fiber elements are to be used, the pretreated feedwater should have an SDI of 3.0 or less and, if spiral wound elements are to be used, an SDI of 5.0 or less. However, spiral wound elements have been used to recover municipal wastes with an SDI in excess of 5.0 after pretreatment.

Pretreatment Section

The pretreatment section of a reverse osmosis plant is designed:

- (1) to reduce suspended solids (1 micrometer or larger) to zero and to minimize the effects of colloids in membrane fouling,
- (2) to adjust the pH of the feedwater,
- (3) to add a threshold inhibitor to the feedwater and
- (4) to remove oxidizing compounds in the feedwater if required.

In some applications, such as the desalination of a well water, suspended solids and colloids may be negligible and the pretreatment section may consist of pH adjustment and addition of a threshold inhibitor. At the other extreme is the desalination of surface waters, municipal wastes and industrial wastes which may require all of the steps outlined above.

The following techniques can be used to remove suspended solids or to mitigate the effects of colloids:

- (1) If the feedwater is a municipal water supply or well water with an SDI of 6.0 or below it may only be necessary to filter the feed. There are a number of filters available that have been successfully employed and these are: diatomaceous earth, single media, dual media and mixed media. A properly designed filter in these applications should be able to reduce the suspended solids load to zero.
- (2) In-line coagulation may be used to reduce colloidal membrane fouling. In this technique, coagulant aids such as alum, ferric chloride and/or a polyelectrolyte are added to the feed stream prior to filtration. It is necessary to conduct on-site jar tests in order to determine which coagulant aid is effective and what is the proper dose for that coagulant aid.
- (3) In the case of a feedwater with a high suspended solids, turbidity and/or SDI, it may be necessary to use conventional coagulation followed by sedimentation prior to filtration. Again, on-site jar tests are required to determine the proper coagulant aid, dose and settling rate.

The feedwater pH is usually adjusted to a pH of between 4 and 6 for the following reasons:

- (1) It is required (usually as a condition of warranty) to minimize the rate of hydrolysis of the cellulose acetate ester. Cellulose acetate hydrolysis reduces the useful life of the membrane by increasing the flux and reducing the rejection of the membrane.
- (2) The flux and rejection of some composite membranes are a function of pH and the optimum pH is between 4 and 6.
- (3) Many natural waters are saturated in calcium carbonate which is highly rejected by the membrane. Consequently, it is concentrated in the feed/reject stream during the reverse osmosis process and it will precipitate on the membrane decreasing the flux and rejection. Lowering the pH of the feedwater to between 4 and 6 converts some of the carbonate or bicarbonate ions to carbon dioxide and this prevents carbonate precipitation.

Some feedwaters contain compounds in addition to calcium carbonate that may become saturated in the feed/reject stream and when that stream is concentrated in the reverse osmosis process, these compounds will precipitate on the membrane with a resultant loss of membrane performance. Some of the more sparingly soluble compounds of concern are:

Calcium sulfate	Barium sulfate
Calcium phosphate	Strontium sulfate
Calcium fluoride	Silica

One method of preventing precipitation is to operate the reverse osmosis unit at a recovery which will not concentrate the feed/reject stream to the compound saturation level. Another method of preventing precipitation is to add a threshold inhibitor, such as sodium hexametaphosphate, certain polyacrylates, organophosphates or phosphonates, to the feedwater. The threshold inhibitors are added at the rate of 1 to 5 mg/l of feedwater, and they serve to disrupt the formation of a crystalline precipitate during the residence time of the feedwater in the reverse osmosis unit. In doing this, they broaden the solubility limits of the sparingly soluble compounds.

As noted above, the polyamide and polyurea membranes cannot tolerate an oxidizing agent, such as residual chlorine in the feed. Consequently, if these membranes are used and the feed has a residual chlorine content, then it is necessary to remove it. This is usually done by adding a stoichiometric excess of sodium bisulfite or sodium thiosulfate in accordance with recommendations from the membrane manufacturer.

Even with the best of pretreatment schemes, membrane elements will foul over a long period of time and they must be cleaned. Each membrane packaging configuration has a different degree of susceptibility to fouling and an ease of cleaning.

The tubular element is the least susceptible to fouling and the easiest to clean. This element is widely used in ultrafiltration applications where the process streams contain suspended solids. The tubular element can be cleaned not only by chemical action but also by mechanical means. A sponge rubber ball is pumped through the tubular element with the chemical cleaning solution to scour the membrane surface.

The spiral wound element requires less pretreatment than the hollow fine fiber element or, stated in another manner, it is less susceptible to fouling. For instance, the hollow fine fiber has been tested and found to foul excessively on municipal wastewater reclamation applications while the spiral wound element has been operated successfully. The spiral wound element must be cleaned with chemical solutions, but there are no mechanical means available to clean this element.

The hollow fine fiber element, with the great number of close packed fibers, is an effective filter in itself. Consequently, it is the most easily fouled membrane configuration and requires the most pretreatment. The hollow fine fiber element can be cleaned with chemical cleaning solutions, but it is not amenable to mechanical cleaning. It is also more difficult to clean than the spiral wound element.

Element manufacturers have developed cleaning solutions for their specific membranes and elements. Generally, there are two types of cleaning solutions—one for removal of organic foulants and another for removal of metal hydroxides. A cleaning system should be provided with each reverse osmosis system and the cleaning system should consist of a tank to mix the cleaning solution and a pump with associated piping, valves and instruments to circulate the cleaning solution through the reverse osmosis elements.

Reverse Osmosis Section

Once the pretreatment study had been completed, it will be possible to decide on the type of elements to be used in the reverse osmosis unit. If the SDI of the pretreated feed is 3.0 or less, then either the spiral wound or hollow fine fiber elements can be used. The choice will depend on economics (element price) and desalination characteristics (flux and rejection). If the pretreated feed SDI is more than 3.0, then the spiral wound element should be used. When the decision as to element type is made, then it is appropriate to forward a copy of the pretreated feed water analysis to reverse osmosis element manufacturers to obtain a prediction of product water quality, recommended type of element, total number of elements required, possible problems with sparingly soluble compounds in the feedwater, allowable recovery, and price and delivery.

Figure 4.13 shows a flow diagram for a reverse osmosis unit with 75% recovery on a brackish feed. The pretreated feed is routed to the high pressure pump where the feed pressure is raised to between 250 and 400 psig as required for brackish water desalination. The pressurized feed is then pumped to the first pass pressure vessels where about 50% of the feed is recovered as product and 50% is reject. The reject from the first pass pressure vessels is then routed to the second pass pressure vessels where, again, about 50% of the first pass reject is recovered as product and 50% is reject which is sent to waste. Thus, the overall recovery of the unit is 75% as product. As can be seen, a normal array for a 75% recovery unit is two first pass pressure vessels feeding one second pass pressure vessel or a 2-1 array. If the system recovery were from 40 to 60%, all of the pressure vessels would be in parallel. However, if the system recovery were raised to between 85 and 90%, the pressure vessels would be arranged in a 4-2-1 array.

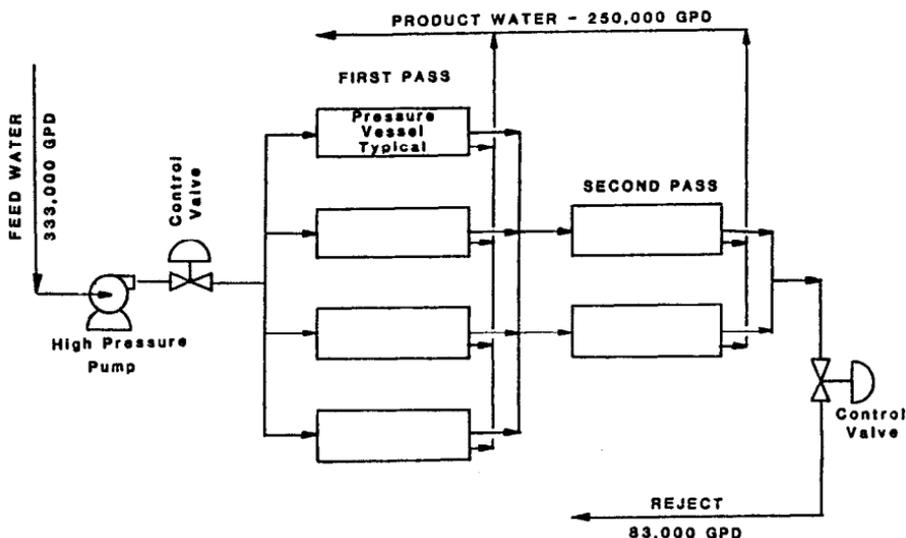


Figure 4.13: Reverse osmosis system.

The quality of the product water is a function of the rejection as shown in the following equation:

$$C_p = C_f * (1 - R)$$

where C_p = Concentration of the product
 C_f = Concentration of the feed
 R = Rejection

This equation produces accurate results for a membrane sample or a small element with a low recovery, e.g., 2% or less. However, a practical reverse osmosis system is designed to recover from 25 to 90% of the feedwater. This means that the concentration of the feed varies throughout the membrane system. At 90% recovery, the initial membranes will have a feed which is about 10 times less concentrated than the feed to the final membranes and the quality of the product water will vary incrementally throughout the system. The product water from the first membrane elements will be less concentrated than the product water from the last elements. The product water from the practical reverse osmosis system is combined in the product water manifold and its concentration is usually represented as the average product water concentration. The average product water concentration is determined by the following formula:

$$\bar{C}_p = \bar{C}_f * (1 - R)$$

where \bar{C}_p = Average or total product water concentration
 \bar{C}_f = Average concentration of the feed and reject streams.

The average concentration of the feed and reject streams as represented by the following equation is not truly representative:

$$\bar{C}_f = \frac{\text{feed concentration} + \text{reject concentration}}{2}$$

As Figure 4.13 shows, there are more elements in the first pass of a 75% recovery reverse osmosis system than in the second pass and the first pass elements produce more product. Thus, a greater percentage of the total product water is derived from the first pass and the total product water will be nearer in concentration to the first pass average than to the second pass average.

Saltonstall¹⁴ has derived an equation to determine the average concentration of the feed and reject streams accurately and this equation is as follows:

$$\bar{C}_f = C_f * \left[\frac{1 - (1 - Y)^{(1 - R)}}{(1 - R) Y} \right]$$

where \bar{C}_f , C_f and R are defined above and Y = system recovery.

The concentration of the total product then becomes:

$$\begin{aligned} C_p &= C_f * \left[\frac{1 - (1 - Y)^{(1 - R)}}{(1 - R) Y} * (1 - R) \right] \\ &= C_f * \left[\frac{1 - (1 - Y)^{(1 - R)}}{Y} \right] \end{aligned}$$

Thus, the product water concentration is dependent on the feed concentration, membrane rejection and system recovery. Figure 4.14 shows the dependence of product concentration on these variables. It is possible to use a lower rejection membrane and a high recovery with a low TDS feedwater. On the other hand, a high TDS feedwater, such as seawater is limited to a low recovery and requires a high rejection membrane.

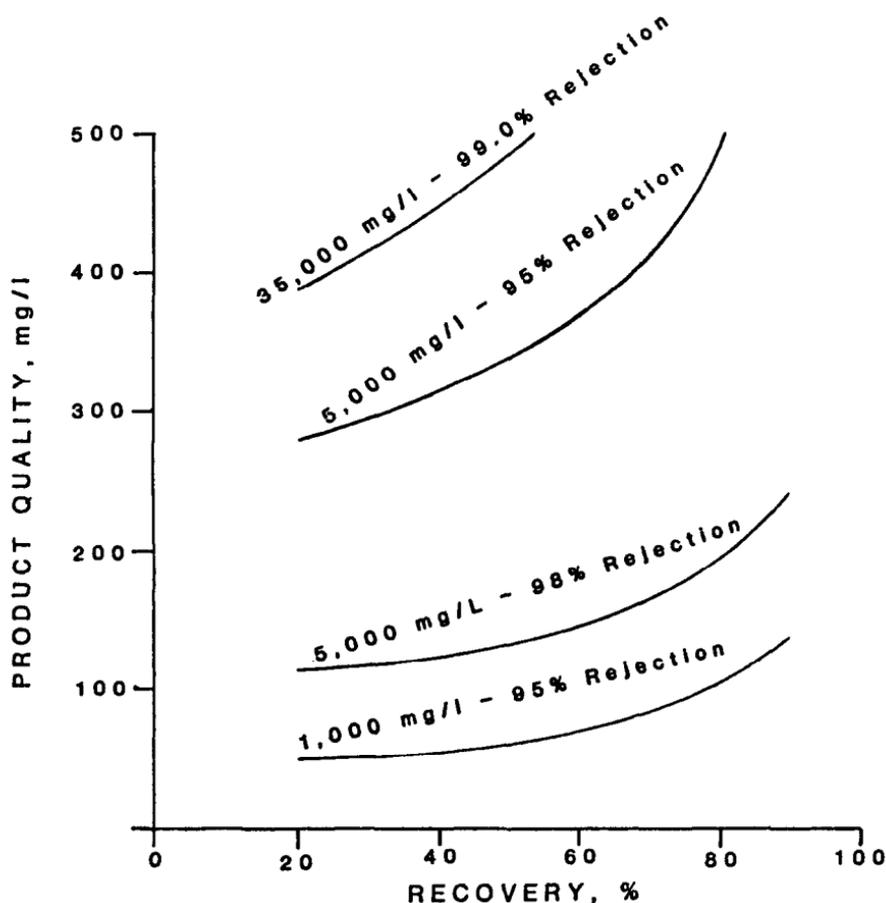


Figure 4.14: Product water vs. recovery.

The membrane flux is also dependent on recovery. The average feed/reject concentration increases with pressure and, since the osmotic pressure is proportional to concentration, the average osmotic pressure will increase with increased recovery. At constant applied pressure, the membrane flux will decrease with increased recovery. Figure 4.15 shows the increase in average osmotic pressure and the decrease in flux for a membrane which would produce 20 gfd at zero recovery operating on a feed of 5,000 mg/l at 450 psi applied pressure. Concentration polarization is assumed to be constant at all recoveries.

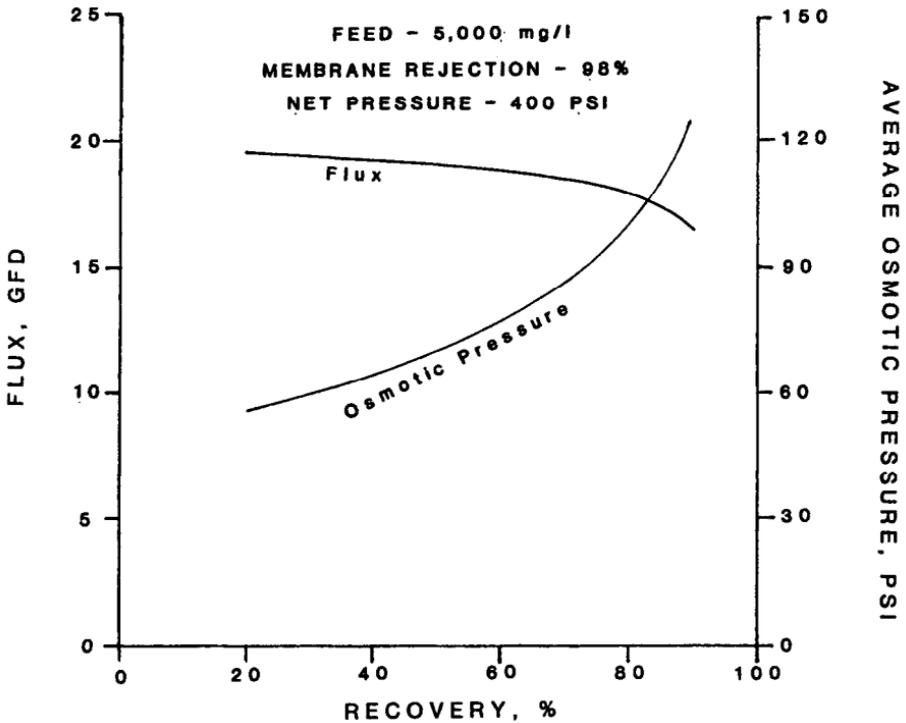


Figure 4.15: Flux and average osmotic pressure vs. recovery.

It is noted that the osmotic pressure of seawater is about 400 psig and an operating pressure of between 800 and 1,000 psig is required to obtain optimum flux and rejection. If a recovery of only 50% were used, the reject osmotic pressure would be in excess of 700 psig and the average osmotic pressure would be about 550 psig. This would present problems in obtaining a potable product of 500 mg/l or less. Consequently, most seawater desalination reverse osmosis units are operated at a recovery of from 25 to 35%.

Since lowering of the feed pH usually means that the feed contains significant quantities of carbon dioxide, the feed and reject streams are extremely corrosive. In addition, the membrane does not reject carbon dioxide and, consequently, the product stream contains about the same concentration of carbon dioxide as the feed and reject streams, i.e., all three process streams are corrosive. General practice is to use PVC pipe and valves for low pressure piping (less than 100 psig) and 316 stainless steel pipe and valves for the high pressure (above 100 psig) process streams. The high pressure pump wetted parts should be either 316 stainless steel, corrosion resistant plastic or other corrosion resistant materials. The pressure vessels are fabricated with fiberglass reinforced plastic which is corrosion resistant.

The reverse osmosis process is relatively simple and instrumentation requirements are minimal. Following is a list of the minimum recommended parameters to be measured in a reverse osmosis system:

<u>Feed</u>	<u>Product</u>	<u>Other</u>
Flow	Flow	Differential Pressure
Temperature	Conductivity	across each array
Pressure		
Conductivity		
pH		

A variety of control schemes can be incorporated in the design of a reverse osmosis plant. However, this subject is beyond the scope of this manual.

Posttreatment Section

In most plants that use reverse osmosis in the preparation of process water, the reject stream is routed directly to waste discharge without any additional posttreatment. In industrial plants that use reverse osmosis to treat industrial wastes, the reject stream may contain valuable materials and this stream would be sent back to the process. In other applications of industrial waste treatment by reverse osmosis, the reject stream may require additional treatment prior to ultimate discharge. In this case, the reverse osmosis unit will have provided a large volume of water that is disposable or can be reused (the product) and a smaller volume of reject which can be treated more economically.

The product water from a reverse osmosis unit will have a low pH and most probably a high concentration of carbon dioxide. The carbon dioxide can be removed and the pH of the product increased by use of a decarbonator. A decarbonator is a packed column in which product water is introduced at the top while either forced or induced air is introduced at the bottom. The air and water flow countercurrently over and around the column packing. The carbon dioxide is stripped from the water and exits from the decarbonator at the top in the air stream. In a well-designed decarbonator, the carbon dioxide content can be reduced to about 5 mg/ℓ in the water effluent.

INDUSTRIAL REVERSE OSMOSIS AT A REFINERY

One of the most innovative industrial uses of reverse osmosis is at the Petromin Refinery in Riyadh, Saudi Arabia. The refinery takes an unusable municipal wastewater, secondary effluent from the Riyadh sewage treatment plant, and by using lime clarification, filtration, reverse osmosis and ion exchange demineralization, it converts that useless waste into the entire process water requirements for the refinery. Figure 4.16 is the process flow schematic for the refinery water treatment plant.

Chlorinated secondary effluent arrives at the pumping station, which is adjacent to the sewage treatment plant in Riyadh, through an open channel which flows to the pump station inlet basins. The 4.63 MGD of effluent is then pumped 19 kilometers to the refinery by either of two full capacity pumps which take suction from the inlet basins. The effluent is pumped through a

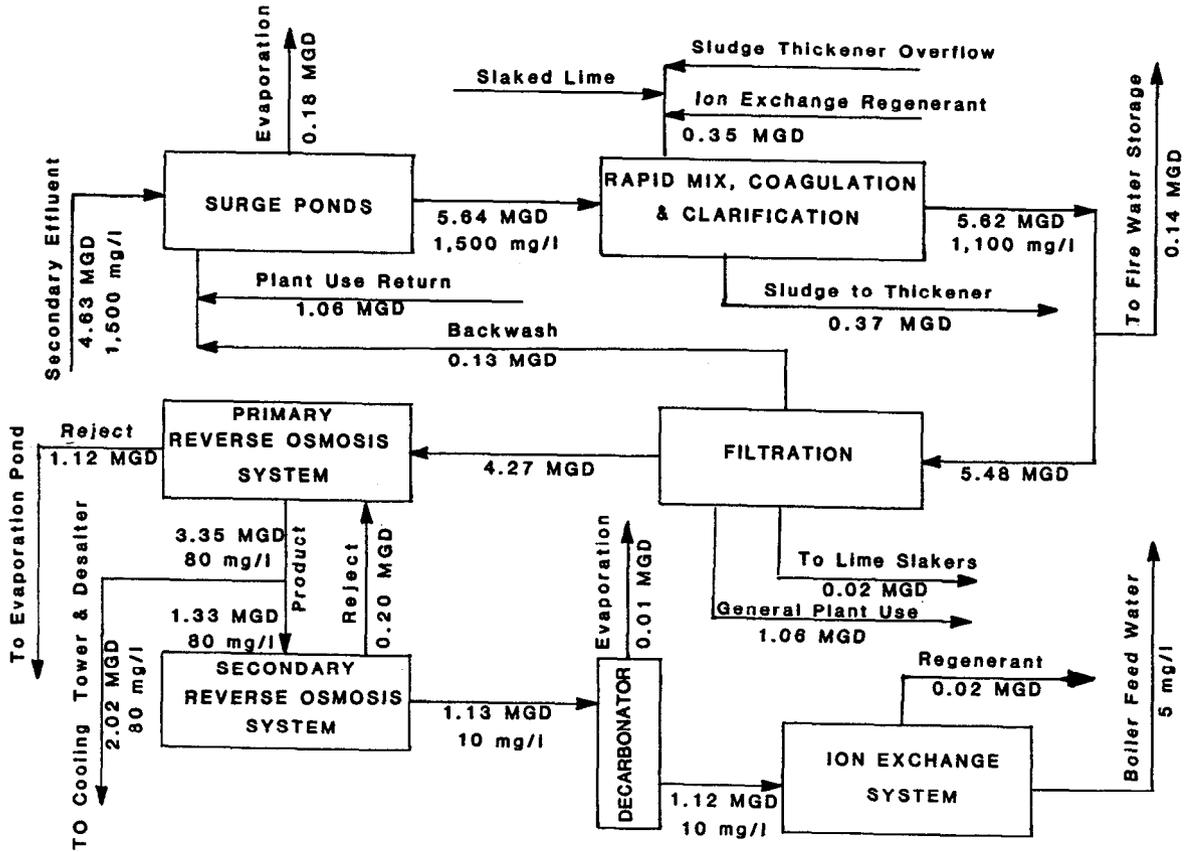


Figure 4.16: Industrial reverse osmosis at a refinery.

24-inch diameter cement lined carbon steel pipeline. The pipeline is equipped with vacuum relief valves and air-release safety valves at the two highest points in the line plus air-release safety valves at the third highest point in the line. The pipeline is also equipped with pigging facilities (for pipeline cleaning) which consist of 2 pigs, a pig launcher and pig receiver.

On arrival at the refinery, the effluent is discharged into a 2 compartment splitting box which divides the flow evenly for discharge into the two on-site surge ponds. The effluent flows from the splitting box by gravity into the surge ponds. At the same time, miscellaneous refinery plant streams and backwash from the refinery water treatment plant filter are returned to the surge ponds and mixed with the secondary effluent. The return stream flow is estimated to be 1.19 MGD with a quality similar to the secondary effluent. Each of the surge ponds is a concrete lined, earth basin with a capacity of 5.3 million gallons. The surge ponds have a static air diffusion system to prevent sewage septicity and improve homogeneity in the secondary effluent. Chlorine can be added to the secondary effluent either at the sewage treatment plant pump station or at the surge ponds at the refinery, if required.

All surface waters and municipal effluents contain suspended solids as well as dissolved solids and the presence of suspended solids dictates the need for a pretreatment section. Experience has shown that effective removal of the suspended solids in pretreatment is a prerequisite to efficient reverse osmosis membrane performance. Suspended solids in secondary effluent are primarily organic in nature and, due to their small size, it is difficult to remove them by settling. Therefore, it is necessary to aggregate the smaller particles into larger particles which can more easily be removed by settling and filtration.

For some time, it has been known that suspended solids or colloids in water possess an electrical charge which is predominantly negative. As a result of this electrical charge, the colloidal particles tend to repel each other and thereby limit the potential for aggregation and the colloidal suspension is said to be stabilized. Before the suspended particles can coalesce, the stabilizing forces must be neutralized and this is usually done by adding chemicals. Trivalent aluminum, ferric hydroxides, and/or lime have been used to destabilize colloidal suspensions and allow suspended particles to grow in size. Another means of particle destabilization is the use of natural or synthetic polymers which are long chain, high molecular weight polyelectrolytes with many active sites. The negatively charged colloids are adsorbed on the active sites with the resultant coagulation or growth of the particles.

The refinery clarification equipment has the capability of adding any of the chemicals mentioned above. However, lime clarification was chosen as the method to be used. The combined secondary effluent and the plant return streams (5.64 MGD) are pumped from the surge ponds to the rapid mix basin in the clarifier. The rapid mix basin has two compartments in series and each compartment has a high speed mixer. Lime and sodium hydroxide are added to the first compartment and these are vigorously mixed with the secondary effluent in both compartments. As a result, the pH of the effluent from the rapid mix basin is raised to between 10.8 to 11.0. At this pH, much of the bicarbonate in the water reacts with the lime and forms an insoluble calcium carbonate and the magnesium in the water reacts with hydroxyl ions to form insoluble magnesium hydroxide.

The water flows from the rapid mix basin to the flocculation basin which has four compartments. Each compartment is further divided into three sec-

tions and all twelve sections are equipped with a slow speed turbine mechanism. The gentle agitation provided in the flocculation basin promotes contact between the calcium carbonate, magnesium hydroxide and suspended matter which results in the formation of larger particles.

The water then flows by gravity from the flocculation basin to the sedimentation or clarification basin. The enlarged floc particles settle to the bottom of this basin and they are removed as a 1% sludge by three circular rakes into a sludge basin or hopper. This amounts to about 0.37 MGD of sludge which is pumped to the sludge thickener tank. The sludge is thickened to about 6% in the sludge thickener and then filtered in a plate and frame filter press to about 50% solids. The filtered sludge may then be transported to a landfill for disposal or sent to the on-site multiple hearth lime kiln for regeneration. Operation of the lime kiln is economically marginal. The overflow from the sludge thickener and the filter press filtrate are returned to the rapid mix basin for reclamation. In addition, part of the water used in the regeneration of the ion exchange demineralizers is also returned to the rapid mix basin for recovery. The total amount of water from these three sources is 0.35 MGD with variable quality.

Effluent from the clarifier is saturated in calcium carbonate and this would precipitate on the filter media to clog the filter which is the next step of pre-treatment. Consequently, the clarifier effluent flows by gravity to the serpentine recarbonator basin where carbon dioxide is added to reduce the pH to between 7.5 and 8.0. The insoluble calcium carbonate and magnesium hydroxide are converted to soluble calcium and magnesium bicarbonate in the recarbonator basin. The 5.62 MGD of effluent from this basin has a suspended solids concentration of about 2 mg/l and a TDS of about 1,100 mg/l.

Following lime clarification and recarbonation, the treated water is pumped to the fire water storage tanks (0.14 MGD) and the remainder (5.48 MGD) is pumped to the water treatment plant cooling tower. The cooling tower is used to reduce reverse osmosis feedwater temperature when required. For a large part of the year, the cooling tower is inoperative. A static mixer is installed in the line upstream of the cooling tower. At this point, sulfuric acid can be added to the process stream to reduce the pH of the water in the event that carbon dioxide is unavailable to the recarbonation system. Chlorine can be added to the process water at the cooling tower influent line or at the chlorine diffuser of the filter aid basin. The water flows by gravity from the cooling tower to the filter aid mix station where it is possible to add coagulant aids to assist in the removal of suspended solids during filtration. Coagulant aids are not being added at the filter aid mix basin at this time. The water then flows by gravity to the dual media filters.

There are four separate filter basins with filter media as described in Table 4.7 below.

Table 4.7: Filter Media Description

<u>Media</u>	<u>Height</u>		<u>Size, mm</u>
	<u>mm</u>	<u>Inches</u>	
Coal	800	31.5	0.6 - 1.6
Sand	400	15.7	0.6 - 1.0
Gravel Support	400	15.7	3 - 35

The process water flows through the filter media by gravity to remove the remaining suspended solids. It then flows through a flow control valve to the filtered water reservoir by gravity. Backwash water overflows from the filtered water reservoir to the wash water storage tank. The filter effluent of 5.48 MGD provides 1.06 MGD of general use water to the refinery, 0.13 MGD of filter backwash water, 0.02 MGD of filtered water to the lime slakers at the rapid mix basin and the remaining 4.27 MGD flows by gravity to the reverse osmosis dechlorination basin.

Chlorine has been added to the feedwater upstream of reverse osmosis pretreatment. However, since chlorine will depolymerize the polyurea membrane barrier layer in the spiral wound element, with subsequent loss of desalination properties, the chlorine is removed in the pretreatment system dechlorination basin. This removal is chemically accomplished by the addition of sodium bisulfite. The chlorine level in the influent and effluent to the dechlorination basin is continuously monitored. The feedwater is then transferred from the dechlorination basin to the cartridge filter feed pumping station by gravity flow and it is then pumped to the cartridge filters.

Sulfuric acid and sodium hexametaphosphate (SHMP) are injected in the feedwater line upstream of the cartridge filters. The sulfuric acid is injected to adjust the feedwater pH to a level of between 4 and 6. The purpose of acid injection is twofold. The primary purpose is to mitigate the possibility of calcium carbonate deposition by conversion of bicarbonate to carbon dioxide. Coincidentally, the rejection performance of the thin film composite membrane is pH sensitive and the optimum performance is at the operating pH level. SHMP is added to the feedwater as a threshold inhibitor to inhibit the crystalline growth of sparingly soluble salts such as calcium sulfate.

After chemical addition, the feedwater is routed to cartridge filters which serve to mix the chemicals which have been added upstream and to insure that any particles that may have escaped the gravity filters, such as sand or other particulate matter is removed. In general, the cartridge filters do not improve the quality of the reverse osmosis feedwater to a large degree and they are not intended as continuous duty filters. The effluent from the cartridge filters is routed to the primary reverse osmosis feed pump wet well.

The primary reverse osmosis system contains five trains of equal capacity. Each train is operated independently and there is a vertical turbine high pressure pump for each train mounted on the primary feed pump wet well. The high pressure feedwater is pumped to the reverse osmosis train where the operating pressure of each train is adjusted by the pressure control valve upstream of the spiral wound element pressure vessels. The elements are located in fiberglass reinforced plastic pressure vessels with six elements per pressure vessel. The elements have nominal dimensions of 40 inches in length by eight inches in diameter and they contain composite membrane with a polyurea membrane barrier layer. Each primary train has 180 elements in 30 pressure vessels which are manifolded in a 20-10 array. The pressurized water is fed to 20 pressure vessels in parallel where about 50% of the feed is recovered as product and the reject from these pressure vessels is manifolded and fed to 10 pressure vessels in parallel. About 50% of the feed to these latter pressure vessels is recovered as product and the reject is manifolded through the flow control valve to waste. The product water from the primary reverse osmosis system is transferred to the cooling tower makeup and

desalt water storage tanks or to the secondary reverse osmosis system feed pump wet well. The overall recovery for each train is 75% and the design capacity of each train is 0.83 MGD of product for a total product water capacity of 4.10 MGD in the first stage. The daily requirements for product water are 3.35 MGD so there is an installed spare capacity of about 20%. The product water requirements are 2.02 MGD for the refinery cooling towers and the desalter and 1.33 MGD for the second stage reverse osmosis system. The reject (1.12 MGD) from the first stage reverse osmosis system is sent to the on-site solar evaporation pond for disposal.

The first stage product water concentration of 80 mg/l is adequate for cooling tower makeup and desalter process water but it is not pure enough for moderate pressure boiler feed. Consequently, the first stage product must be further treated with ion exchange demineralization to achieve the desired purity. A second stage of reverse osmosis is used at the refinery to pretreat the first stage product prior to ion exchange demineralization. An improved quality feed to the ion exchangers will improve not only the quality of ion exchanger effluent but it will reduce the quantity of regenerative chemicals required by the ion exchangers.

The secondary reverse osmosis system contains three trains of equal capacity. As in the primary reverse osmosis system, each train is operated independently and there is a vertical turbine high pressure pump associated with each train. The pressurized feedwater is pumped to the secondary trains where the operating pressure of each train is adjusted by the pressure control valve upstream of the element pressure vessels which are fabricated from fiberglass reinforced plastic. Each secondary train has 21 pressure vessels (126 elements) which are manifolded in a 12-6-3 array. The spiral wound elements in the secondary trains are identical to the elements in the primary stage. The pressurized feedwater is fed to the first 12 pressure vessels in parallel where about 50% of the feed is recovered as product and the remaining 50% is manifolded and fed to the next six pressure vessels in parallel. Product recovery in these pressure vessels is again about 50% and the reject is manifolded and fed to the last three pressure vessels where about 40% of the feed is recovered as product. The reject from the last three pressure vessels is manifolded through the flow control valve in each train to the dechlorination basin. The reject flow of 0.20 MGD from the secondary reverse osmosis trains is of better quality than the feed to the primary reverse osmosis system and, thus, it is used as feed to the primary system. The overall design recovery of each of the secondary trains is 85%. Each of the trains is rated at 0.57 MGD for a total capacity of 1.71 MGD. The total required capacity is 1.31 MGD which can be produced by two trains with one train in standby. The recovered product is routed to the forced air decarbonators in the ion exchange demineralization system.

The product water from the secondary reverse osmosis system contains a high concentration of carbon dioxide as a result of pH adjustment in the primary reverse osmosis system which converts bicarbonate and carbonate alkalinity to carbon dioxide. Since the membrane is "transparent" to carbon dioxide, it passes through both the primary and secondary reverse osmosis systems into the secondary system product water. Although the ion exchange resin would remove the carbon dioxide, it is more economical to do so in the decarbonator.

There are two degasifiers (decarbonators) in the refinery water treatment plant and each of these are in parallel and normally in operation, i.e., one de-

gasifier is not used as a standby. The secondary product is fed to the top of the degasifiers where it is allowed to cascade over the degasifier packing. Air is blown through the packed section from the bottom and it rises in the packed column countercurrent to the water. It is estimated that 0.01 MGD of product water are lost to evaporation in the decarbonator. Since the water is at a low pH, carbon dioxide is transferred from the liquid phase to the gas phase. The result is that the carbon dioxide concentration in the water is reduced to about 5 mg/l. Two transfer pumps, both full capacity, are installed to pump the decarbonated water through the ion exchange system.

There are two full capacity ion exchange trains in the system and each train consists of a cation tank, an anion tank, instrumentation, and the necessary valves and piping for process control. The ion exchange demineralization is accomplished in a two-step process involving treatment with both cation and anion resins in separate process vessels. The water is first passed through the strong acid cation exchange resin (Amberlite IR-20) to exchange the cations for the hydrogen ion. The effluent is then passed through a Type I, strong base anion exchange resin (Amberlite IRA-410) where the anions are exchanged for the hydroxide ion. The result is that cations and anions are substituted by water molecules and a high purity effluent is available for use as boiler feedwater. When the ion exchange capacity of one train is depleted, this train is removed from service, the standby train is placed in operation and the depleted train is regenerated with acid and caustic. About 0.02 MGD of decarbonator effluent are used for regeneration of the resins and much of this is returned to the rapid mix basins for recovery.

REVERSE OSMOSIS AND ION EXCHANGE

The preceding example of a reverse osmosis industrial application at a refinery showed that the process is capable of:

- (1) treating a feedwater with high suspended solids and dissolved solids concentrations;
- (2) reclaiming a water that is considered by most as unusable; and
- (3) developing a number of process streams with different quality requirements.

Reverse osmosis also has been used to treat municipal water supplies for industrial purposes even though these supplies are generally low in turbidity, suspended solids and dissolved solids. A large number of reverse osmosis systems have been installed in industrial plants to prepare industrial process water with municipal water as the feed source. A significant number of these industrial applications are to either replace ion exchange demineralization or to pretreat municipal supplies prior to ion exchange demineralization.

Reverse osmosis systems now commercially available will remove 95% or more of the dissolved solids normally removed in ion exchange, and as will be discussed later, a few that are not. The ionized solutes are not all removed to the same degree by reverse osmosis any more than ion exchange resins have the same effect on all solutes. Divalent and multivalent ions, such as calcium, magnesium, sulfate, iron and manganese, can be rejected to greater than 99%. So-

dium, potassium and chloride are normally rejected to the 95% level or better. The net effect is to reduce the number of regenerations required of the ion exchange columns by a factor of 20 or more. This results in a significant reduction in the amount of waste regenerant solutions that must be disposed of and a material reduction in the dissolved solids that might normally be discharged to the environment. A concentrate or reject is produced by reverse osmosis, but there is little change in the environmental salt budget. Concomitant results are a major decrease in the space and equipment necessary for regenerant storage and an extension of the useful life of the resins, owing to reduced resin attrition. In addition, substances difficult to remove from the resin, which also affect its performance, are greatly reduced or are removed by reverse osmosis pretreatment.

When reverse osmosis is used for preliminary demineralization, the variation in the quality of the ion exchange demineralized water is reduced. The amount of dissolved solids in the feed to the ion exchange beds is 5% or less than when the raw water is fed directly. It is, therefore, obvious that the variation in solids in the finished water will also be less when breakthrough occurs. As a result, the resins may be utilized more efficiently. Additional reliability and control can be gained by measuring the solids concentration following reverse osmosis treatment. This precaution virtually eliminates shock loadings on mixed bed polishing columns. Where small point of use polishing columns are used, such as in microelectronics manufacturing, the danger of a rapid breakthrough becomes considerably reduced, and there is an improvement in production and product quality.

An important factor to be remembered is that in some cases water supplies unsatisfactory for processing to high purity water may be the only sources available. Preliminary demineralization by reverse osmosis will make this water suitable for subsequent demineralization by ion exchange. It is thus apparent that such an economically important factor as plant site location, which may be dependent on the availability of suitable water, can be made more flexible through the use of reverse osmosis. It may be possible now to utilize seawater as a source of industrial process water.

During the 1960's, reverse osmosis was compared with other methods of demineralization. It was indicated in these comparisons that reverse osmosis could not compete favorably with ion exchange at dissolved solids concentrations below 700 mg/l and that its most favorable area of use would be from about 1,200 to 5,000 mg/l dissolved solids. This idea has been totally refuted because some of the most successful applications of reverse osmosis, particularly as part of the process to produce high purity water, have been in treating low dissolved solids water. Water containing 200 mg/l dissolved solids or less has been treated at costs equal to or lower than those of ion exchange alone.

Two substances that are frequently of concern in ion exchange demineralization are silica and organics. The organics are frequently present in natural waters as aromatic polycarboxylic acid derivatives known as humic and fulvic acids. Silica may be the limiting factor in the efficiency of the anionic resins, and (particularly in boiler feedwater applications) the lower the concentration before ion exchange demineralization, the better. Reverse osmosis will frequently produce 90% or greater reductions in total silica concentrations. However, performance should be tested on the specific water to be treated since the results can be variable and the reason for differences between waters is not yet understood.

Where the silica concentrations in the raw water are high, reverse osmosis has been most effective. Even in trace quantities, humic and fulvic acids have been responsible for impairing the life of anionic resins and affecting performance of ion exchange columns. These organics are readily removed by reverse osmosis membranes.

REVERSE OSMOSIS AND POLLUTION CONTROL

Pollution control legislation has made industry aware of the economic penalty for inefficient use of raw materials. The plating industry was one of the first industries to experience this enforced awareness. Loss of raw materials in this industry¹⁵ can result in five distinct costs:

- (1) replacement of the plated material that was discharged to waste prior to initiation of pollution control practices;
- (2) removal of that material from the wastewater after pollution control enforcement;
- (3) disposal of the residue from item 2;
- (4) replacing of the process water (at times quite expensive) lost in wastewater; and
- (5) treatment of the wastewater in a publicly owned treatment works after discharge into a sewage system.

In response to the plating industry's increased awareness of the above costs, plating shops have modified their practices to reduce their material and economic losses. While pollution control was greeted by many with coolness, the resultant cost savings have shown that pollution control is not as onerous as expected.

Reverse osmosis has been installed in many plating shops as a way to resource recovery and minimizing the size of waste treatment equipment and volume. Reverse osmosis is particularly suited to waste treatment in the plating industry because most of the toxic ions in the plating solutions are well rejected by commercially available membranes. Table 4.8 shows the rejection ranges of some of the more common toxic materials in the plating industry.¹⁶

Table 4.8: Reverse Osmosis Rejection Range

<u>Ion</u>	<u>Rejection Range, %</u>
Nickel	98 - 99
Copper	98 - 99
Cadmium	96 - 98
Chromate	90 - 98
Cyanide	90 - 95
Zinc	98 - 99

The nickel plating industry is a typical candidate for the use of reverse osmosis in pollution control. Figure 4.17 shows a schematic of this industrial application. The workpiece travels from the plating bath with a concentration of 270,000 mg/l to the rinse tanks. There are three rinse tanks in series and rinse water flows countercurrent to the workpiece. The work piece drags out plating bath to the first rinse tank, first rinse tank solution to the second rinse tank and second rinse tank solution to the third rinse tank. Consequently, the first, second and third rinse tanks have concentrations of 3,000 mg/l, 333 mg/l and 37 mg/l, respectively.

About 100 gallons/hour (GPH) are pumped from the first rinse tank through a cartridge filter and into a reverse osmosis unit. The reject stream contains 99% (59,400 mg/l) of the nickel in the feed stream with 1% (32 mg/l) remaining in the product stream. The reject stream is routed through an activated carbon column to the plating bath. The reverse osmosis product stream is combined with 5 GPH of tap water makeup, which is added to compensate for surface evaporation in the plating tank, and the combined stream is returned to second rinse tank. The waste stream (10 GPH) is sent to waste treatment which is a precipitation process.

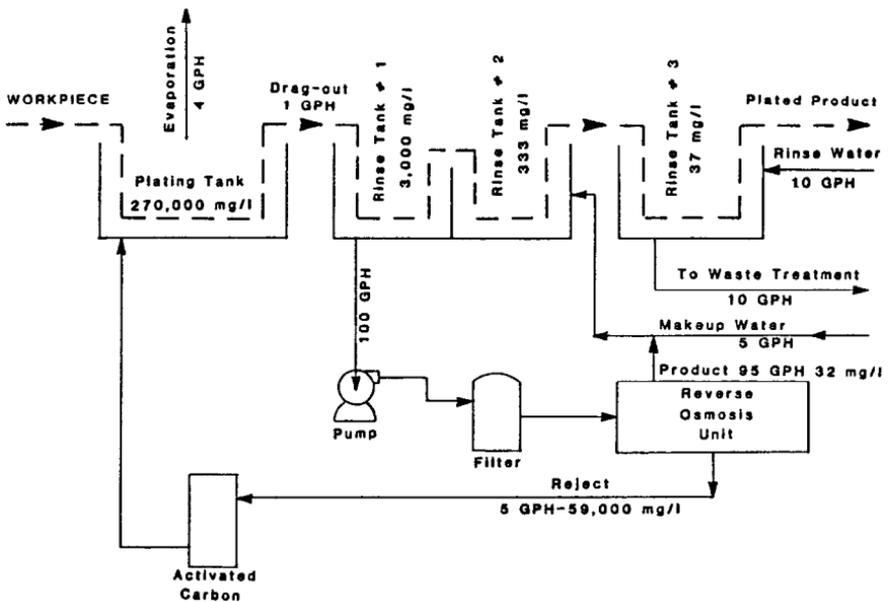


Figure 4.17: Reverse osmosis in nickel plating.

REVERSE OSMOSIS AND SEAWATER DESALINATION

It has been estimated that the oceans cover about 70% of the earth's area and contain about 80% of the water on or in the earth. Seawater concentration varies from 25,000 mg/l of total dissolved solids up to over 50,000 mg/l in the Arabian Gulf. The average composition of seawater is about 35,000 mg/l with the major ions being shown in Table 4.9.

Table 4.9: Major Constituents of Seawater

<u>Constituent</u>	<u>Concentration, mg/l</u>
Sodium	10,500
Magnesium	1,350
Calcium	400
Potassium	380
Chloride	19,000
Sulfate	2,700
Bicarbonate	142
Bromide	65
Other Ions	33
Total	34,570

While seawater is abundant, it is to a large degree unusable and it is the constant dream of the desalination industry to desalt seawater in an economical manner that would allow the product to be used for agricultural purposes. This goal is a long way from reality. On the other hand, distillation and now reverse osmosis are desalting seawater efficiently enough to be used for preparing potable waters in affluent areas. Seawater, with a concentration of 35,000 mg/l, has an osmotic pressure of almost 400 psig and this mandates a membrane element with high pressure capabilities. If the reverse osmosis seawater plant were required to deliver a product water of potable quality (500 mg/l or less), then the membrane element would require a minimum rejection of 98.6% to attain that product quality at 0% recovery. At higher recoveries, both pressure and rejection must increase to obtain potable quality. Spiral wound and hollow fine fiber elements have been developed with the capability of operating at 800 to 1,000 psig and sodium chloride rejections up to 99.5%. This capability has placed reverse osmosis in the seawater desalination business and reverse osmosis is beginning to replace distillation as the result of the clear cut economic advantage of reverse osmosis.

An example of reverse osmosis seawater desalination for industrial purposes is the system installed in a thermoelectric power plant in Venezuela in 1980.¹⁷ The original segment of the plant is designed to produce 800,000 GPD of boiler feedwater and potable water. A process flow diagram for this system is shown in Figure 4.18.

Seawater is pumped from an intake channel to the roughing filters which are the first part of the pretreatment process. Ferrous chloride and chlorine are added to the filter influent line. The chlorine disinfects the seawater and oxidizes the ferrous ion to the ferric ion which forms insoluble ferric hydroxide. The ferric hydroxide acts as a coagulant aid. A large percentage of the coagulated particles are removed in the roughing filter which reduces the feedwater SDI from 15 to about 3. There are five gravity roughing filters which have a dual media of sand and anthracite. Four of the filters are in operation while one is

in standby to be placed in operation when one of the operating filters is back-washed.

The effluent from the roughing filter clearwell is pumped to the pressurized polishing filters which further reduce the SDI to less than 3. There are four operating polishing filters and one standby filter. The filter media is anthracite and green sand.

Sodium bisulfite is added to the effluent from the polishing filters to remove residual chlorine which would be harmful to the polyamide reverse osmosis membrane. Sulfuric acid is then added to the filter effluent to adjust the pH to about 6.5 and the pretreated water is routed to the primary reverse osmosis system.

The first stage reverse osmosis systems consist of four trains which are capable of producing 200,000 GPD each of product water. Each train consists of a 5- μ cartridge filter with a high pressure pump feeding two subunits of 25 DuPont B-10 permeators each plus the necessary piping, valves and instrumentation. The first stage system reverse osmosis units are operated at 30% recovery with a pump discharge pressure of 900 psig. The total dissolved solids content of the feed was reported to be 37,000 mg/l and the product concentration from this stage was 490 mg/l. This product water is sent to a 262,000-gallon storage tank which provides feed to the two second stage reverse osmosis trains and it is the source of potable water for the power plant. The reject from the first stage is routed to waste. A high pressure recovery turbine is not used and a considerable amount of energy is wasted.

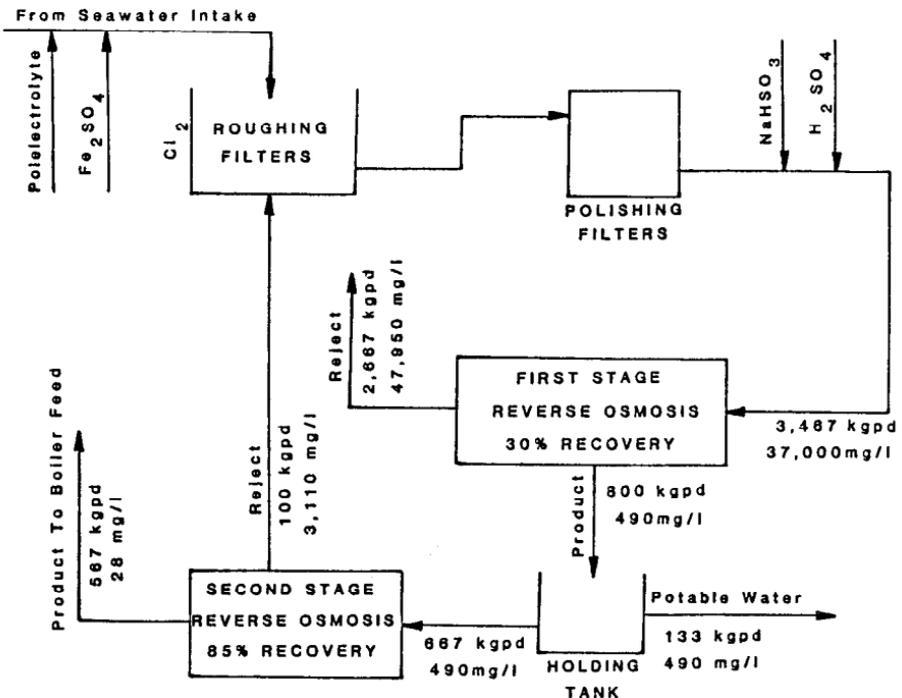


Figure 4.18: Power plant process water from seawater.

The first stage product quality meets potable standards but it is far from the quality required for boiler feedwater. Preparation of boiler feedwater is done in an ion exchange system and, with an influent of 490 mg/l, the operating cycle would be short and chemical regenerant cost high. In this case, it is economical to further treat the first stage reverse osmosis product in a second stage reverse osmosis system to further decrease the load on the ion exchange resins.

About 85% of the first stage reverse osmosis system product water is pumped to the second stage reverse osmosis system which consists of two trains. Each train consists of a 5- μ cartridge filter, a high pressure pump, DuPont B-9 permeators, piping, valves and instrumentation. Each train has 22 permeators in a 12-7-3 array and it is operated at 85% recovery. The product from the second stage has a TDS of 28 mg/l and this is sent to the ion exchange system. The reject from the second stage is of much better quality than the incoming seawater and this is routed to the roughing filter clearwell.

The reverse osmosis system has operated well and the capacity of the plant has been increased to twice that of the initial segment described above.

GENERAL APPLICATIONS OF REVERSE OSMOSIS

The above applications were specific examples that were chosen to demonstrate the versatility of the reverse osmosis process. The following list enumerates a wide variety of applications for which the reverse osmosis process may be considered:

- (1) Municipal Potable Water
 - General quality improvement of present supplies
 - Upgrade total municipal supply
 - Potable water from degraded supplies
 - Brackish water desalination
 - Seawater desalination
 - Removal of nitrates, fluorides, heavy metals, etc.
 - Bottled water production
- (2) Industrial Water
 - Provide usable water where none available
 - Brackish water desalination
 - Seawater desalination
 - Pure water production
 - Industrial rinse waters
 - Food industry
 - Electroplating
 - Power plant boiler feed
 - Beverage production
 - Medical
 - Ultrapure water production
 - Pharmaceutical

- Electronics
- Medical
- (3) Municipal Wastes
 - Reclaim municipal wastewaters (sewage) for
 - Ground water recharging
 - Agricultural or landscape irrigation
 - Industrial process water
 - Improve effluent quality to meet discharge requirements
- (4) Industrial Wastes
 - Reclaim industrial wastewaters for
 - Reuse within industrial plant
 - Zero discharge
 - Agricultural or landscape irrigation
 - Removal of toxic substances prior to discharge
 - Resource recovery
 - Separate or concentrate valuable materials
 - Electroplating industry
 - Dairy industry
- (5) Miscellaneous
 - Production of pure water for high value crops
 - Recovery of agricultural irrigation drainage
 - Production of shipboard drinking water

COSTS OF REVERSE OSMOSIS

There is no such thing as the typical cost of a reverse osmosis system or of the product water from that system as these costs depend on a number of things:

Economic conditions	Power costs
Market conditions	Site preparation
Plant size	Product quality requirements
Local labor rates	Feedwater availability
Chemical costs	Accounting procedures

The DuPont Permasep Engineering Manual¹¹ has published a "guide" for the capital, operating and maintenance costs for both a brackish water system and a seawater system. The brackish water system costs are shown in Table 4.10. They are based on a large brackish water system built in the southern United States in 1982. The estimated capital cost of the plant is \$1.25 per gallon per day of product water installed. This cost includes the cost of wells, a reverse osmosis system with pretreatment, a building for the reverse osmosis systems and office. The above installed capacity cost does not include the cost of land nor an independent power source.

Table 4.10: Total Water Cost for Brackish Water RO

	<u>Cost/1,000 Gals. Product</u>
Energy (0.06/KWH)	\$0.36
Chemicals	0.09
Labor	0.12
Maintenance and Repair	0.05
Membrane Replacement	0.10
Amortization (12%/20 years)	<u>0.48</u>
Total	\$1.20

Data are provided for a seawater system that will produce 10 million gallons per day of product water. The system is a two stage system similar to the one shown in Figure 4.18 where the product water from the first stage high pressure seawater system is treated in a second stage lower pressure brackish water system. The product water TDS is 200 mg/l. The estimated 1982 cost for such a system was \$45 million which includes the RO system with pretreatment and a building for the RO system, controls and office. It does not include the cost of land or an independent energy source. The total water cost for such a plant is shown in Table 4.11.

Table 4.11: Seawater RO System

	<u>Cost/1,000 Gals. Product</u>
Energy (\$0.06/KWH)	\$1.80
Chemicals	0.14
Labor	0.19
Maintenance	0.22
Membrane Replacement	0.90
Amortization (12%/20 years)	<u>1.75</u>
Total	\$5.00

Both the brackish and seawater reverse osmosis product water costs are based on 1982 costs and they are indicative of specific plants in an assumed location in the southern United States. The cost of energy in the seawater system assumes that the reject from the first stage high pressure reverse osmosis system is sent to an energy recovery system which reduces the overall energy requirements for the total system by 31%.

FUTURE PROJECTIONS

The initial projections of 20 years ago have proven to be unrealistic in that reverse osmosis has not caused deserts to bloom, nor does every household contain a reverse osmosis unit to improve the tap water. Yet, the process has been of economic value in providing process water to industry, potable water to high income arid regions and a method of reclaiming municipal and industrial wastes. As of 1985, it was estimated that the worldwide market for reverse osmosis membrane elements (not total systems) was about \$50 million.

The future of the process rests in the research that will result in the development of a product that will compete with and win over other separation processes that can do the same thing. Reverse osmosis has gone a long way toward becoming more economical than multistage flash distillation in the production of potable water from seawater. The development of membranes with higher fluxes and improved rejections will be incorporated into single stage seawater plants with lower operating pressures. This will not only permit reverse osmosis to win the competition for new seawater desalination capacity but it will allow the RO process to replace distillation plants that are being retired in the Middle East where the enormous capacity of these plants exists. The low pressure membranes which are being developed for brackish applications will also further potential municipal wastewater reclamation as an alternate source of potable and industrial water. The reduced costs which will result from lower pressure/higher rejection systems will effectively compete with pipelines, dams and other water supply schemes that proliferate in the western United States.

While not totally essential to progress, it is suggested that oxidizing-agent-resistant membranes will be developed in a thin film composite membrane. This will broaden the applications available to RO and, at the same time, reduce the cost and complexity of existing plants which presently use chlorine sensitive membranes.

An elusive goal has been the development of ion specific membranes. Based on limited knowledge of worldwide research programs, it appears that this goal will remain elusive for the foreseeable future.

Finally, as the world becomes more aware of the environmental damage caused by indiscriminate waste disposal it is apparent the RO process will play a key role in mitigating that problem. It appears that the next market for the RO process will be in industrial waste treatment in the United States to be followed by application in other countries. Eventually, the world will be forced to use reverse osmosis to reclaim municipal wastewater on a large scale and to put the reclaimed water to a number of already demonstrated beneficial uses.

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