CNWRA-EAA-01

PRELIMINARY FEASIBILITY ASSESSMENT OF EDWARDS AQUIFER SALINE WATER TREATMENT AND USE

Prepared for

Edwards Aquifer Authority
San Antonio, Texas

Prepared by

Center for Nuclear Waste Regulatory Analyses
Southwest Research Institute
San Antonio, Texas

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EXECUTIVE SUMMARY

The Edwards Aquifer is the primary source of water for more than one million people in and near San Antonio, Texas, and supplies potable water as well as water for agricultural, industrial, and recreational purposes throughout an 8,000-square-mile area of south-central Texas. In the last five decades, increased population growth and water demand in San Antonio and surrounding areas have resulted in dramatic increases in average pumping from Edwards Aquifer wells. Water resource planners project that by the year 2020, demand for water in the Edwards Aquifer region could be 863,000 acre-feet per year. Water authorities and other entities in the Edwards Aquifer region, as well as in other parts of Texas, are concerned about potential shortfalls in future water supplies relative to future needs. For example, the 2002 Texas State Water Plan warns that about 900 cities and water user groups in Texas could face water shortages during droughts within the next fifty years unless demand for water is reduced and/or additional water sources are developed. In response to these concerns, public and private entities in Texas are evaluating and implementing desalination projects as alternative water sources. Desalination is a process that removes substances and minerals from brackish (or saline) ground and surface water, or seawater. Desalination is recommended in the 2002 State Water Plan as a water management strategy to produce additional water supplies in several regions of Texas, including the south-central Texas region. The technology is particularly promising for the Edwards Aquifer region because of the presence of an extensive saline-water zone in the Edwards Aquifer, downdip of the fresh water zone. The saline-water zone has not been tapped as a water source due to the high dissolved solids concentration (1,000 to over 10,000 mg/L), but it has the potential to be an important source of water if desalination technology becomes an economically viable choice.

As demand for water in the Edwards Aquifer region starts to exceed the availability of fresh water from the aquifer, desalination of water pumped from the saline-water zone likely will become increasingly important. This document was prepared for the Edwards Aquifer Authority and presents a preliminary study on the feasibility of pumping and treating saline water from the Edwards Aquifer to produce potable water. The report provides background technical information on reverse osmosis (RO) and electrodialysis reversal (EDR)—the dominant desalination technologies for municipal water treatment plants, and summarizes information on existing and planned municipal water treatment plants employing similar technologies. Preliminary cost curves are calculated for the treatment of saline waters as a function of operating parameters. The estimated costs are compared with cost data from existing plants and with estimated costs taken from the literature or provided by a vendor. In addition, preliminary estimates of potential yield from wells that may tap the Edwards Aquifer saline-water zone are made to determine if sufficient saline water could be drawn from the aquifer for use as feedwater for the desalination facility.

ACKNOWLEDGMENTS

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1 INTRODUCTION

The Edwards Aquifer is the primary source of water for more than one million people in and near San Antonio, Texas, and supplies potable water as well as water for agricultural, industrial, and recreational purposes throughout an 8,000-square-mile area of south-central Texas. About 54 percent of that water is used for municipal supply. For example, San Antonio obtains its entire municipal water supply from the Edwards Aquifer and is the largest U.S. city to rely on a single groundwater source. In addition, the Edwards Aquifer is the sole-source of water for a unique system of aquatic life, including several threatened and endangered species.

EAA Hydrologic Data Report

In the last five decades, increased population growth and water demand in San Antonio and surrounding areas have resulted in dramatic increases in average pumping from Edwards Aquifer wells. In San Antonio alone, population increased from about 200,000 people in 1940 to more than one million in 1990. Regional pumping, which was 100,000 acre-feet[1] of water per year in 1934, reached a maximum of 542,000 acre-feet of water per year in 1989 (Votteler, 2002). In the 1990s, the amount of water pumped from the aquifer ranged from 327,000 acre-feet in 1992 to 493,000 acre-feet in 1996. This increased demand for water was exacerbated by periods of drought. For example, in the 1950s, a seven-year drought drastically lowered water levels in the aquifer, while in the 1980s and 1990s, droughts of shorter duration occurred that required heavy pumping from the aquifer. Water resource planners project that by the year 2020, demand for water in the Edwards Aquifer region could be 863,000 acre-feet per year. Recharge of the aquifer, which is entirely dependent on rainfall, is not expected to keep up with this demand. The average recharge recorded from 1934 to 2000 was only 680,000 acre-feet per year (http://www.edwardsaquifer.net).

It is expected that the Edwards Aquifer will continue to be the primary source of water for the region. Water authorities and other entities in the Edwards Aquifer region are concerned about potential shortfalls in future water supplies relative to future needs. This concern is not unique to the Edwards Aquifer region. The 2002 Texas State Water Plan (Texas Water Development Board, 2002) states that supplies from existing water sources in Texas are expected to decrease by 19 percent, from 17.8 million acre-feet per year in the year 2000 to 14.5 million acre-feet per year in 2050, whereas municipal and manufacturing demands are projected to increase by 67 and 47 percent, respectively, between 2000 and 2050. Thus, the 2002 State Water Plan warns that about 900 cities and water user groups in Texas could face water shortages during droughts within the next fifty years unless demand for water is reduced and/or additional water sources are developed.

In response to concerns about shortfalls in future water supplies, public and private entities in Texas are evaluating and implementing desalination projects as alternative water sources. Desalination is a process that removes substances and minerals (including, but not limited to salt) from brackish (or saline) ground and surface water, or seawater. Desalination systems can be membrane-based, such as reverse osmosis (RO) and electrodialysis reversal (EDR), or thermal-based, such as multi-stage flash and multiple-effect distillation. Membrane-based systems work by forcing saline water through a semi-permeable membrane, which restricts salt and other minerals, but allows water molecules to pass through. Thermal methods involve heating the saline water to produce steam, which is then condensed to produce water with a low salt concentration and few of the other impurities contained in the original water. Depending on the technology used, the final water product is generally high quality, ranging from 1 to 500 milligrams per liter (mg/L) total dissolved solids.[2]

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[1] One acre-foot of water is equal to 325,850 gallons of water. One acre-foot is approximately the amount of water needed to supply two families of four with enough water for one year.

[2] The recommended drinking water standard in Texas for total dissolved solids is ≤1,000 mg/L.

As of 1996, 76 desalination plants were operating in Texas and were producing 40 million gallons per day (MGD) of purified water for municipal and industrial uses from inland brackish surface water and groundwater (Krishna, 2002; Sharp, 1998). Municipal desalination accounted for about 23 MGD whereas industrial desalination was approximately 17 MGD. More than half (49) of the desalination plants in Texas use the RO method, 15 plants use the EDR process, and the others use the distillation process (Sharp, 1998). Newer plant installations use the membrane process because it is more efficient compared to thermal systems.

As demand for water increases and starts to exceed the supply of fresh water, desalination likely will become increasingly important for the Edwards Aquifer region and for the state of Texas. In the short term, a potentially important factor is the cap on aquifer withdrawals set by the Texas legislature[3] that will reduce total withdrawals from the current 450,000 acre-feet per year to 400,000 acre-feet annually by 2008.[4] Desalination is recommended in the 2002 State Water Plan as a water management strategy to produce additional water supplies in several regions of Texas, including the south-central Texas region. The technology is particularly promising for the Edwards Aquifer region because of the presence of an extensive saline-water zone in the Edwards Aquifer, downdip of the fresh water zone. The saline-water zone has not been tapped as a water source due to the high dissolved solids concentration (1,000 to over 10,000 mg/L). However, it has the potential to be an important source of water if desalination technology becomes economically viable and if withdrawals from the saline zone are not counted towards the cap on aquifer withdrawals.

This document, *Preliminary Feasibility Assessment of Edwards Aquifer Saline Water Treatment and Use*, was prepared for the Edwards Aquifer Authority and presents a preliminary study on the feasibility of pumping and treating saline water from the Edwards Aquifer to produce potable water. The report provides background technical information on RO and EDR—the dominant desalination technologies for municipal water treatment plants, and summarizes information on existing and planned municipal water treatment plants employing similar technologies. Preliminary cost curves are calculated for the treatment of saline waters as a function of operating parameters. The estimated costs are compared with cost data from existing plants and with estimated costs taken from the literature or provided by a vendor. In addition, preliminary estimates of potential yield from wells that can tap the Edwards Aquifer saline-water zone are made to determine if sufficient saline water could be drawn from the aquifer for use as feedwater for the desalination facility.

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[3] Senate Bill 1477, or the Edwards Aquifer Authority Enabling Act, adopted by the Texas Legislature on May 30, 1993. The Act created a conservation and reclamation district, named the Edwards Aquifer Authority, which was charged with regulating groundwater withdrawals pursuant to the Conservation Amendment in the Texas Constitution, Article XVI, § 59, replacing the rule of capture in five counties and portions of three others, with a permit system.

[4] However, the Edwards Aquifer Authority may seek to raise the authorized pumping limit to 500,000 acre-feet or more annually (Votteler, 2002).

2 DESCRIPTION OF MEMBRANE DESALINATION TECHNOLOGY

To understand how membrane desalination systems operate, it is useful to know some general terms common to both RO and EDR systems. Feedwater is the source of influent water to the selected membrane process. A membrane can be defined as a thin film separating two aqueous phases and acting as a selective barrier to the transport of matter. Its performance depends on its structure, on the nature of the constituents present in the aqueous phases, and on the applied driving force. The feedwater is separated into two streams at the membrane — permeate and concentrate. The permeate is the stream that has passed through the membrane and is the demineralized product water. The concentrate stream contains the dissolved materials that were removed from the feedwater by the membrane. The total dissolved solids concentration of the concentrate stream is much greater than that of the permeate stream. Water recovery is the percent of feedwater recovered as product water. Salt rejection quantifies the reduction in total dissolved solids concentration from the feedwater to the product water.

Both the RO and EDR desalting systems are capable of removing high percentages of dissolved salts. RO systems use pressure to force pure or nearly pure water to pass through the semi-permeable membrane while retaining dissolved minerals from the feedwater on the concentrate side of the system. EDR systems use membranes that are made from ion-exchange resin materials supported by an open weave cloth cast in the resin for physical strength. In contrast to RO systems, EDR is electrically driven. When an EDR membrane is subjected to electrical current, the solution on one side of the membrane becomes partially desalted while the solution on the other side becomes more concentrated.

The RO and EDR desalination processes have several characteristics in common (HDR Engineering, Inc., 2000). Both require some form of pretreatment. At a minimum, pretreatment will include cartridge filtration and chemical conditioning. Chemical treatment may include chlorination, pH adjustment, and scale control. Water produced for domestic use also undergoes post-treatment to ensure that the product water meets the health standards for drinking water as well as recommended aesthetic and anti-corrosive standards. Pure desalinated water has low pH and is thus corrosive to pipes, so it has to be mixed with other sources of water that are piped onsite or otherwise adjusted for pH, hardness, and alkalinity before being piped offsite. The product water generally is more pure than drinking water standards, so when the product water is intended for municipal use, it is usually mixed with water that has higher total dissolved solids concentration. Furthermore, RO and EDR processes produce concentrate that requires disposal. In using these desalination technologies, there is a need to optimize the system recovery to minimize the total cost, including concentrate disposal. Also, the membranes used in RO and EDR systems require careful monitoring and routine maintenance, including cleaning. All membranes have a finite useful life and must be periodically replaced.

The following sections provide details on RO and EDR desalination processes.

2.1 REVERSE OSMOSIS

Osmosis is the phenomenon of water flow from a dilute (low concentration) solution to a more concentrated solution through a semi-permeable membrane (Figure 2-1a). A semi-permeable membrane allows water molecules to pass through but prohibits the solute (i.e., dissolved material) from passing through. The flow of water produces a measurable pressure, termed osmotic pressure. This flow may be stopped, or even reversed, by applying an external pressure on the side with the higher concentration (Figure 2-1b). If the external pressure is slightly lower than the osmotic pressure, water will flow from the dilute side to the concentrated side. If the external pressure is equal to the osmotic pressure, no water flow across the membrane will occur. However, if the applied pressure is higher than the osmotic pressure, it will force water to flow from the more concentrated side, through the membrane, to the dilute side. This phenomenon is called reverse osmosis.

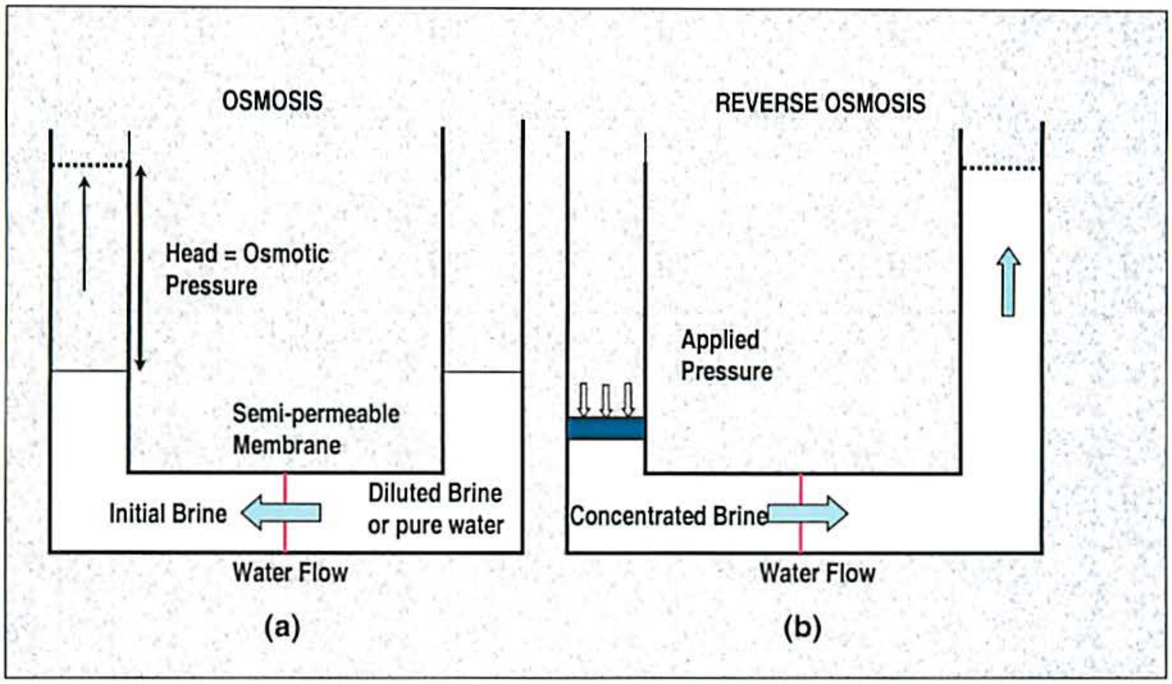


Figure 2-1. Schematic of osmosis and reverse osmosis. Figure taken from Morales and Barrufet (2002).

The osmotic pressure and, thus, the external pressure required to overcome it, increases with the concentration of the solute. For example, the osmotic pressure for a 1,000 mg/L NaCl solution is 11.4 pounds per square inch (psi), whereas the osmotic pressure for a 35,000 mg/L NaCl solution is 398 psi. A rule of thumb, which is based on NaCl, that works well for most natural waters is that the osmotic pressure increases by approximately 0.01 psi for each milligram per liter of solute. However, high-molecular-weight organics produce a much lower osmotic pressure. For example, sucrose causes approximately 0.001 psi increase for each milligram per liter.

In practice, the saline feedwater is pumped into a closed vessel where it is pressurized against the membrane. The driving force pressure must be higher than the osmotic pressure of the feedwater and the membrane resistance in order to move water through the membrane. As a portion of the water passes through the membrane, the remaining feedwater increases in salt content. At the same time, a portion of this feedwater is discharged without passing through the membrane. Without this controlled discharge, the pressurized feedwater would continue to increase in salt concentration, creating problems such as precipitation of supersaturated salts and increased osmotic pressure across the membranes. The amount of the feedwater discharged to waste in this brine stream varies from 20 to 70 percent of the feed flow, depending on the salt content of the feedwater.

2.1.1 Reverse Osmosis System

An RO system is made up of the following basic components: (i) pretreatment, (ii) high-pressure pump, (iii) membrane assembly, and (iv) post-treatment (Figure 2-2).

Pretreatment removes suspended solids and helps prevent salt precipitation or microorganism growth on the membranes. This step is important because membrane surfaces are prone to fouling by particulate matter, inorganic scales (i.e., carbonate and sulfate salts of alkaline earth metals; oxides and hydroxides of aluminum and iron), organic material (i.e., humic, tannic, etc.), and biological material (e.g., bacteria, fungi, algae). Usually, pretreatment consists of fine filtration and the addition of acid or other chemicals to inhibit precipitation. Filtration typically consists of a sand filter, an activated carbon filter, and a depth cartridge filter. The sand filter is used to remove larger impurities, the activated carbon filter absorbs low molecular weight organics and reduces the amount of chlorine or other halogens, and the depth cartridge filter traps the remaining particles (in the 1 to 100 micrometer range) in the complex openings of a filter material (constructed of cotton, cellulose, synthetic yarns or "blown" microfiber such as polypropylene). These filters are often disposable. As particles accumulate, the pressure drop across the filter increases. When the pressure difference between filter inlet and outlet has increased by 5 to 10 psi relative to the starting point, the filter is backwashed or replaced.

In the pretreatment step, a number of chemicals may be added to prevent membrane fouling. Scaling of RO membrane surfaces is caused by the precipitation of sparingly soluble salts from the concentrated brine (especially $CaCO_3$ and $BaSO_4$). Sulfuric acid (H_2SO_4) or hydrochloric acid (HCl) may be added to the feedwater to reduce the pH and prevent $CaCO_3$ precipitation. However, H_2SO_4, although safer and less expensive than HCl, will increase the content of sulfate ions in the feedwater and increase the potential for $CaSO_4$ precipitation. Polyphosphates or, more recently, polycarboxylates can be added to prevent scaling due to $CaSO_4$ precipitation.

Chlorination is a common pretreatment method for preventing the growth of bacteria and algae that may cause fouling in the system or degradation of cellulose acetate membranes. The amount of chlorine required is determined by the amount of organic matter in the feedwater and by the water temperature. Some RO membranes cannot tolerate chlorine, so de-chlorination techniques are required where those are used. Ozone or ultraviolet light may also be used to remove organisms, but if ozone is used, it must be removed with chemicals before it reaches the membranes because it may react with the membranes.

After pretreatment, a membrane assembly, consisting of a pressure vessel containing the membrane module, is used to pressurize the feedwater. A high-pressure pump is used to supply pressure exceeding the osmotic pressure to enable the permeate to pass through the membrane, leaving dissolved salts and other contaminants behind with the concentrate, which is drawn off as waste. RO removes virtually all organic compounds, 90 to 99 percent of inorganic ions, and 99.9+% of viruses, bacteria, and pyrogens (Osmonics, Inc., 1997).

The RO membrane must be able to withstand the drop of the entire pressure across it. This pressure ranges from 250 to 400 psi for brackish water and from 800 to 1,180 psi for seawater. The semi-permeable membranes vary in their ability to pass fresh water and reject the passage of salts. No membrane is perfect in its ability to reject salts, so a small amount of salts passes through the membrane and appears in the product water.

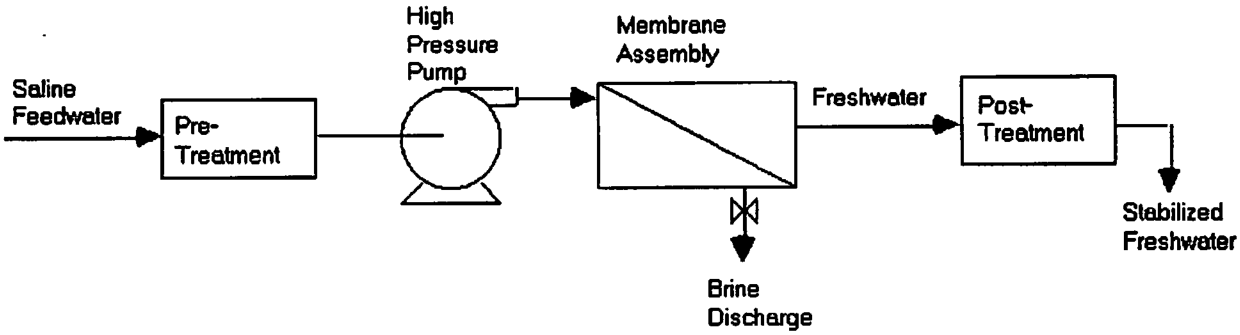


Figure 2-2. Flow diagram of a reverse osmosis system (Khan, 1986)

RO membranes come in a variety of configurations—tubular, hollow-fiber, flat-sheet, or spiral-wound. Two of the most commercially successful are spiral-wound and hollow-fiber. Both of these configurations are used to desalt brackish and sea water, although the construction of the membrane assembly will vary depending on the manufacturer and expected total dissolved solids of the feedwater. Spiral-wound membrane elements have gained the greatest acceptance in the market because they are the most rugged, leak-free, and pressure-resistant configuration. It consists of a continuous sheet of flat membrane rolled up like a large paper towel roll (Figure 2-3). The spiral design allows for optimum membrane surface area and fluid dynamics to produce a high permeate flow for the size of equipment required and are relatively easy to maintain with a routine cleaning program. A major advantage is enhanced "self-cleaning" due to the turbulent flow at the membrane surface. This mechanism dramatically reduces fouling, thereby enhancing performance and membrane life.

Hollow fine-fiber elements (Figure 2-4) consist of hollow fibers each roughly the size of a human hair. Thousands of fibers are closely bundled in each housing. As the pressurized feedwater flows slowly over the outside of the fibers, pure water permeates to the center and is collected out of the potted tube sheet. This type of membrane element provides more surface area and higher productivity than spiral-wound elements, but it is more prone to fouling and requires a significant amount of pre-filtration to keep the tightly packed membrane surfaces free of severe fouling (Amjad et al., 1998). The advantages and disadvantages of hollow-fiber and spiral-wound elements are compared in Table 2-1.

Spiral membrane elements are loaded in a serial configuration in a pressure vessel (one to seven membranes per pressure vessel), whereas hollow fiber membrane elements are loaded in a parallel configuration (one to hundreds of elements per pressure vessel). Multiple pressure vessels may be connected in a serial or parallel flow path to enhance the product water quality or quantity.

Over 100 different materials are used to make RO membranes. However, the two most commonly used materials are cellulose acetate and polyamide thin film composite. The characteristics and performance of these membranes are compared in Table 2-2.

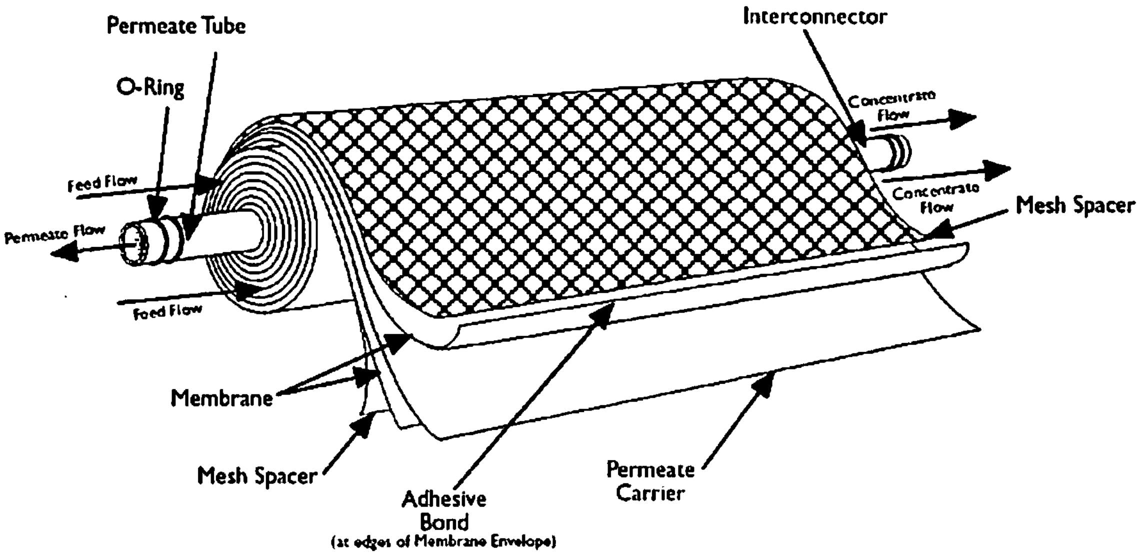


Figure 2-3. Spiral-wound membrane element (Osmonics, Inc., 1997)

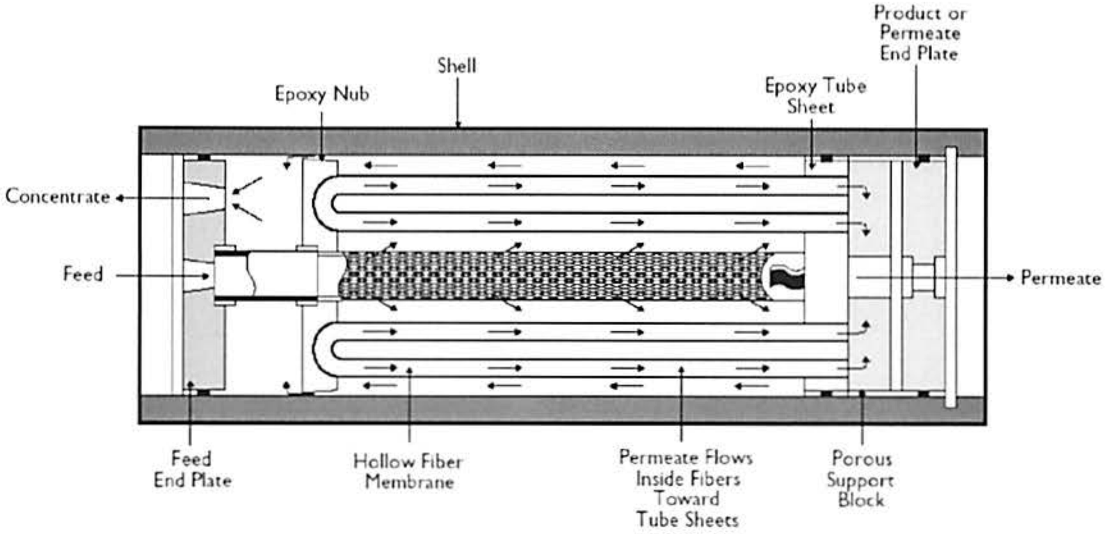


Figure 2-4. Hollow fine-fiber membrane element (Osmonics, Inc., 1997)

All RO membranes require regular maintenance. Membrane cleaning is recommended when feed pressures increase by 10% to maintain the same product flow, or when a 10% drop in product flow occurs at constant pressure. Reduction in RO performance results from foulant accumulation, e.g., colloidal silt, colloidal silica, and ferric iron, or from precipitation of $CaCO_3$, $CaSO_4$, $BaSO_4$, $SrSO_4$, ferrous iron, silicates, and other scale. It is important that all inorganic, organic, and biological foulants be removed from the membrane surface, otherwise irreparable damage may result such as extrusion to the membranes from excessive pressure or salt rejection loss due to membrane surface abrasion by the foulants or crystal growth into the pores of the membrane. The cleaning method and frequency depend on the type of foulant and the membrane's chemical resistance. Cleaning methods include mechanical cleaning (i.e., direct osmosis, flushing with high velocity water, ultrasonic, sponge ball or brush cleaning, air sparging, etc.), chemical cleaning (use of chemical agents), or a combination of both. The most common generic inorganic acids used for membrane cleaning are HCl (industrial grade known as muriatic acid) and H_2SO_4.

Table 2-1. Comparison of hollow-fiber and spiral-wound membranes (Amjad et al., 1998)

Membrane	Advantages	Disadvantages
Hollow-Fiber	1. High membrane surface area to volume ratio 2. High recovery in individual RO unit 3. Easy to troubleshoot 4. Easy to change bundles in the field	1. Sensitive to fouling by colloidal materials 2. Limited number of membrane materials and manufacturers
Spiral-Wound	1. Good resistance to fouling 2. Easy to clean 3. Variety of membrane materials and manufacturers	1. Moderate membrane surface area 2. Difficult to achieve high recovery

Table 2-2. Comparison of cellulose acetate and thin film composite membranes (Amjad et al., 1998)

Parameter	Cellulose Acetate	Thin Film Composite
Operating pressure (psi)	410 to 600	200 to 500
Operating temperature (°C)	0 to 30	0 to 45
Operating pH	4 to 6.5	2 to 11
Membrane degradation potential	Hydrolyzes at low and high pHs	Stable over broad pH range
Permeate flux (gfd*)	5 to 18	10 to 205
Salt Rejection (%)	70 to 95	97 to 99
Stability to free chlorine	Stable to low (<1 ppm) levels	Attacked by low levels (>0.1 ppm)
Resistance to biofouling	Relatively high resistance	Low resistance
Manufacturer	Several	Several
Cost	Lower	50 to 100 % more

*gallons per square foot per day

The product water from the membrane assembly usually requires post-treatment before being transferred to the distribution system for use as drinking water. This post-treatment might consist of the removal of gases, such as hydrogen sulfide (H_2S), and adjusting the pH from a value of approximately 5 to a value close to 7.

2.1.2 Effect of Operation Parameters on Reverse Osmosis Performance

The performance of membrane elements operating in a reverse osmosis system is affected by the feedwater temperature, applied pressure, water recovery ratio, and feedwater composition. The effects of pressure, temperature, and water recovery on RO membrane flux and water quality are illustrated in Figure 2-5.

Effect of Pressure

As shown in Figures 2-5a and 2-5b, water flux and product quality increase with an increase in applied pressure. However, above a certain value, higher pressures could result in decreased water flux due to scaling. At high pressure, foulants that normally would flow out of the system at lower pressure get impacted onto the membrane surface and hung up in the spacer. Concentration polarization, a normal phenomenon of salt build-up at the membrane surface, may increase at higher pressure. As the concentration increases, the local osmotic pressure increases. Under normal operating conditions, ions diffuse away from the membrane surface and the bulk flow carries them out of the system. But at the higher operating pressures, a higher surface osmotic pressure is required before diffusion away from the membrane can take place, thus increasing the potential for scaling (Wilbert et al., 1998). In addition, membrane compaction and deformation can be caused by excessive pressure.

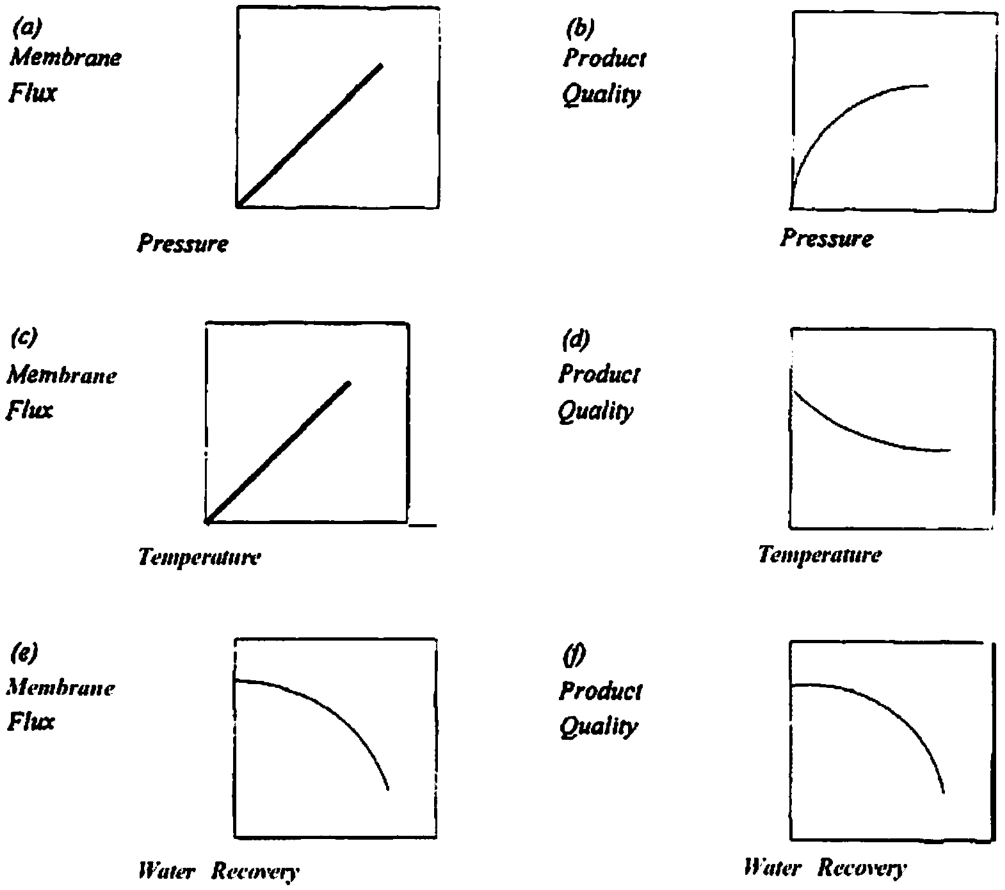


Figure 2-5. Effects of applied pressure, feedwater temperature, and water recovery on membrane flux and product water quality (figure taken from Wilbert et al., 1998). The graphs are generalized curves intended to show trends and are not based on actual data points.

Effect of Temperature

An increase in temperature increases the rate of diffusion through the membrane. Thus, the permeate flux is greater at higher temperatures than at lower temperatures (Figure 2-5c). A general rule of thumb is a 3% change in productivity per degree Celsius change in temperature. For example, if the feedwater temperature were 10 °C during winter, the membrane element flux would be approximately 55% of the 25 °C rated capacity. If the water temperature during summer were 30 °C, then the estimated flux would be 115% of the 25 °C capacity. Thus, the range in membrane flux can be large due to temperature fluctuations and needs to be accounted for in selecting the RO system parameters.

Since RO facilities are designed to operate at constant output, the feed pressure is adjusted to compensate for permeate flux changes resulting from changes in temperature. Figure 2-6 plots, as a function of temperature, the feed pressure necessary to keep the permeate flux constant across two membrane types; a high salt rejection membrane (CPA) and a high permeate flux membrane (ESPA). As shown in the figure, the required feed pressure decreases with increasing temperature.

To a certain extent, salt diffusion through the membrane is also increased as temperature is increased. As shown in Figure 2-6, permeate salinity increases with increasing feedwater temperature, thus product

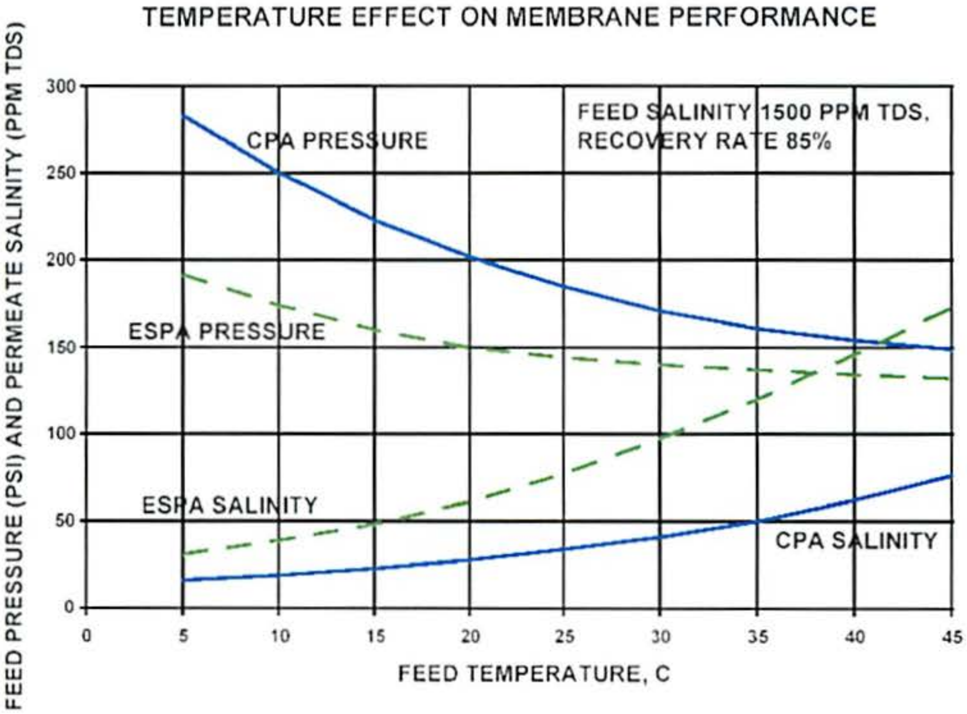


Figure 2-6. Effect of feedwater temperature on reverse osmosis system feed pressure and permeate salinity (figure taken from http://www.membranes.com/docs/trc/desparam.pdf). TDS = total dissolved solids

quality decreases with increasing temperature (Figure 2-5d). However, at normal operating temperatures, this effect is considered not important (Wilbert et al., 1998). In addition, higher temperature also increases the potential for carbonate scaling, membrane compaction and loss of porosity due to compression under pressure, and fouling caused by increased water flux.

Most cellulosic and thin film composite membranes have maximum temperature limits of 40 to 45 °C, which should be adequate for most surface and groundwater sources. Streams at higher temperatures should be used in a heat transfer process before treatment. At low temperatures, membranes become more rigid and water flux decreases, but most membranes can handle feedwater temperatures as low as 1 °C without a problem.

Effect of Water Recovery Ratio

Water recovery is defined as the ratio of permeate flow to feedwater flow. Recovery can be reduced by increasing the feedwater flow or by decreasing the operating pressure. If all other parameters remain the same, increasing the recovery ratio decreases the membrane flux and product quality (Figures 2-5e and 2-5f). Also, increasing the recovery ratio increases the concentration of all substances in the concentrate stream, thereby increasing the potential for membrane fouling and scaling. Because recovery rate strongly affects process economics, there is a tendency to design operation of RO systems at the highest practical value.

Effect of Feedwater Composition

Feedwater supplied to RO systems may exhibit variations in composition due to seasonal fluctuations or to mixing of water sources with different compositions. If the varying feedwater compositions do not require a change in RO system recovery ratio, changes in feedwater composition will affect only the required feed pressure and the permeate salinity. Figure 2-7 shows the change in required feed pressure and estimated permeate salinity as a function of feedwater salinity. The calculations were done for a system operating at an average flux rate of 15 gfd and recovery rate of 85% using two membrane types (ESPA and CPA2). The figure shows that feed pressure and permeate salinity both increase with an increase in feedwater salinity.

The pH of feedwater can affect the membrane structure and the potential for scale formation. The pH, as well as temperature, of the feedwater may need to be adjusted to control scaling of the concentrate flow stream. Calcium carbonate, for example, is more soluble at low pH and low temperatures, thus lowering the pH or temperature, or adding anti-scalants can minimize carbonate scaling. Silica, on the other hand, has higher solubility at pH greater than 7.7 and at higher temperatures, thus silica scaling can be mitigated either by raising the pH or the temperature of the feedwater.

Cellulosic membranes have a narrow operational pH range of 4 to 6. If exposed to pH outside this range, hydrolysis occurs, bonds in the membrane matrix are broken and replaced with hydroxyl ions, leaving holes in the matrix (Murphy, 1990). Thin film composite membranes generally have a much broader operational pH range, some as large as 2 to 11.

2.2 ELECTRODIALYSIS REVERSAL

Dissolved salts dissociate in water into positively and negatively charged ions. These ions are attracted to electrodes that have an opposite electric charge. When electrodes connected to an external source of direct current, such as a battery, are placed in a container of saline water, electrical current is carried through the solution and the ions migrate to the electrode with an opposite charge. A technology called electrodialysis (ED) takes advantage of this phenomenon by placing membranes, which allow either cations or anions (but not both) to pass, between a pair of electrodes (Figure 2-8). These membranes are arranged alternately, with an anion-selective membrane followed by a cation-selective

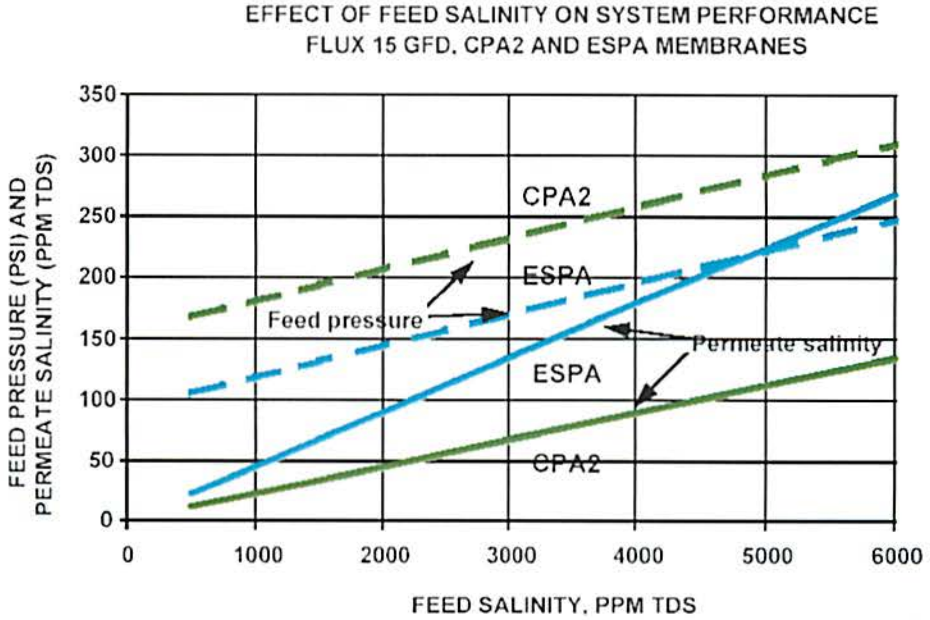


Figure 2-7. Effect of feedwater salinity on system performance (feed pressure and permeate salinity) (figure taken from http://www.membranes.com/docs/trc/desparam.pdf)

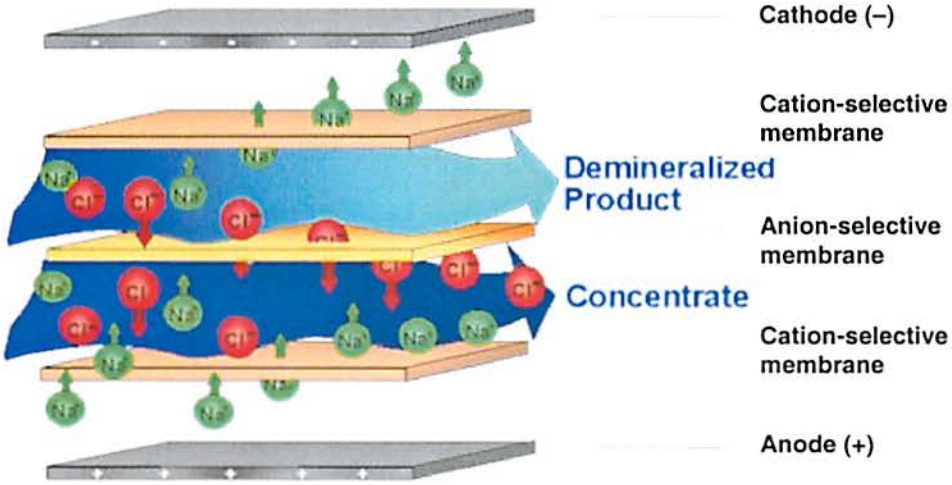


Figure 2-8. Simplified diagram of electrodialysis cell for sodium chloride removal (from http://www.ionics.com/toolbox/edr.htm)

membrane. A spacer sheet that permits water to flow along the face of the membrane is placed between each pair of membranes. One spacer provides a channel for feed (and product) water flow, while another provides a channel for flow of the concentrate.

As the electrodes are charged and saline feedwater flows along the product water spacer, the anions in the water are attracted and diverted towards the positive electrode (anode) and the cations are attracted towards the negative electrode (cathode). This reduces the salt content of the demineralized water in the product water channel. The anions pass through the anion-selective membrane, but cannot pass any farther than the cation-selective membrane, which blocks its path and traps the anion in the concentrate. Similarly, cations, under the influence of the negative electrode (cathode), move in the opposite direction through the cation-selective membrane to the concentrate channel on the other side. Here, the cations are trapped because the next membrane is anion-selective and prevents further movement towards the electrode.

By this arrangement, concentrated and dilute solutions are created in the spaces between the alternating membranes. These spaces, bounded by two membranes (one anionic and the other cationic) are called cells. The cell pair consists of two cells, one from which the ions migrated (the dilute cell for the product water) and the other in which the ions concentrate (the concentrate cell for the brine stream). The basic ED unit consists of several hundred cell pairs bound together with electrodes on the outside and is referred to as a membrane stack (Figure 2-9). Feedwater passes simultaneously in parallel paths through all of the cells to provide a continuous flow of desalinated water and brine to emerge from the stack.

One of the problems in ED is that fouling or scaling of the membrane and other active surfaces tend to occur over time due to deposition of organic and inorganic substances present in the feedwater. Electrodialysis reversal (EDR) was developed in the early 1970s to deal with this problem. In an EDR unit, the polarity of the electrodes is reversed and the flows are simultaneously switched at intervals of several times an hour, so that the brine channel becomes the product water channel and the product water channel becomes the brine channel. The result is that the ions are attracted in the opposite direction across the membrane. Immediately following the reversal of polarity and flow, enough of the product water is dumped until the stack and lines are flushed out, and the desired water quality is restored. This flush takes about 1 or 2 minutes, and then the unit can resume producing water. The reversal process is useful in

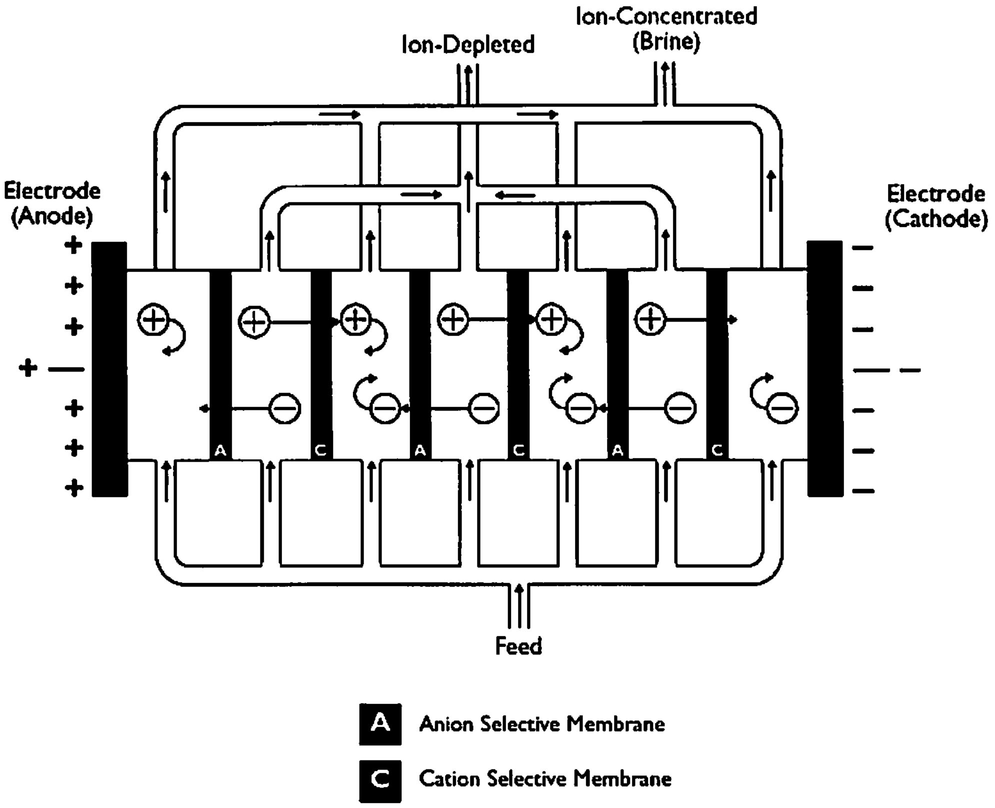


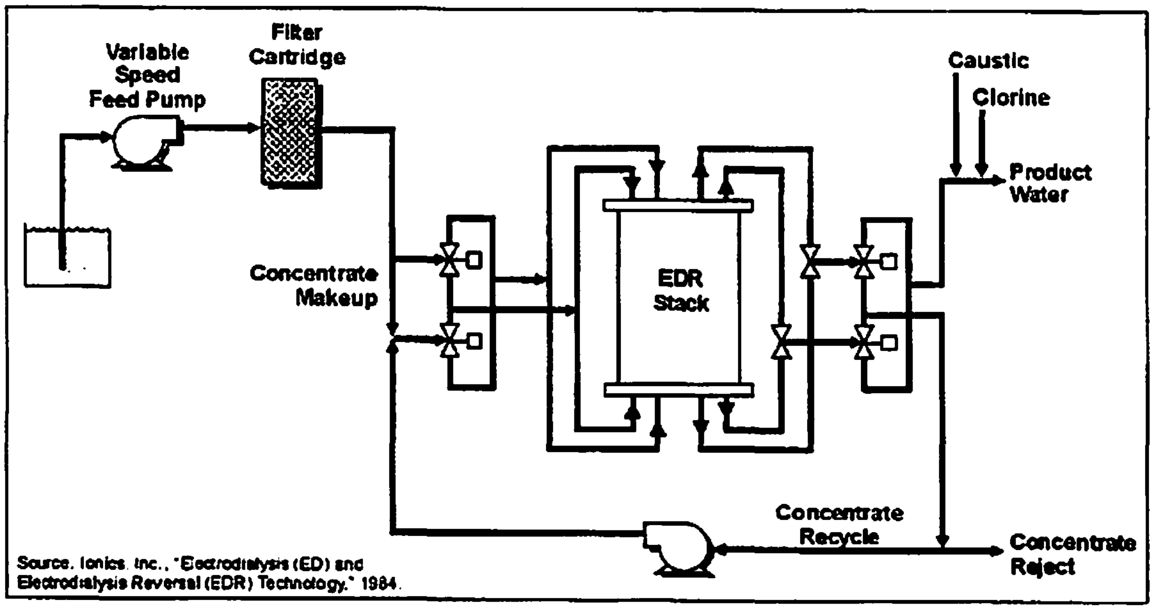
Figure 2-9. Electrodialysis reversal membrane stack (Osmonics, Inc., 1997)

breaking up and flushing out scale, slimes, and other deposits in the cells before they can build up and create a problem. Flushing allows the unit to operate with fewer pretreatment chemicals and minimizes membrane fouling.

2.2.1 Electrodialysis Reversal System

An EDR system (schematic shown in Figure 2-10) is made up of the following basic components: (i) pretreatment train, (ii) membrane stack, (iii) low-pressure circulating pump, (iv) power supply for direct current (a rectifier), and (v) post-treatment. The raw feedwater must be pre-treated to prevent impurities that could harm the membranes or clog the narrow channels in the cells, such as organic matter, colloidal substances, microorganisms, and insoluble salts, from entering the membrane stack. The pretreatment system is site-specific depending on the feedwater quality, although cartridge filtration is typically used. An EDR system requires periodic chemical cleaning to remove foulants that have accumulated on the membrane surface. Three methods used to remove scale and other contaminants are polarity reverse flow, clean-in-place, and stack disassembly. The latter involves disassembly of the membrane stack and scrubbing of the membranes to remove the contaminants. EDR systems periodically flush an acid solution across the electrodes to prevent scale from depositing on the electrode surface (HDR Engineering, Inc., 2000).

The feedwater is circulated through the stack with a low-pressure pump with enough power to overcome the resistance of the water as it passes through the narrow passages. A rectifier is generally used to transform alternating current to the direct current applied to the electrodes on the outside of the



Source. Ionics, Inc., "Electrodialysis (ED) and Electrodialysis Reversal (EDR) Technology." 1984.

Figure 2-10. Typical electrodialysis reversal flow schematic (figure taken from HDR Engineering, Inc., 2000)

membrane stacks. Post-treatment consists of stabilizing the water and preparing it for distribution. This post-treatment might consist of removing gases such as hydrogen sulfide and adjusting the pH.

As with RO, the EDR systems are modular in nature. Therefore, additional capacity can be easily and economically added to increase treatment capacity once the design criteria are established and if infrastructure for the treatment facilities are adequately planned and engineered in the initial installation.

2.2.2 Effect of Operation Parameters on Electrodialysis Reversal Performance

The feedwater characteristics, design parameters, and equipment selection control the rate of ion removal during EDR. The water quality and temperature of the feedwater determine the system recovery and rate of mass transfer. Ion removal increases as the temperature and ionic charge increase. The maximum temperature in EDR stacks used to be 40 °C, but recently developed membranes allow operation at temperatures up to 60 °C (http://ameridia.com/html/elep.html).

The key design parameters that affect system performance are current density, current efficiency, and diluate and concentrate concentrations. The current density is the driving force of the process because it determines the quantity of ions transported across the membranes. Running at a high current density reduces the required surface of EDR cells, which makes the process more attractive, but results in a disproportionate cell voltage increase and much higher power consumption. For a given current density, the cell voltage increases with time as the membranes are either chemically affected or physically fouled by contaminants. The cell voltage also increases as the active sites in the polymeric structure of the membranes disappear. These processes determine the time at which membranes need to be replaced.

Current efficiency also determines the surface area of membranes required for a given application. This design parameter takes into consideration all the parasitic phenomena occurring in the stack, such as non-perfect permselectivity of membranes or physical leakage (leading to impurities in the products) that can be reduced by optimized stack design and membrane selection.

The ratio of the concentrations (conductivities) of the input and output streams affects the current efficiency, limiting the maximum concentration for the concentrate stream. In most cases, 20 is the maximum concentration factor that can be obtained (provided that the solubility of potential precipitates is high enough), unless more than one stage is used. This concentration factor is generally much higher than with RO.

Organics and weakly charged inorganics are not removed by EDR. It also does not provide a barrier to pathogenic microorganisms, in contrast to RO systems. Also, the membranes cannot tolerate many organic solvents and most oxidizing chemicals.

The total dissolved solids concentration affects the relative economics of EDR more than any other factor. As the total dissolved solids concentration increases, more electrical power is required; conversely, as the total dissolved solids concentration decreases, less electrical power is required. EDR has been used mainly for saline water with less than 3,000 mg/L total dissolved solids (HDR Engineering, Inc., 2000).

2.3 CONCENTRATE DISPOSAL

The concentrate stream is typically 10 to 25 percent of the feedwater flow for brackish water and greater for seawater desalination. Therefore, a significant volume of concentrate from a desalination facility requires disposal. The quantity and chemistry of the concentrate will depend on the quality of the source water and the desalination process employed, but in any case the discharge of the concentrate will require an environmentally acceptable disposal method that meets the applicable regulatory requirements of appropriate Federal, state, and local agencies.

Various options are available for concentrate disposal. The most common methods are: (i) discharge to a surface water body (e.g., ocean outfall), (ii) discharge to a wastewater treatment plant, (iii) discharge with storm water, (iv) land application, (v) deep injection well, and (vi) evaporation by either thermal or solar application (Wilbert et al., 1998) (HDR Engineering, Inc., 2000). The different methods of concentrate disposal must be evaluated carefully because the selected method will be a major factor in the overall cost of the desalination project and can affect the ability of the facility to meet regulatory requirements. The ability to discharge to a surface water body, fresh or saltwater, is limited by regulatory constraints for the receiving body and the cost of the discharge system infrastructure. Discharge into a municipal wastewater system can be cost effective if the desalination facility is located in close proximity to a wastewater treatment plant. Where the desalination facility is not co-located with a wastewater treatment facility, the concentrate can be delivered through the municipal collection system, but analyses must be performed to determine the impact of the concentrate influent to the wastewater treatment process.

Deep well injection disposal is most commonly used in inland desalination facilities. In many cases, the lack of any surface water body within a reasonable distance makes this method the only disposal option. Deep well injection involves disposing the concentrate in a geologic zone that contains lower quality water and is separated from potential potable water aquifers by low permeability zones. This method is potentially useful for disposal of concentrates from an Edwards Aquifer region desalination plant. However, its feasibility is very site specific and requires detailed geological investigations for specific locations. Deep well injection also has a high up front cost.

Disposal of concentrate to a land surface evaporation pond is an option available under very specific conditions. The requirements for effective disposal through land application include: (i) sufficient land availability, (ii) high evaporation rates, (iii) low precipitation rates, (iv) low concentrate discharge volumes, and (v) adequate pond liner material (HDR Engineering, Inc., 2000). This method is used for low discharge volumes (<0.01 MGD) associated with industrial use facilities and is not viable for public water supply facilities that will require an excessive amount of land for effective evaporation.

Concentrate disposal is regulated under several Federal laws, and state and local governments may impose additional regulations. A published report (HDR Engineering, Inc., 2000) describes the potential state and federal regulatory issues that may be involved in the disposal of concentrate from a desalination facility in the State of Texas. The report emphasized the required permits, codified rules, and the regulatory considerations that may be involved in the disposal of concentrate by means of surface water discharge, discharge into municipal wastewater system, land application, and deep well injection.

3 SUMMARY INFORMATION ON EXISTING AND PLANNED MUNICIPAL DESALINATION FACILITIES

3.1 EXISTING FACILITIES

The following tables give a summary of information on 10 municipal water desalination plants in the United States, seven using reverse osmosis and three using electrodialysis reversal. The tabulated information includes the startup year, process used, plant capacity, recovery rate, pre- and post-treatment methods, feed and product water compositions, concentrate disposal method, and supplier of equipment and membrane. The information presented in the tables was taken from the website http://www2.hawaii.edu/~nabil/depow.htm).

Jupiter Reverse Osmosis Water Plant **Jupiter Water Department** **P.O. Box 8900** **Jupiter, FL 33458**	
Startup Year	1997
Process	Reverse Osmosis
Capacity (MGD)	29.7*
Recovery Rate (%)	75
Pretreatment	anti-scalant, microfiltration
Post-Treatment	disinfection, blending with raw well water (4.2 MGD) and with lime-softened water (13.5 MGD)
Feed Water Composition	TDS=2000 ppm
Product Water Composition	TDS=200 ppm
Concentrate Disposal	injection well
Equipment Supplier	Hydropro, Inc., Lake Park, FL
Membrane Supplier	Hydranautics, Wilmington, DE

*1997 expansion of a 6.0 MGD RO plant previously built in 1990

Cape Coral Reverse Osmosis Water Treatment Plant **3300 SW 20th Avenue** **Cape Coral, FL 33914**	
Startup Year	1992*
Process	Reverse Osmosis
Capacity (MGD)	16.8*
Recovery Rate (%)	80
Pretreatment	pH adjustment, cartridge filtration
Post-Treatment	degassing, disinfection
Feed Water Composition	TDS: 1500 ppm
Product Water Composition	TDS: 80 ppm
Concentrate Disposal	discharge to wastewater treatment plant
Equipment Supplier	American Engineering Services, Tampa, FL
Membrane Supplier	Fluid Systems, Venice, FL

*1992 was a 6.5 MGD expansion. In 1985, the rated capacity was 8.8 MGD. Blending amounts to 1.5 MGD. The original plant was a 3 MGD RO plant built in 1976.

Joe Mullins Reverse Osmosis Water Treatment Facility **Melbourne Water Treatment Facility** **6055 Lake Washington Road** **Melbourne, FL 32934**	
Startup Year	1995
Process	Reverse Osmosis
Capacity (MGD)	23.0*
Recovery Rate (%)	85
Pretreatment	pH adjustment, antiscalant, pre-filtration
Post-Treatment	disinfection, blending with raw well water and, for overall peak capacity, blending with the South Water Treatment Plant product water
Feed Water Composition	chloride content of 500 ppm
Product Water Composition	potable water
Concentrate Disposal	discharge to saltwater lagoon
Equipment Supplier	Professional Water Technologies, Escondido, CA

*The permitted peak capacity of the South WTP is 16.5 MGD. The capacity of the RO plant is 6.5 MGD when 1.5 MGD of raw well water is blended with RO product water. As a result, the total drinking water capacity now available is 23.0 MGD. Currently, the average daily demand is being met through the production and blending of 9.0 MGD of water from the South WTP and 5.0 MGD of water from the RO WTP.

Lee Hall Reverse Osmosis Plant **36 Reservoir Road** **Newport News, VA 23608**	
Startup Year	1998
Process	Reverse Osmosis
Capacity (MGD)	5.7
Recovery Rate (%)	81
Pretreatment	sulfuric acid, anti-scalant, cartidge filtration
Post-Treatment	degassing, pH adjustment, blending with finished water from a conventional treatment plant
Feed Water Composition	TDS: 3600 mg/L, elevated trihalomethanes
Product Water Composition	TDS: 120 mg/L
Concentrate Disposal	aerated concentrate to outfall at James River
Equipment Supplier	WaterLink Technologies, Inc., West Palm Beach, FL
Membrane Supplier	Hydranautics, Wilmington, DE

Mount Pleasant Waterworks
RO Plant 1, King St.; RO Plant 2, 7th Ave.; RO Plant 3, Labor Camp
Mount Pleasant, SC 29464-3927

Startup Year	1991, 1992, 1994
Process	Reverse Osmosis
Capacity (MGD)	6.8
Recovery Rate (%)	80
Pretreatment	anti-scalant, pH adjustment, corrosion control
Post-Treatment	blending with deep well water from Middendorf Aquifer
Feed Water Composition	salinity: 190 mg/L, TDS: 900 mg/L
Product Water Composition	salinity: 50 mg/L; TDS: 60 mg/L
Concentrate Disposal	discharge to wastewater treatment plant
Membrane Supplier	Hydranautics, Wilmington, DE

Arlington Basin Groundwater Desalter
11615 Sterling Ave.
Riverside, CA 92503

Startup Year	1990
Process	Reverse Osmosis
Capacity (MGD)	5.4
Recovery Rate (%)	82
Pretreatment	sulfuric acid
Post-Treatment	Air stripping, decarbonation
Feed Water Composition	TDS: 1100 mg/L; nitrate
Product Water Composition	TDS: 50 mg/L; groundwater recharge
Concentrate Disposal	ocean discharge via Santa Ana Regional Interceptor (SARI) line (30-mile)
Equipment and Membrane Supplier	Hydranautics, Oceanside, CA

Marco Island Reverse Osmosis Water Treatment Plant
S. Heathwood Drive
Tract G, Unit 25
Marco Island, FL 34145-5029

Startup Year	1997[*]
Process	Reverse Osmosis
Capacity (MGD)	5.0
Recovery Rate (%)	75
Pretreatment	lime softening
Post-Treatment	Chlorination
Feed Water Composition	TDS: 7,500-10,500 ppm
Product Water Composition	TDS: 186 ppm
Concentrate Disposal	discharge to ocean outfall
Equipment Supplier	Harn RO Systems Inc., Venice, FL
Membrane Supplier	Fluid Systems Corporation, San Diego, CA

[*] 1 MGD expansion in 1997 to 5 MGD

Sherman Municipal Water Treatment Plant **243 Lacima Road** **Sherman, TX 75091**	
Startup Year	1993
Process	Electrodialysis Reversal
Capacity (MGD)	6.0
Recovery Rate (%)	85
Pretreatment	flocculation, sedimentation, filtration
Post-Treatment	blending with raw water, disinfection
Feed Water Composition	TDS: 1350 mg/L
Product Water Composition	TDS: 650 mg/L
Concentrate Disposal	surface water discharge
Equipment and Membrane Supplier	Ionics, Watertown, MA

Carlton Water Treatment Plant **1255 Mabry Carlton Parkway** **Venice, FL 34292**	
Startup Year	1995
Process	Electrodialysis Reversal
Capacity (MGD)	12.0
Recovery Rate (%)	85
Pretreatment	None
Post-Treatment	aeration, chlorination
Feed Water Composition	TDS: 1300 ppm
Product Water Composition	TDS: 250 ppm
Concentrate Disposal	groundwater injection
Equipment and Membrane Supplier	Ionics, Inc., Watertown, MA

Washington Water Treatment Plant **522 N. 4th Ave** **Washington, IA 52353**	
Startup Year	1993
Process	Groundwater Electrodialysis Reversal
Capacity (MGD)	1.8
Recovery Rate (%)	88
Pretreatment	None
Post-Treatment	aeration, chlorination
Feed Water Composition	Brackish water TDS: 1200 mg/L
Product Water Composition	TDS: 600 mg/L
Concentrate Disposal	surface water discharge
Equipment and Membrane Supplier	Ionics, Inc., Watertown, MA

3.2 PLANNED FACILITIES

In addition to currently operating municipal desalination plants such as those listed in the previous section, several projects are in various stages of development, from preliminary design to construction, to treat both brackish water and seawater. Construction is underway on two large-scale desalination plants in Florida and California, and additional plants have been proposed in those states and in Texas (American Water Works Association, 2002). For example, El Paso Water Utilities and Fort Bliss officials have agreed to work together in building the country's largest inland desalination plant. The plant will draw brackish water from the Hueco Bolson, an underground aquifer that provides about 40 percent of El Paso's municipal water supply. Water is currently extracted from the aquifer 25 times faster than it can be replenished, a rate that would deplete the Texas portion of fresh water within 30 years. A U.S. Geological Survey study indicated that the aquifer contains a vast quantity of brackish water that can be treated and used. The two parties will collaborate on building a 27.5 MGD (84.4 acre-ft/day) desalination plant capable of meeting the needs of the city and the base. Construction of the plant, to be located on Fort Bliss property in East El Paso, is scheduled to begin in 2003 and completion is scheduled for 2004.

In Florida, a $110 million state-of-the-art reverse osmosis plant is under construction that is projected to supply 25 MGD (77 acre-ft/day) of potable water from treated seawater (American Water Works Association, 2002). It is the first seawater desalination plant in Florida, where several small reverse osmosis plants already convert brackish water into fresh water. The RO system is supplied by Hydraunatics of Oceanside, California, and incorporates features and polyamide membranes that will reduce capital costs and power consumption to desalt water. The desalination plant is located next to the Tampa Power Company's Big Bend Power Station. The co-location is advantageous because the desalination plant can take some of the very large volume of seawater used as cooling water by the power plant—about 14,000 MGD (43,000 acre-ft/day)—as RO feedwater. Also, co-location allows the RO concentrate to be discharged back into the power plant's cooling water discharge stream, at a 1:70 ratio, to minimize the salinity load before being discharged into a canal to Hillsborough Bay. The new technology will increase water costs by four times—$1.71/1,000 gal ($557/acre-ft) the first year and $2.08/1,000 gal ($678/acre-ft) the next 29 years—but will be much less expensive than desalination using older technologies. Tampa's customers will pay an additional $7.50 per family each month (American Water Works Association, 2002). The plant is scheduled to be on-line by January 2003.

On the Texas Gulf Coast near Freeport, Dow Chemical is collaborating with Poseidon Resources to build a 25 MGD (77 acre-ft/day) seawater desalination plant that would be upgradeable to 100 MGD (307 acre-ft/day), making it potentially the largest seawater desalination plant in the country (Krishna, 2002). The City of Corpus Christi is considering a 5 MGD (15 acre-ft/day) plant on Mustang Island using brackish groundwater, and may later build a 25 to 30 MGD (77 to 92 acre-ft/day) plant in conjunction with the Barney Davis Power Station. In North Texas, the City of Wichita Falls plans to complete by 2003 a 15 MGD (46 acre-ft/day) microfiltration and RO plant using water from Lake Kemp. The Southmost Regional Water Authority in Cameron County is planning to build desalination plants using brackish groundwater, and later from seawater (Krishna, 2002).

4 ESTIMATED COST OF MEMBRANE DESALINATION SYSTEMS

Many factors influence the cost of producing potable water by membrane desalination. The essential cost-determining factors are (Ludwig, 2002): (i) basic process design parameters such as salinity, composition, and temperature of the feedwater and their annual fluctuations; (ii) capital cost for process equipment, civil works, and infrastructure facilities; (iii) energy import and export cost (heat and/or power); (iv) operation cost of personnel, consumables, and maintenance; (v) service life of membranes and membranes replacement cost; and (vi) site-related cost factors. An evaluation of the economic feasibility of any desalination facility proposed for the Edwards Aquifer region will need to consider these various factors.

4.1 ESTIMATED COSTS CALCULATED USING THE WATER TREATMENT ESTIMATION ROUTINE

Preliminary cost estimates for RO and EDR plants treating Edwards Aquifer saline water were calculated using the Water Treatment Estimation Routine (WaTER), which is an Excel spreadsheet developed by the Bureau of Reclamation for estimating the cost of desalination treatment systems (Wilbert et al., 1999). WaTER is based primarily on the U.S. Environmental Protection Agency (EPA) report "*Estimating Water Treatment Costs, Vol. 2, Cost Curves Applicable to 200 MGD Treatment Plants*" (Gumerman et al., 1979). The basic assumptions used in the estimation methods and the generic configuration of each desalination process are provided in the EPA report. The cost estimates derived using WaTER are based on theoretical equipment sizes and chemical requirements and are not intended to be final design cost estimates. Nevertheless, those values are useful as tools for comparing different process options during the early stages of the planning process.

WaTER requires information on the required plant feedwater flow rate, desired plant product flow rate, water analysis data, and cost indices. Construction cost, operation and maintenance costs, and sizing calculations for the different processes are performed in linked worksheets, and process design parameters can be adjusted to refine the cost estimates. For capital cost estimation, direct capital costs in WaTER are the sum of the costs for membranes, RO skids, building, electrical, instrumentation and control, high pressure pumps, raw water transfer pumps, product water pumps, degasifiers, odor control, process piping, yard piping, chemical feed with pumps, cartridge filters, membrane cleaning equipment, contractor engineering and training, concentrate treatment and piping, generators, and site work (Wilbert et al., 1999). The equations used for most of the direct capital costs are from the technical paper by Suratt et al. (1995). Indirect capital costs in WaTER are the sum of interest during construction, contingencies, architectural and engineering (A&E) fees and project management, and working capital. Operation and maintenance costs include electricity, labor, chemicals (e.g., acid, caustic, anti-scalant, and chlorine), membrane replacement, cleaning chemicals, cartridge filters, repairs and replacement, insurance, and laboratory fees.

In this study, cost estimates were calculated for a wide range of plant capacity, feedwater total dissolved solids concentration, and product quality. The following input parameters in WaTER were varied as indicated:

- Plant output capacity of 10, 50, and 100 acre-feet per day (3.3, 16.3, 32.6 MGD)
- Feedwater total dissolved solids of 1,000, 3,000, 5,000, and 10,000 mg/L (total suspended solid concentration assumed negligible)
- Target post-treatment total dissolved solids of 500 and 1,000 mg/L

Additional calculations were also done for an RO plant capacity of 30.7 acre-feet per day (10 MGD) to provide a better constraint on the RO cost curve for plant capacities between 10 and 50 acre-feet per day. Cost estimates for EDR plants with product flow rates less than 10 acre-feet per day were also derived to permit comparison with values given in published literature.

In the absence of detailed information on potential Edwards Aquifer region desalination facilities, a number of parameters required by WaTER were set equal to the default values in the spreadsheet. These parameters include facility design (high pressure pumps, transfer pumps, product water pump, number of RO modules per vessel, number of vessels per skid, etc.), chemical costs, membrane life, cleaning rate, and operational labor. The values for plant availability due to downtime, planned operation time per day, and percent recovery were fixed at 85%, 100%, and 75%, respectively. Membrane data applicable for a brackish water membrane (Film Tec BW30-400) were taken from the WaTER user manual (Wilbert et al., 1999). The cost components were adjusted to December 2002 using the cost indices published in Engineering News Record (http://enr.construction.com).

The concentration of total dissolved solids and the chemical composition of saline Edwards Aquifer waters vary over a wide range. The measured or calculated total dissolved solids concentration and the water types of several Edwards Aquifer saline water samples reported by W.F. Guyton Associates (1986), Groschen (1994), and Oetting et al. (1996) are tabulated in Table 4-1, and the chemical compositions are plotted in Figure 4-1. For the purpose of cost estimation, the composition of the sample from Monitor Well A-3 reported by W.F. Guyton Associates (1986), tabulated in Table 4-2 and plotted as a red circle in Figure 4-1, was selected as the reference composition. The Monitor Well A-3 sample is a Ca–Na–Mg–SO_4–Cl brine with a calculated total dissolved solids concentration of 4,397 mg/L (measured value is 4,200 mg/L), which is on the high end of the range of values listed in Table 4-1. For the cost calculations, the dissolved species concentrations of the Monitor Well A-3 sample were increased or decreased proportionately to the ratio of the feedwater total dissolved solids concentration and 4,397 mg/L, and were input into the "H_2O Analysis" worksheet in WaTER. For example, for feedwater with a total dissolved solids concentration of 5,000 mg/L, the Monitor Well A-3 concentrations listed in column-2 of Table 4-2 were increased to the values listed in column-3 of the table.

The estimated annual costs ($/acre-foot) of producing potable water from Edwards Aquifer saline water using RO and EDR are plotted in Figures 4-2 and 4-3, respectively, as a function of plant capacity (acre-feet/day) and total dissolved solids concentration (mg/L) of feed and product water. The estimated costs are also plotted in Figures 4-4 and 4-5 in units of dollars per 1,000 gallons ($/1,000 gal) to facilitate comparison with literature values given in those units. EDR costs for a narrower range of plant capacity (2 to 30 acre-ft/day or 1 to 10 MGD) are illustrated in Figures 4-3(b) and 4-5(b). For reference, examples of input and output WaTER worksheets are included in the appendix.

The desalination costs estimated using WaTER show that the unit cost of producing potable water from both RO and EDR systems depends strongly on the total dissolved solids concentration of the feedwater and on plant capacity. For RO systems, economies of scale are observed. The WaTER results suggest that RO water unit costs decrease significantly with increasing capacity up to about 30 acre-ft/day (10 MGD). A further increase in plant capacity results in a smaller decrease in unit cost. EDR systems also show economies of scale, but in a narrower range of plant capacity. EDR water unit costs decrease significantly with increasing plant capacity up to about 3 acre-ft/day (1 MGD), remain relatively constant to about 20 acre-ft/day (6 MGD), and increase significantly at even higher plant capacity, especially for feedwater with total dissolved solids greater than 3,000 mg/L. The increased cost at higher total dissolved solids concentration and higher plant capacity is most likely due to higher electric power consumption. EDR is cost competitive with RO for small to moderate-sized facilities [~10 acre-ft/day (~3.3 MGD)] if the feedwater total dissolved solids concentration does not exceed 3,000 mg/L.

Table 4-1. Total dissolved solids (TDS) concentration and water types of Edwards Aquifer saline water samples. Chemical analysis data taken from W.F. Guyton Associates (1986), Groschen (1994), and Oetting et al. (1996).

Sample	Water type	Reference	TDS (mg/L)
Facies A-1	Ca-SO_4	Oetting et al. (1996)	2,635*
Facies A-2	Ca-Mg-SO_4	Oetting et al. (1996)	3,183*
Facies A-3	Ca-SO_4	Oetting et al. (1996)	2,568*
Facies B-1	Ca-Na-Cl-HCO_3	Oetting et al. (1996)	938*
Facies B-2	Ca-SO_4-HCO_3-Cl	Oetting et al. (1996)	852*
Facies B-3	Na-Ca-Cl-SO_4-HCO_3	Oetting et al. (1996)	1,106*
Facies B-4	Ca- SO_4	Oetting et al. (1996)	3,019*
Facies C-1	Ca-Na-Mg-SO_4-Cl	Oetting et al. (1996)	3,133*
Facies C-2	Ca-Na-Mg-SO_4-Cl	Oetting et al. (1996)	4,190*
Facies C-3	Ca-Na-Mg-SO_4-Cl	Oetting et al. (1996)	4,714*
Facies D-1	Na-Ca-Mg-Cl- SO_4	Oetting et al. (1996)	1,997*
Facies D-2	Na-Mg-Ca-Cl-SO_4-HCO_3	Oetting et al. (1996)	999*
Facies D-3	Na-Cl-SO_4	Oetting et al. (1996)	8,394*
Facies E-1	Na-Ca-SO_4-Cl	Oetting et al. (1996)	1,611*
Facies E'-1	Na-HCO_3-Cl-SO_4	Oetting et al. (1996)	1,161*
Facies E-2	Na-HCO_3-SO_4-Cl	Oetting et al. (1996)	474*
Facies E'-2	Na-Cl-SO_4-HCO_3	Oetting et al. (1996)	1,230*
Facies E-3	Na-SO_4-Cl-HCO_3	Oetting et al. (1996)	1,450*
Facies E'-3	Na-HCO_3-Cl-SO_4	Oetting et al. (1996)	507*
Monitor Well A-1	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	4,200
Monitor Well A-2	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,300
Monitor Well A-3	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	4,200
Test hole A, 1180 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,390
Test hole A, 1228 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,510
Test hole A, 1279 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,500
Test hole A, 1331 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,300
Test hole A, 1384 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,580
Test hole A, 1437 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,306
Test hole A, 1489 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,600
Test hose A, 1021ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	2,190
Test hose A, 1071ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	2,200
Test hose A, 1123 ft	Ca-Na-Mg-SO_4-Cl	W.F. Guyton Associates (1986)	3,090
Trinity Well T-1	Na-Cl-HCO_3-SO_4	Oetting et al. (1996)	1,112*
Trinity Well T-2	Na-Cl-HCO_3-SO_4	Oetting et al. (1996)	1,093*
Trinity Well T-3	Na-SO_4-Cl	Oetting et al. (1996)	2,420*
Well A1	Ca-Na-Mg-SO_4-Cl	Groschen (1994)	4,200
Well A2	Ca-Na-Mg-SO_4-Cl	Groschen (1994)	3,400
Well A3	Ca-Na-Mg-SO_4-Cl	Groschen (1994)	4,200

*Calculated using Aquachem Version 3.7.42 software

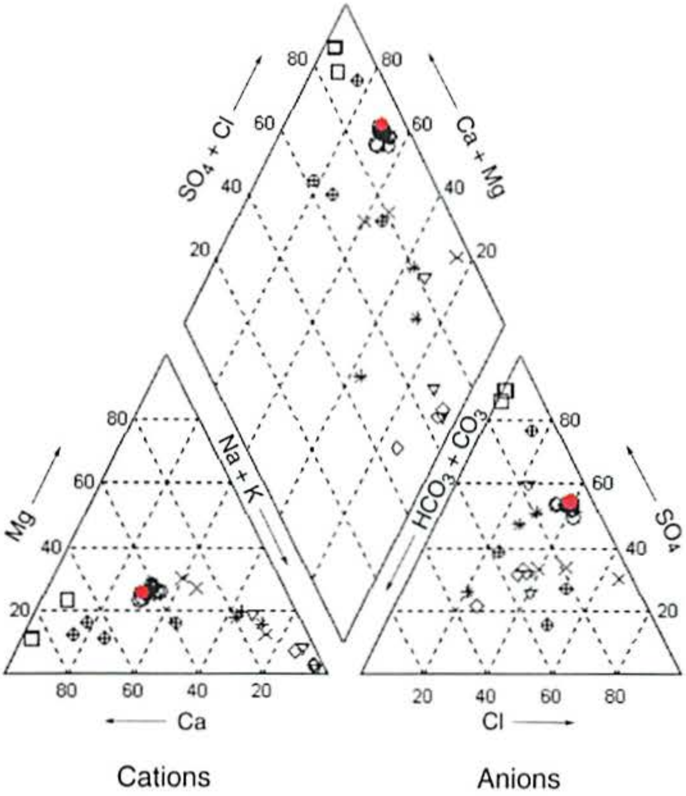


Figure 4-1. Piper diagram illustrating the wide range in chemical composition of Edwards Aquifer saline water samples reported by W.F. Guyton Associates (1986), Groschen (1994), and Oetting et al. (1996). The red circle is the composition of the sample from Monitor Well A-3, which was selected as the reference composition for the cost estimation using WaTER.

Table 4-2. Composition (column-2) of sample from Monitor Well A-3 (W.F. Guyton Associates, 1986) used as reference composition. Adjusted composition (column-3) used as input values in WaTER calculations for feedwater with 5,000 mg/L total dissolved solids.

Species	Well A-3 Sample Concentration (mg/L)	Concentration (mg/L) Adjusted to 5,000 mg/L Total Dissolved Solids
Na^+	510	580
K^+	29	33
Ca^{2+}	580	660
Mg^{2+}	220	250
Cl^-	940	1,069
SO_4^{2-}	1,800	2,047
HCO_3^-	294	334
F^-	3.1	3.5
SiO_2	21	24
Measured pH	6.7	
Sample temp. (°C)	31	
Calculated TDS (mg/L)	4,397	5,000

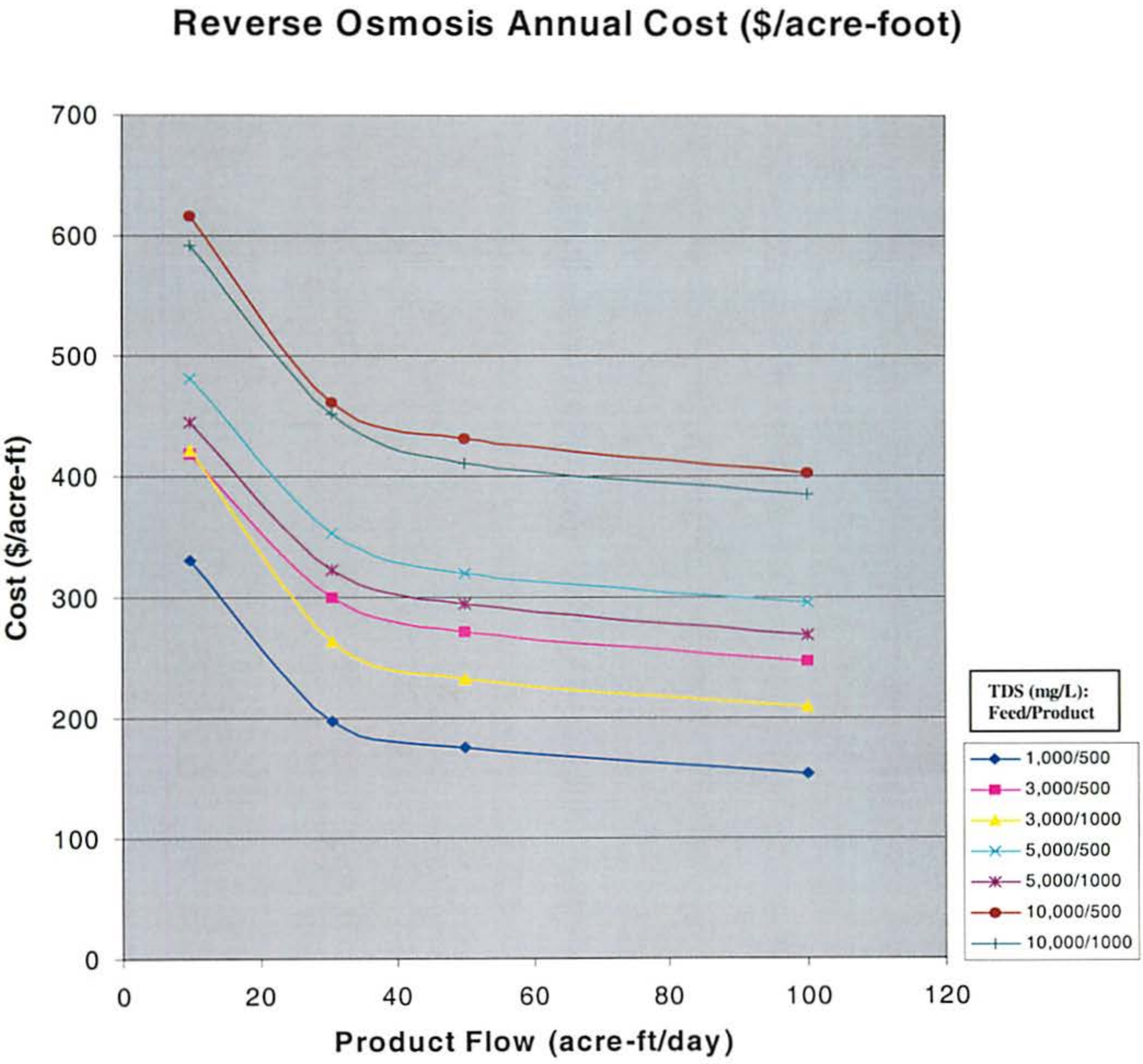


Figure 4-2. Estimated annual cost ($/acre-foot) of producing potable water by reverse osmosis as a function of plant capacity (acre-feet/day) and total dissolved solids concentration (mg/L) of feed and product water

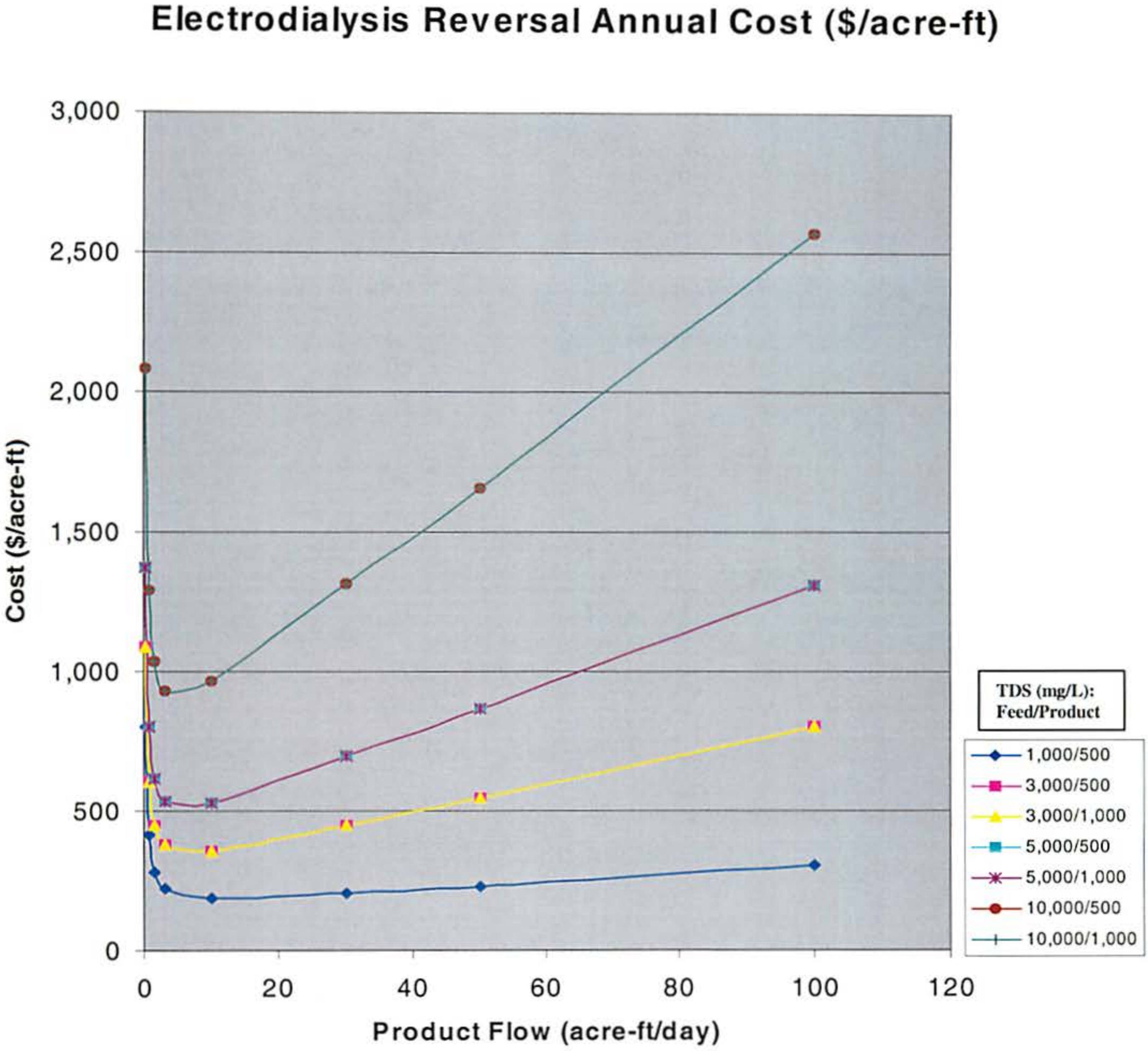


Figure 4-3(a). Estimated annual cost ($/acre-foot) of producing potable water by electrodialysis reversal as a function of plant capacity (acre-feet/day) and total dissolved solids concentration (mg/L) of feed and product water

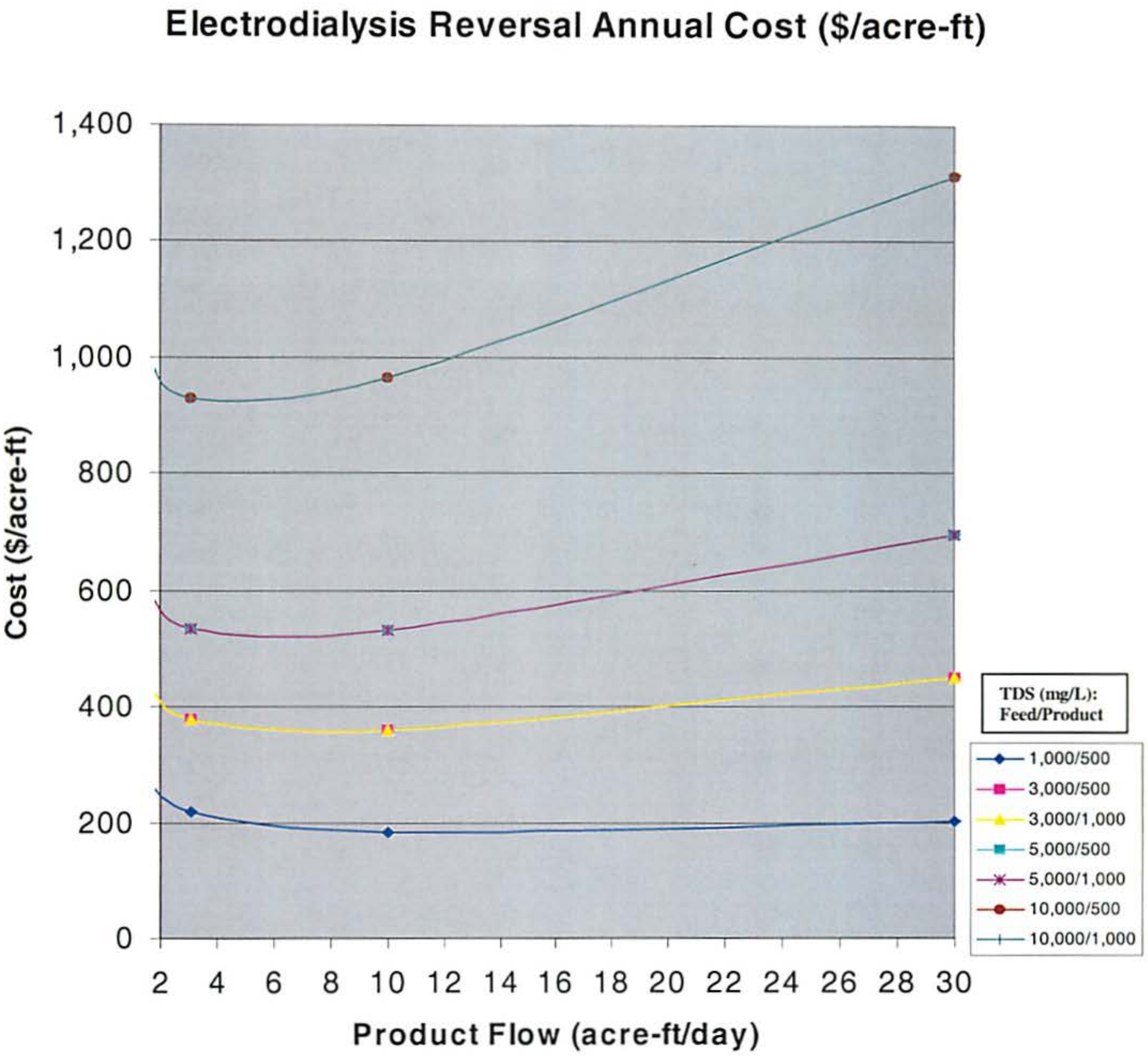


Figure 4-3(b). Estimated annual cost ($/acre-foot) of producing potable water by electrodialysis reversal for plant capacity ranging from 2 to 30 acre-feet/day as a function of total dissolved solids concentration (mg/L) of feed and product water

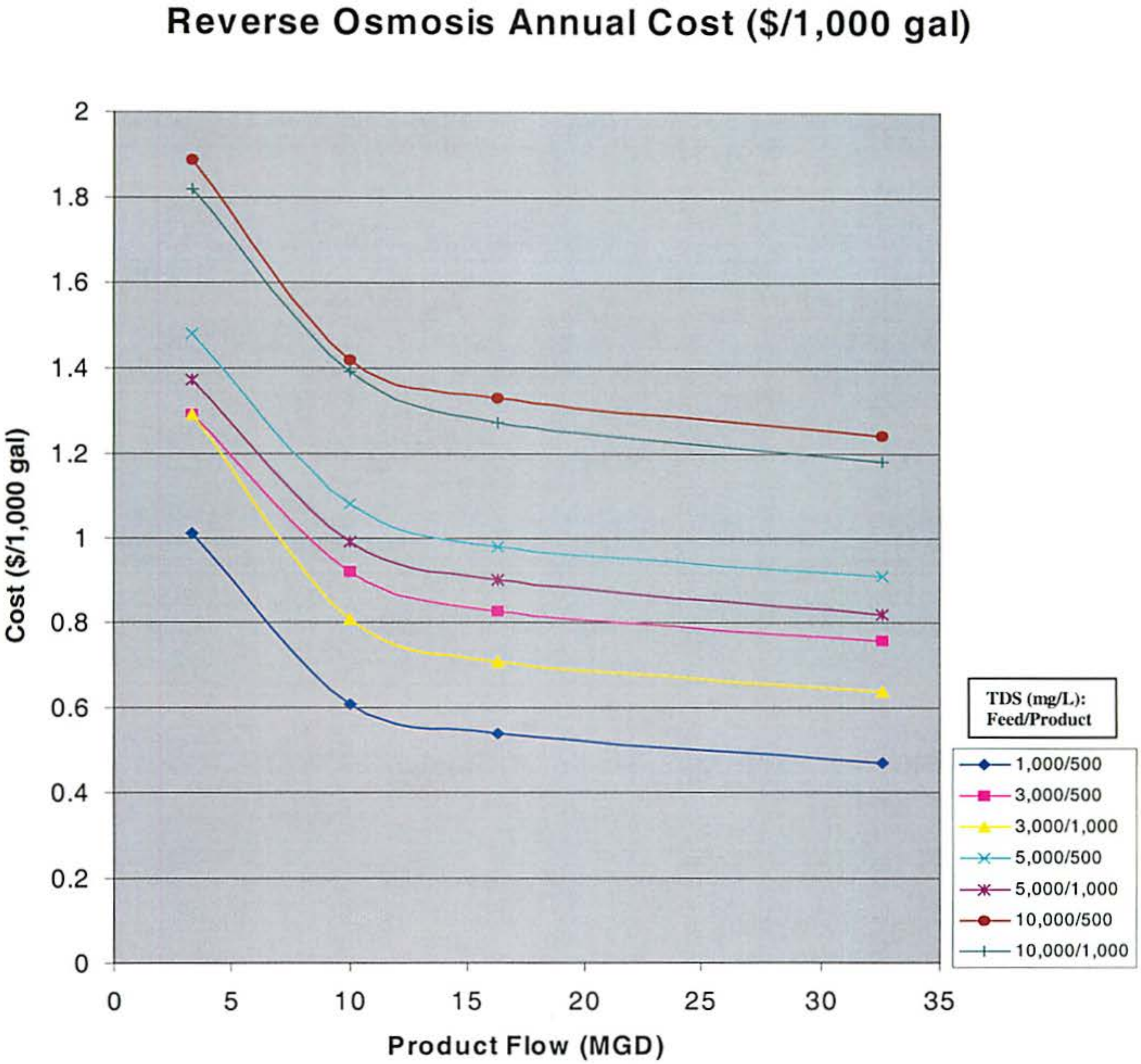


Figure 4-4. Estimated annual cost ($/1,000 gal) of producing potable water by reverse osmosis as a function of plant capacity (million gallons per day, MGD) and total dissolved solids concentration (mg/L) of feed and product water

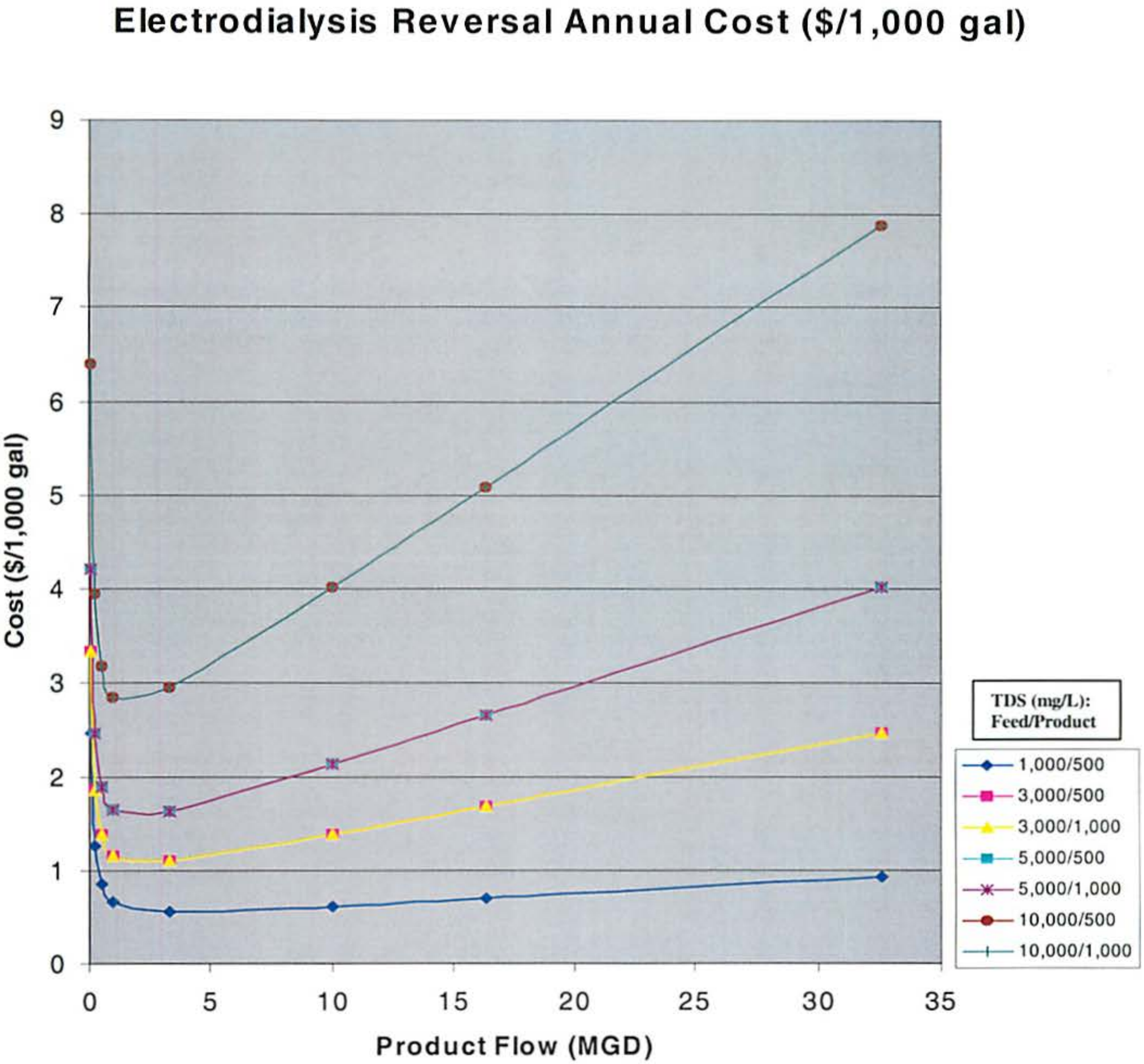


Figure 4-5(a). Estimated annual cost ($/1,000 gal) of producing potable water by electrodialysis reversal as a function of plant capacity (million gallons per day, MGD) and total dissolved solids concentration (mg/L) of feed and product water

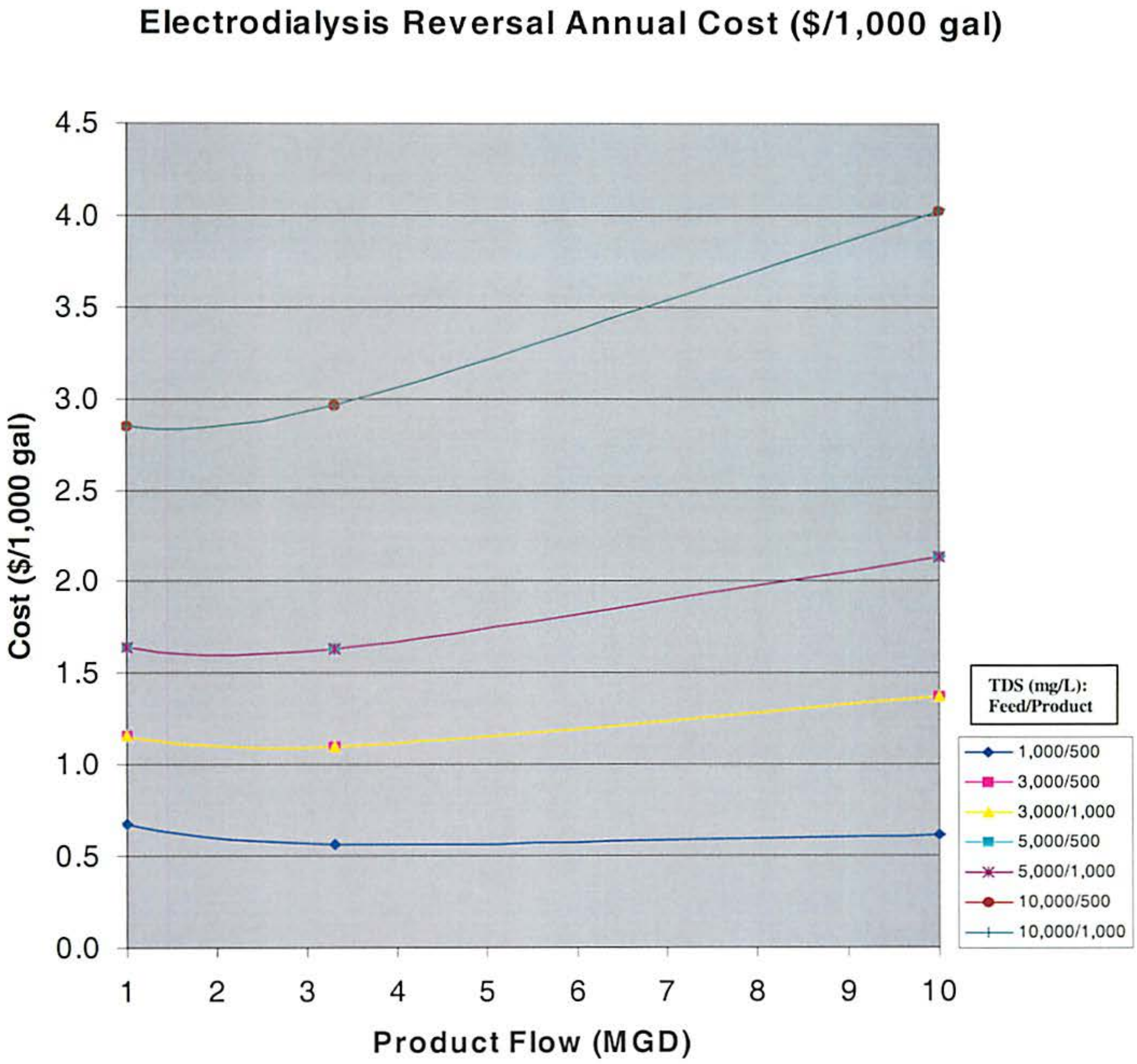


Figure 4-5(b). Estimated annual cost ($/1,000 gal) of producing potable water by electrodialysis reversal for plant capacity ranging from 1 to 10 MGD as a function of total dissolved solids concentration (mg/L) of feed and product water

The WaTER calculations indicate that the unit cost for RO water is also sensitive to the target quality of product water but the unit cost for EDR water is not, which is a surprising result. The insensitivity of EDR cost to product water quality arises because the formulas used in the WaTER "IonicsEDR" worksheet do not account for differences in product water total dissolved solids concentration. Unfortunately, the formulas used in the worksheet are not well documented. It would be useful to critically evaluate the equations and parameter values used in the WaTER calculations, which is beyond the scope of the present study.

4.2 REVERSE OSMOSIS AND ELECTRODIALYSIS REVERSAL COST DATA FROM OPERATING PLANTS

Cost estimation methods, such as WaTER or those used by other investigators (e.g., HDR Engineering, Inc., 2000), do not consider all factors that may influence the cost of the desalination project, such as concentrate disposal cost, permit costs, site capacity and land cost, and general local and nationwide economic conditions (Qasim et al., 1992). Actual project costs are highly site-specific and must be developed for individual circumstances. For example, the cost to effectively dispose of desalination concentrate will vary greatly according to a host of factors (HDR Engineering, Inc., 2000), such as (i) distance from plant facility to discharge point, (ii) quantity of concentrate discharge, (iii) quality of concentrate discharge, (iv) method of disposal, (v) permitting requirements, and (vi) monitoring requirements. Therefore, it is useful to compare estimated costs with information on construction and operation and maintenance costs derived from actual plant experience.

A survey of operating desalination plants in the United States recently was undertaken by HDR Engineering, Inc. to gather cost and performance data and identify trends in the costs associated with construction, operation and maintenance, and concentrate disposal (HDR Engineering, Inc., 2000). A questionnaire was sent to 117 public water systems believed to operate some form of desalination water treatment using membranes, mostly in Texas but also some in Florida and California. Survey responses were received from 17 plants, 13 of which use RO and 4 use EDR. Based on the cost data derived from the survey, cost curves representing capital, operation and maintenance, and total treatment costs were developed by statistical regression. The cost data derived from the survey and the cost curves derived by regression to the data are plotted in Figures 4-6 to 4-8. Capital cost, operation and maintenance cost, and total cost are plotted in Figure 4-6, 4-7, and 4-8, respectively, as functions of plant capacity. Capital costs include costs of initial construction and later expansions, adjusted to the year 2000, the year the report by HDR Engineering, Inc. was prepared. Present day (year 2000) capital cost was divided by the plant capacity to yield the unit cost for plant construction in dollars per gallon per day ($/gpd). Operation and maintenance costs generally included labor, chemicals, power, membrane/parts replacement, concentrate disposal, and other recurring costs. The operations and maintenance costs were reported as if the plants were operating at 100% of its design capacity. Of these items, labor and power are generally the items of greatest cost. The total treated water cost curves were calculated as the sum of the amortized capital costs and the operations and maintenance costs. Annual debt service was computed using 8 percent interest over a 20-year period. Figure 4-6 shows that the unit costs for the different operating treatment plants are highly variable. The survey does not account for differences in source water quality, except for water source type (surface water versus groundwater). Groundwater desalination capital costs range from $2/gpd to $4/gpd. Figure 4-8 shows the total treated water cost for groundwater desalination as reported by the survey. Total treated water costs range from $1.50/1,000 gal to $2.75/1,000 gal and exhibit economies of scale.

In the survey conducted by HDR Engineering, Inc., only two plants using EDR responded. These plants have a design capacity in the 7 MGD range and use EDR to desalinate surface water and conventional pretreatment to control fouling. The total capital costs reported for the survey were

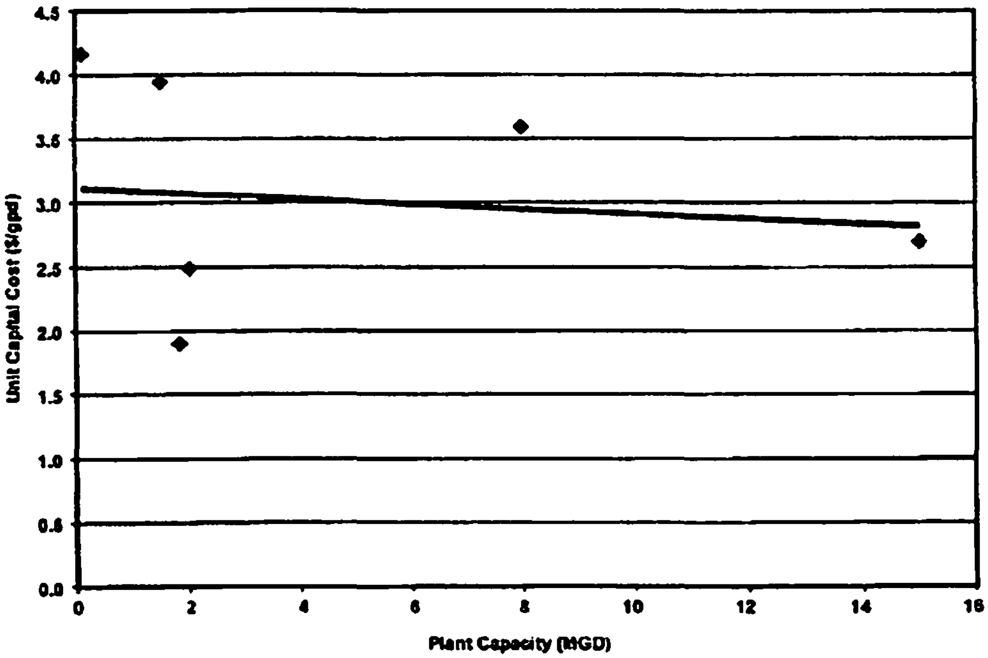


Figure 4-6. Unit capital cost of groundwater desalination by reverse osmosis based on a survey of operating plants (symbols) conducted by HDR Engineering, Inc. (figure taken from HDR Engineering, Inc., 2000). The curve is a regression line to the survey data.

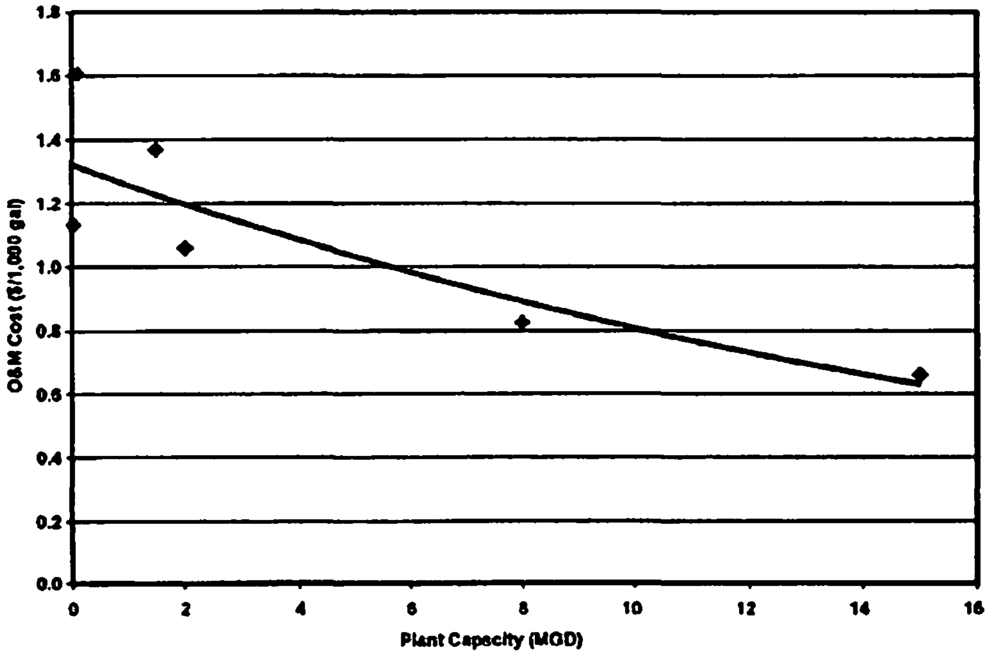


Figure 4-7. Operation and maintenance (O&M) cost of groundwater desalination by reverse osmosis based on a survey of operating plants (symbols) conducted by HDR Engineering, Inc. (figure taken from HDR Engineering, Inc., 2000). The curve is a regression line to the survey data.

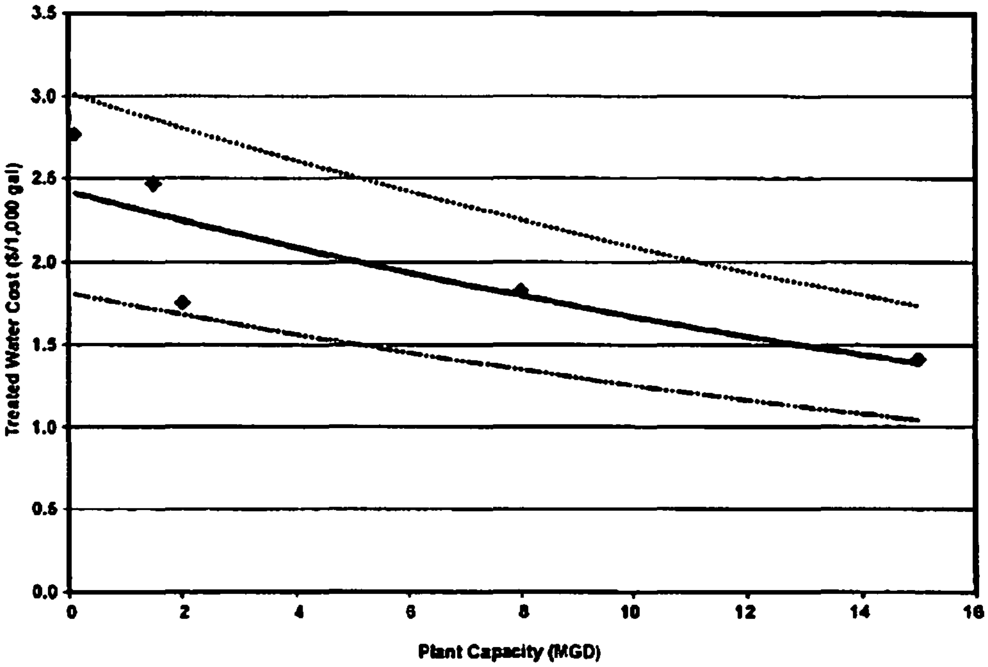


Figure 4-8. Total cost of groundwater desalination by reverse osmosis based on a survey of operating plants (symbols) conducted by HDR Engineering, Inc. (figure taken from HDR Engineering, Inc., 2000). The curve is a regression line to the survey data.

$2.05/gpd and $1.15/gpd. The operation and maintenance costs reported were $0.62/1,000 gal and $0.66/1,000 gal, and the total treated costs were reported to be $1.00/1,000 gal and $1.20/1,000 gal.

Another survey of EDR plants was published recently by Leitner and Associates (1997). The survey results included data on the desalination process used, capital costs, and operation and maintenance costs. The survey results for 17 brackish water EDR plants are reproduced in Tables 4-3 to 4-5.

4.3 ESTIMATED COST OF REVERSE OSMOSIS AND ELECTRODIALYSIS REVERSAL FROM VENDOR AND FROM LITERATURE

Estimated costs for reverse osmosis treatment of Edwards Aquifer saline waters were solicited from three RO system vendors. Only one of the vendors (James Almond, sales engineer – municipal market, Osmonics, Inc.) responded with information (Table 4-6) in time to be included in this report. The equipment pricing is for the RO equipment only and includes cartridge prefiltration, chemical pretreatment, high pressure RO pump, membrane housings and elements skid mounted, instrumentation, controls, clean in place system, freight and start up assistance. The price does not include any installation costs. The operating costs are for everything associated with the RO system: power costs, membrane replacement, cartridge filter replacement, cleaning, and chemical pretreatment. It does not include any cost associated with getting the water to the plant or additional pretreatment or post-treatment that may be required.

Table 4-3. Brackish water electrodialysis reversal process data from a survey conducted by Leitner and Associates (1997)

Plant name/owner	Rated capacity (GPD)	Annual production (gal)	Quantity of water added (GPD)	Feed salinity (mg/L)	Product salinity (mg/L)	Product recovery (%) .	cost of electrical power ($/kWh)
Kona Village Resort, Hawaii	30.000	11980.000	None	1,600/4,000	250	60	0.11
Dell City, Texas	100,000	12980,000	120,000	3.900	500	45	40/day
Georgetown, S. Carolina	190,000[1]	34,800,000	50,000	956	347	65	0.104-.073
Granbury, Texas WTP	300,000[2]	47.500000		600	47		0.098
Nokomis, FL (Sorento)	300,000	60,000,000	340,000(RO)	3,000	350	80	700/mo
Alta, Iowa WTP	432,000	75,500,000		1.200	340	85	0.078
Melville, Saskatchewan, Canada	500,000	184,000,000	325,000	1.900	600	80	0.135
Ewa Beach, Hawaii	500,000	66,430,000	125,000	600	40	70	
Coupville, Town of (WA)	528.000	44,000,000	150,000	900	450	75	
Yuma Proving Ground (AZ)	600,000	51,000,000		1.800	200	85	0.035
Buckeye, Town of (AZ)	1,000,000	346.000.000		1.600	600	75	
Washington, City of (Iowa)	1.900,000	300,000,000	98,000	1.200	628	13(?)	3,000/mo
Foss Reservoir, Oklahoma	2,800,000	449,000,000	0	1.050	240	70	0.028
Lake Granbury, Texas WTP	3.500.000	500,000,000	35%	321	97	75	0.058
Suffolk, City of (VA) WTP	3.750.000	730.000000	1,000,000	193	50	94	0.06
Sherman, City of (Texas)	4,500,000	821,000,000	2,500,000	1.350	650	85	0.049
Venice, FL (City of Carlton)	12,000,000		0	1.030	298	87	

[1] Annual production based on rated capacity (190,000 GPD) with unit operating at an average 0112 hours/day.

[2] Three-year average.

Table 4-4. Brackish water electrodialysis capital costs based on a survey conducted by Leitner and Associates (1997)

Plant name/owner	Building	Plant equipment	Wells or feed intake	Product tanks	Brine disposal	other costs	Total capital costs
Kona Village Resort Hawaii							
Dell City, TX	14.260	373,682	27,941	149,493			665.376
Georgetown, SC WTP							
Granbury, Texas WTP	700,000	500,000				67.645	1,267,645
Nokomis, FL (Sorento)							550,000
Alta, Iowa WTP	50,000	850.000					900,000
Melville, Saskatchewan, Canada	383.000	752.700	134.000		437.500	160.80	1,868,000
Ewa Beach, Hawaii	2.968.000	2,039,000	616.000	422.000	77,000	0	6.386.000
Coupville, Town of (WA) WTP						264.00	Approx. 750,000
Yuma Proving Ground (AZ)			Not included			0	2,200,000
Buckeye, Town of (AZ) WTP							1,050,000
Washington, City of (Iowa) WTP	280,000	1.212.600					1.807,000
Foss Reservoir, Oklahoma			Not included	Not included			
Lake Granbury, Texas WTP	3,900,000	2,200,000	32,000	600,000	230.000	315.00	6,100,000
Sherman, City of (TX)	2.000.000	3.000,000	8,000,000			0	13,000,000
Venice, FL (City of Carlton)							97,000,000

Table 4-5. Brackish water electrodialysis operation and maintenance costs based on a survey conducted by Leitner and Associates (1997)

Plant name/owner	Power required (kWh)	Annual power cost	Personnel (number)	Annual personnel cost	Annual parts/chem cost	Replacement membrane cost	Total annual costs	(*)
Kona Village Resort, Hawai		6.531	3	20.610	15.751		40.953	
Dell City, TX		8.352	1	36,000	1.500	5.000	50,852	NNY
Georgetown, SC WTP		15,000	1	35,000	1.000		50,000	YYY
Granbury, Texas WTP		5220	4	78.021	50265		133,506	YNN
Nokomis, FL (Sorento)		100.000	2	60,000			160.000	YYY
Alta, Iowa WTP	112	84,000	1	25.000	15.000		48.400 (est.)	
Melville, Saskatchewan, Canada	60	46.041	4	145.000	10.842	17,780	219.663	YNN
Ewa Beach, Hawai								
Coupville, Town of (WA) WTP	65							NNY
Yuma Proving Ground, AZ	553.400/yr[1]	19.369	3	252,000	23,000	5.000	299.369	NNY
Buckeye, Town of (AZ) WTP	164	109,573		34,000	14,000		157.573	YYY
Washington, City of, (Iowa) WTP	152	36.000	2	21.000	18,000	18.750	93.750	NNN
Foss Reservoir, Oklahoma	438	124.989	6	160,769	239,809	35260	560.829	NNN
Lake Granbury, Texas WTP	263	82,000	5	200.000			400,000	NNN
Suffolk, City of (VA) WTP	315	175.569	18	355,137	36,180	139,335	706,221	YNY
Sherman, City of (TX)		120.000	15	560.000	60,000	65,000	825,000	
Venice, FL Carlton WTP			8				1.236.766	

Note: (*) = Questions: Does power consumption include well pump power (Y/N)? Is the well pump on the same meter (Y/N)? Does power consumption include distribution power (Y/N)?

[1] Yuma Proving Ground, kWh/yr includes nonpotable system distribution for irrigation use. This system is separate from the potable system and pumps approximately 250 million gallons per year.

Table 4-6. Estimated cost of reverse osmosis systems as a function of plant capacity and total dissolved solids concentration of feed and treated water. Information provided by James Almond of Osmonics, Inc.

Plant Capacity	Total Dissolved Solids (mg/L)						
	Feed	3,000	3,000	5,000	5,000	10,000	10,000
	Treated	500	1,000	500	1,000	500	1,000
2,000 gpm (2.88 MGD)							
RO Equipment Price		$1,350,000	$1,100,000	$1,575,000	$1,325,000	$1,650,000	$1,560,000
Annualized RO Equipment Price*		$119,917	$97,710	$139,903	$117,696	$146,565	$138,571
RO Equipment Price/1000 gal		$0.11	$0.09	$0.13	$0.11	$0.14	$0.13
Operating Cost/1000 gal		$0.25–0.30	$0.25–0.30	$0.33–0.37	$0.28–0.33	$0.38–42	$0.36–0.40
Total Cost/1000 gal		$0.36–0.41	$0.34–0.39	$0.46–0.50	$0.39–0.44	$0.52–56	$0.49–0.53
10,000 gpm (14.4 MGD)							
RO Equipment Price		$6,100,000	$5,035,000	$6,800,000	$6,040,000	$7,500,000	$7,125,000
Annualized RO Equipment Price*		$541,847	$447,246	$604,027	$536,518	$666,206	$632,895
RO Equipment Price/1000 gal		$0.10	$0.09	$0.11	$0.10	$0.13	$0.12
Operating Cost/1000 gal		$0.25–0.30	$0.25–0.30	$0.33–0.37	$0.28–0.33	$0.38–42	$0.36–0.40
Total Cost/1000 gal		$0.35–0.40	$0.34–0.39	$0.44–0.48	$0.38–0.43	$0.51–55	$0.48–0.52

*Amortization at 8% and 20 yr

Table 4-6 (continued). Estimated cost of reverse osmosis systems as a function of plant capacity and total dissolved solids concentration of feed and treated water

Plant Capacity		Total Dissolved Solids (mg/L)					
	Feed	3,000	3,000	5,000	5,000	10,000	10,000
	Treated	500	1,000	500	1,000	500	1,000
20,000 gpm (28.8 MGD)							
RO Equipment Price		$11,000,000	$9,065,000	$12,300,000	$10,935,000	$13,670,000	$12,950,000
Annualized RO Equipment Price*		$977,102	$805,221	$1,092,577	$971,328	$1,214,271	$1,150,315
RO Equipment Price/1000 gal		$0.09	$0.08	$0.10	$0.09	$0.12	$0.11
Operating Cost/1000 gal		$0.25–0.30	$0.25–0.30	$0.33–0.37	$0.28–0.33	$0.38–42	$0.36–0.40
Total Cost/1000 gal		$0.34–0.39	$0.33–0.38	$0.43–0.47	$0.37–0.42	$0.50–54	$0.47–0.51

*Amortization at 8% and 20 yr

Estimated costs are also available from published literature. For example, the Bureau of Reclamation report *"Water Treatment Primer for Communities in Need"* (Jurenka et al., 2001) includes total plant cost fact sheets for nine of the more common best available technology water treatment techniques. Figures 4-9 and 4-10, taken from the Bureau of Reclamation report, illustrate the estimated 2001 total plant cost for RO and EDR systems, respectively. The estimated costs were calculated also using the WaTER program (Jurenka et al., 2001).

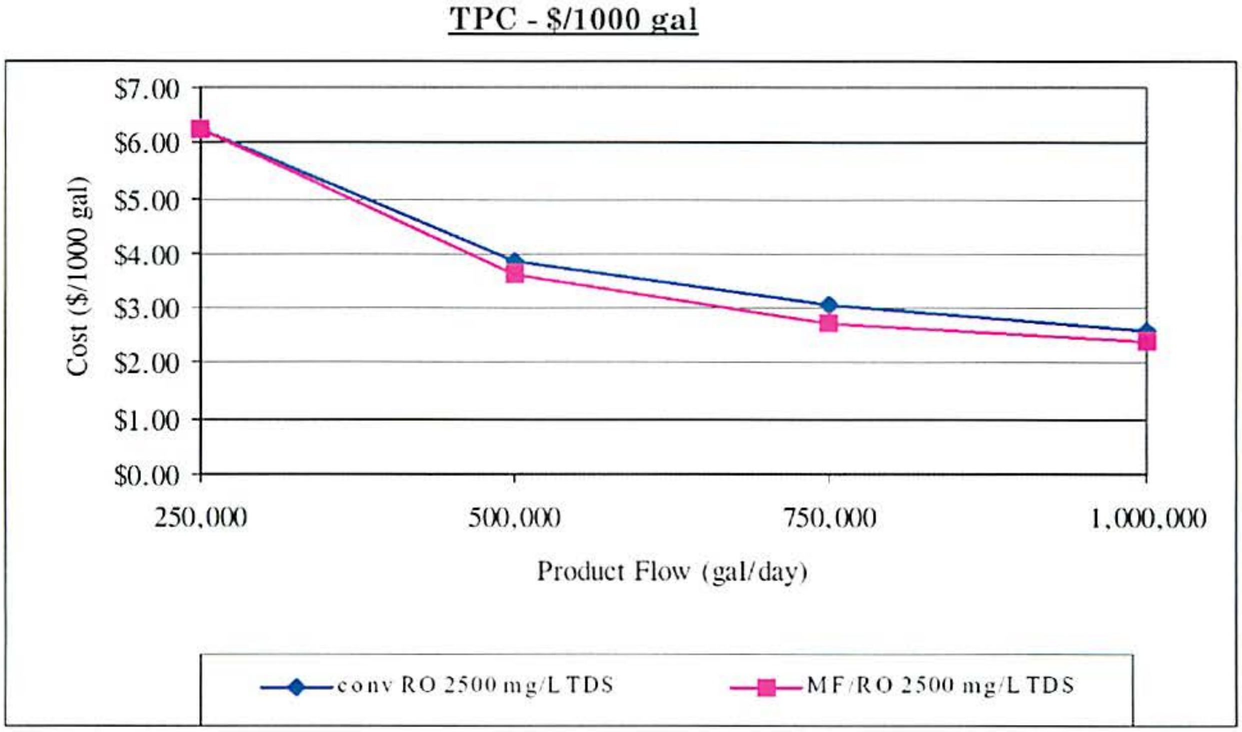


Figure 4-9. Estimated 2001 total plant cost (TPC, $/1000 gal) for a reverse osmosis system, with or without a microfiltration (MF) unit, as a function of plant capacity (gal/day). Figure taken from Jurenka et al. (2001).

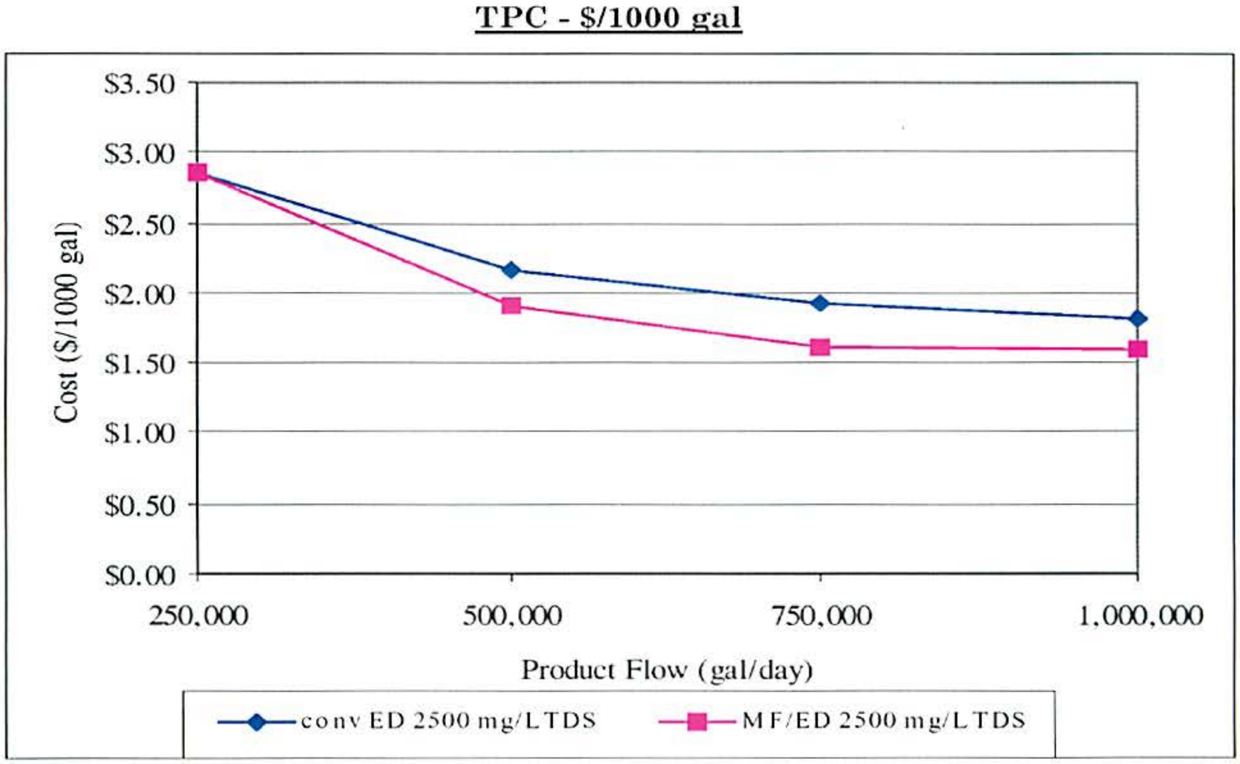


Figure 4-10. Estimated 2001 total plant cost (TPC, $/1000 gal) for an electrodialysis reversal system, with or without a microfiltration (MF) unit, as a function of plant capacity (gal/day). Figure taken from Jurenka et al. (2001).

5 POTENTIAL WELL YIELD FROM EDWARDS AQUIFER SALINE-WATER ZONE

In evaluating the feasibility of an Edwards Aquifer region desalination facility, an important question that needs to be addressed is whether sufficient water can be pumped from the zone to supply feedwater to the desalination plant. In this section, an attempt is made to address this question by calculating the potential well yield from Edwards Aquifer saline zone wells based on information from existing wells.

5.1 GEOHYDROLOGICAL SETTING

The fresh water portion of the Edwards Aquifer is bounded to the south and southeast by a high salinity zone, arbitrarily demarcated at locations where the groundwater total dissolved solids concentration exceeds 1,000 mg/L. Although the transition interface is commonly referred to as the saline-water (or badwater) line where the 1,000 mg/L interface intersects the ground surface, it is actually a surface whose geometry and orientation are complex. The length of the saline-water line exceeds 180 miles, from west of Uvalde to Kyle, Texas. The nature of salinity along the boundary is not uniform and has been delineated into six facies by Oetting et al. (1996).

In general, the aquifer is thicker (i.e., 500–700 ft) and has higher porosity (i.e., ~ 0.3) in the saline-water zone compared with the fresh water zone (< 500 ft and ~ 0.2, respectively). However, the flow velocities in the fresh water zone (90 to 900 cm/day) exceed those in the saline-water zone (1 to 2 cm/day) (Oetting et al., 1996). The location of transition between the fresh- and saline-water zones has been refined using borehole data gathered by the San Antonio Water System (SAWS) during their ongoing 10-year saline-water line investigation. The most current map of the saline-fresh water interface (Figure 5-1) was presented by Alvin Schultz at an Edwards Aquifer Technical Advisory Group meeting in April 2001 based on work performed for SAWS.

The width of the transition zone is spatially variable. If the width of the transition zone is determined using the distance between the 1,000 and 10,000 mg/L surfaces, the transition zone is relatively narrow in Uvalde and Medina Counties, increases in width in Bexar and Guadalupe Counties, and decreases at the east end in Comal County. Because the cost of desalination, in general, increases with the total dissolved solids concentration, it is desirable to use desalination feedwater with low total dissolved solids concentration. Therefore, it would be beneficial to locate wells for a desalination facility in an area with groundwater that has total dissolved solids concentration closer to 1,000 mg/L than to 10,000 mg/L. In addition, it would be desirable to locate the wells in an area with a small gradient in total dissolved solids concentration to insure a large pool of water with low total dissolved solids concentration, i.e., an area with a broad transition zone. Based on this criterion, southeastern Bexar County or western Guadalupe County would be preferable well sites relative to other areas. It should be noted that decisions on where to locate extraction wells would be based on several criteria, not only those discussed in this section. Therefore, the following discussion includes wells that represent much of the 180-mile Edwards Aquifer saline-fresh water transition zone, not only southeastern Bexar and western Guadalupe Counties.

5.2 SALINE-FRESH WATER TRANSITION ZONE HYDRAULIC INFORMATION

Information from existing wells has been assembled to evaluate the availability of water from the saline portion of the Edwards Aquifer. Data from 13 wells located either partially or entirely in aquifers with saline quality groundwater (i.e., > 1,000 mg/L total dissolved solids) are used to estimate the potential yield of wells that could provide feedwater for a desalination facility. Information on these

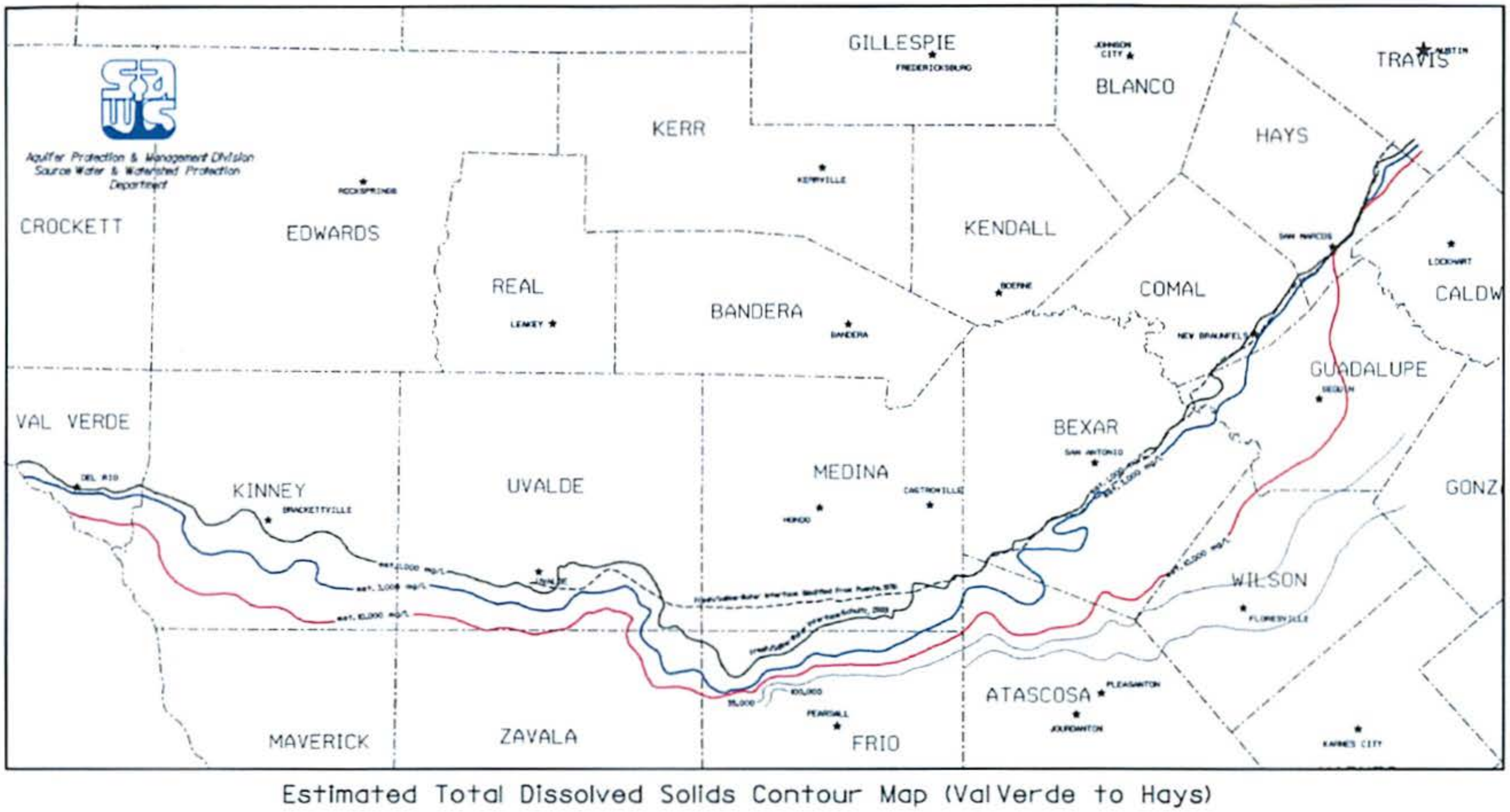


Figure 5-1. Location of the saline/fresh-water interface (Schultz, 2002).

Table 5-1. Summary of Edwards Aquifer wells included in this evaluation

Well	Conductivity[#] (mmhos/cm)	Depth of Well (ft)	Saline Zone (ft)
Artesia A	2,680–6,650	1,489	Total depth
Artesia C*	3,860–5,870	1,398	859–1,398
Artesia D*	6,380	1,384	1,225–1,384
New Braunfels A	2,413–5,540	936	Total depth
New Braunfels B*	990–3,750	916	472–916
New Braunfels C*	1,050–4,190	959	518–959
San Marcos B	12,420–14,680	890	Total depth
San Marcos C	13,000–14,710	920	Total depth
San Marcos D	11,230–16,405	774	Total depth
Tri-County No.3	7,800	1,219	Total depth
Tri-County No.4	5,000	1,562	Total depth
Uvalde No.3	2,980	1,403	Total depth
Uvalde No.4	2,680	1,466	Total depth

*Denotes wells that penetrate both fresh and saline zones in the Edwards Aquifer

[#] Total dissolved solids concentration is 0.55–0.75 times the conductivity (Todd, 1980)

wells is summarized in Table 5-1. Seven of the wells are located entirely in the saline zone. Five of the wells penetrate both saline and fresh water in the Edwards Aquifer.

The 13 wells are part of the saline-water line study led by SAWS. Because there has been no compelling motivation to measure the hydraulic properties of the saline portion of the Edwards Aquifer, minimal hydraulic testing of these wells has been performed. Therefore, only limited hydraulic information for this region is available. Nonetheless, approximations of aquifer hydraulic property have been made to estimate the potential yield of the aquifer. Two methods of analysis were used to estimate the potential yield of the wells, one for those cases where the hydraulic properties of the well are available and another where the hydraulic properties are not available but specific capacity can be estimated. The following sections describe each method.

5.3 POTENTIAL YIELD ESTIMATED USING HYDRAULIC PROPERTY VALUES

Results from aquifer tests performed at the New Braunfels and San Marcos well transects are reported in Poteet et al. (1992). The aquifer tests were performed using multiple wells at each well transect. The aquifer thicknesses at all three well locations in New Braunfels are assumed constant at 357.3 ft. Similarly, the aquifer thickness at San Marcos is 503.6 ft. The transmissivity was calculated for three observation wells (5,900 gpd/ft at A, 4,600 gpd/ft at B, and 9,000 gpd/ft at C) and for one pumping well (3,200 gpd/ft at C) during the aquifer tests at the New Braunfels transect. Transmissivity was calculated for one observation well (5,800 gpd/ft at B) and one pumping well (3,200 gpd/ft at C) at the San Marcos transect.

The potential yield at these wells, assuming a homogeneous and infinite aquifer, can be estimated using the Theis equation for a confined aquifer (Driscoll, 1989):

$$s = \frac{114.6QW(u)}{T} \tag{5-1}$$

where:

s = drawdown, in ft
$W(u)$ = well function
Q = well yield, or pumping rate, in gpm
T = transmissivity, in gpd/ft

and u is defined as

$$u = \frac{1.87r^2S}{Tt} \tag{5-2}$$

where:

r = distance, in ft, from the center of a pumped well to a point where the drawdown is measured
S = coefficient of storage, dimensionless
t = time, days

The potential well yield was calculated for three wells at the New Braunfels well transect and for two wells at the San Marcos well transect assuming a 30-day duration of pumping and assigning a maximum drawdown at the pumping well equivalent to two-thirds of the available drawdown. Available drawdown is the difference between the total well depth and the static water level. Transmissivity was not measured at the San Marcos D well, thus well yield was not estimated for this well. A well radius of 1 ft was assumed in the calculations. Results from these calculations are summarized in Table 5-2.

5.4 POTENTIAL YIELD ESTIMATED USING SPECIFIC CAPACITY

Pump and recovery data collected during well testing were used to estimate the potential well capacity for seven wells: Artesia wells A, C, and D, Tri-County wells No. 3 and No. 4, and Uvalde wells No. 3 and No. 4. In the absence of controlled aquifer testing, drawdown measured during the pump/recovery tests was used to roughly estimate the specific capacity of the wells. Specific capacity of a well is defined as its yield per unit of drawdown, expressed here as gallons of water per minute per foot of drawdown (gpm/ft) (Driscoll, 1989). Specific capacity typically varies with time. As the duration of pumping increases, the specific capacity decreases. Therefore, it is not possible to simply calculate the specific capacity for a well after a short duration of pumping and extrapolate this value to the available

Table 5-2. Summary of potential well yield estimated using aquifer test results

Well	Static Water Level (ft)[#]	Well Depth (ft)	67 % Saturation Well Depth (ft)	Storage	Trans-missivity (gpd/ft)	Potential Well Discharge (gpm)
New Braunfels A	50	936	590	0.000092	5,900	1,500
New Braunfels B*	50	916	580	0.001598	4,600	1,350
New Braunfels C*	50	959	610	0.000064	9,000	2,575
San Marcos B	50	890	560	0.000198	5,800	1,450
San Marcos C	50	920	580	0.000198[‡]	3,200	860
San Marcos D	50	774	480	—	—	—

*Denotes wells that penetrate both fresh and saline zones in the Edwards Aquifer
[#] Below ground level. Assumed values because static water levels are unknown.
[‡] Storage for San Marcos B assigned to San Marcos C

drawdown to estimate the theoretical yield of the well.

Analyses by Driscoll (1989) indicate that the minimum specific capacity is approximately 50 percent of the maximum specific capacity. Further analysis by Driscoll (1989) indicates that at 67 percent of maximum drawdown, 90 percent of the maximum yield is obtained. Thus, it is generally uneconomical to operate a well with drawdown greater that 67 percent. Note that the maximum specific capacity theoretically corresponds to zero drawdown because there is no reduction in the saturated thickness (Driscoll, 1989).

Preliminary estimates of well yield for these seven wells were made using the criteria by Driscoll (1989). In summary, the short-term pumping specific capacities reported by Denise Renaghan (personal communication to R. Green, January 2003) were considered as the maximum specific capacity. These values were reduced by one-half and the potential well yield was calculated for 67 percent of the total available drawdown. The estimated potential yield for these eight wells is presented in Table 5-3.

5.5 SUMMARY OF POTENTIAL WELL YIELD ESTIMATES

The potential well yield was estimated for eleven wells. Estimates that were made using the relatively long-duration aquifer test (i.e., 9 hours) have higher degrees of confidence than estimates of specific capacity that were made using the relatively short-duration pump/recovery tests (i.e., ~30 minutes). It is therefore not surprising that the range of potential well yield values estimated using specific capacity is large. In particular, the potential well yield calculated for Uvalde No. 3 and No. 4 may be unrealistically high. This uncertainty can be reduced using results from aquifer tests performed by the U.S. Geological Survey. However, these results were not available at the time of this analysis.

Nevertheless, the potential yield from wells in the Edwards Aquifer saline zone estimated using two methods indicate that sufficient feedwater can be pumped from the saline portion of the Edwards Aquifer to supply a moderate-sized desalination facility. For example, assuming a well yield of 1,300 gallons per minute and a well utilization of 12 hours per day, a single well can produce 936,000 gallons per day of feedwater. Five such wells would be needed to supply feedwater to a desalination plant that has a 75% water recovery and a capacity of 10 acre-feet per day (3.3 MGD). Desalination plants with higher capacity would require more wells and/or wells with higher yield.

Table 5-3. Summary of potential well yield estimated using measured specific capacity

Well	Static Water Level (ft)#	Well Depth (ft)	67 % Sat. Well Depth (ft)	Specific Capacity (gpm/ft)	Potential Well Yield (gpm)
Artesia A	25	1489	980	0.3–5.0	150–2,450
Artesia C*	50	1398	900	1.9–3.9	855–1,750
Artesia D*	20	1384	910	–	–
Tri-County No.3	20	1219	800	4.7	1,880
Tri-County No.4	10	1562	1,030	1.12	575
Uvalde No.3	90	1403	930	30	14,000
Uvalde No.4	90	1466	920	30–60	14,000–28,000

*Denotes wells that penetrate both fresh and saline zones in the Edwards Aquifer

#Below ground level

6 SUMMARY AND CONCLUSIONS

Desalination of saline groundwater potentially can be an important technology for ensuring a reliable water supply for some communities in the Edwards Aquifer region. The benefits of desalination include (Burroughs, 1997): (1) flexibility in facility size, (2) minimal reliance on extended delivery systems, (3) the opportunity for local control of water supplies, (4) reduced dependence on the Edwards Aquifer freshwater zone, (5) high quality potable water, and (6) a reliable water source even in times of drought. However, desalination facilities have high-energy consumption and require expensive capital investment. These factors make the cost of producing water by desalination substantially greater than by traditional pumping of Edwards Aquifer fresh water. Notwithstanding the higher cost of desalination, the technology must be considered by water authorities and entities in the Edwards Aquifer region because of the projected shortfalls in future water supplies derived primarily from the fresh water zone.

In this report, a preliminary study was presented on the feasibility of pumping and treating saline water from the Edwards Aquifer to produce potable water. Background technical information on RO and EDR was presented, along with summary information on existing and planned municipal water treatment plants employing similar technologies. Preliminary cost curves for the treatment of saline waters were calculated as a function of operating parameters and compared with cost data from existing plants, from the literature, and from a vendor. In the absence of detailed information on any proposed desalination facility (e.g., location, feedwater composition, plant capacity), the cost curves calculated using the WaTER program have limited accuracy, but the cost information derived from the surveys of existing plants could be used as reference points in evaluating the WaTER results.

The WaTER cost estimates are useful for comparing different process options. In addition, some qualitative conclusions regarding RO and EDR desalination systems can be made based on those cost estimates. For example, it is clear that the unit cost of producing potable water from both systems depends strongly on the total dissolved solids concentration of the feedwater and on plant capacity. For RO systems, economies of scale are observed. EDR systems also show economies of scale, but in a narrower range of plant capacity. EDR is cost competitive with RO for small to moderate-sized facilities [~10 acre-ft/day (~3.3 MGD)] if the feedwater total dissolved solids concentration does not exceed 3,000 mg/L.

Preliminary estimates of potential yield from eleven wells that tap the Edwards Aquifer saline-water zone were made to determine if sufficient saline water could be drawn from the aquifer for use as feedwater for a desalination facility. The results indicate that sufficient water is available to supply feedwater to a desalination plant, at least to one of moderate size. It may be possible to supply a plant with larger capacity by using more wells or wells with higher yield. Based on a recent map of the Edwards Aquifer saline-fresh water interface, zones with a broad saline-fresh water transition zone are present in southeastern Bexar County or western Guadalupe County. These areas are potentially suitable locations for wells supplying feedwater to a desalination facility.

The fact that many communities in Texas and elsewhere have existing desalination plants, or are planning or in the process of building one, shows that desalination can be an economically viable technology. However, the conditions under which desalination is appropriate for communities in the Edwards Aquifer region must be carefully identified and considered in the context of an overall water supply management strategy that includes consideration of other water sources, conservation measures, and long-term economic and environmental ramifications.

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APPENDIX

Input and Output Worksheets of WaTER Cost Estimation For a Reverse Osmosis or Electrodialysis Reversal Plant with a Capacity of 10 acre-ft/day (3.3 MGD), Feed and Product Water Total Dissolved Solids Concentration of 5,000 and 500 mg/L, Respectively

FLOW RATE INPUT PAGE, WATER DATA REPORT		Links to this page are ORANGE			
Yellow colored cells are mandatory input cells					
Enter Availability.					
Plant availability due to down time (used to estimate production/year):	0.85				
Planned operation time per day (used to calculated energy & chemical cost):	1.00				
	L/M	GPH	GPD	MGD	Acre-Ft/year
INPUT CELLS: enter flowrate in ONE of these cells, set rest cells to 0=>		0	0	3.3	0
Flow rate converted to Liters/second and entered in workbook calculations.	0.00	0.00	0.0	144.6	0.0
Flow rates converted to a variety of units.	8674	137,500	3,300,000	3.30	3696.24
PLANT FLOW RATES	**L/S**	**GPM**	**MGD**		
Required Plant Feed Flow Rate:**	**193**	**3056**	**4.4**		
Desired Plant Product Flow Rate:	**145**	**2292**	**3.3**		
**Feed Flow = Plant Product Flow / RO Recovery entered on cost report.					
WATER DATA REPORTS (based on Water Analysis)					
Total dissolved solids (TDS):	4997	mg/L			
Average equivalent wt.:	31.5	g/equiv			
Total equiv./L:	0.158	eq/L	1.10E-01	mol/L	
Total equiv./L (Valence >+1):	0.054	eq/L	2.76E-02	>1 valence	
Average MW	45.25	g/mol			
Ionic Strength:	0.127	mole*charge^2/L			
Delta G:	-9.094				
LSI:	-6.545				
Tendancy to corrosion, may need remineralization.					

Cost Index

COST INDICES DATA:

Input: Current Cost Indicies Values

These are derived from the ratio of the origianal 1978 basis values for the original indexes and their values in May of 2000

$I_r*(I_c/I_{5/2000})$ — I_c is the current value of the ENR indices, I_r is the 1978 value of the BLS index, $I_{5/2000}$ is the 5/2000 value of the BLS index.

Cost Indices Categories:	Input Current Values Here December 2, 2002	Used For	New 1978 Basis	Ratio for 1978	Ratio for 2000
ENR Construction Cost Index	6,562.73	Manufactured & Electrical Equipment	3384.05	1.94	1.05
ENR Building Cost Index	3,640.11	Housing	3113.20	1.17	1.02
ENR Skilled Labor Index	6,338.14	Excavation, Site Work & Labor	5335.92	1.19	1.11
ENR Materials Index	1,991.51	Piping & Valves	1973.80	1.01	0.90
ENR Steel Cost ($/cwt)	26.42	Steel	29.68	0.89	0.93
ENR Cement Cost ($/ton)	82.70	Concrete	67.54	1.22	1.02
ENR Materials Index	1,991.51	Maintanace Materials	2036.81	0.98	0.90
Electricity Cost ($/kWhr)	0.07	Power	0.03	2.33	1.00
ENR Labor Rate ($/hr)	35.18	Labor	21.22	1.66	1.11
Interest Rate (%)	8.00	On Construction & Bond Money			
Amortization time (yr)	20.00	For Bond Period			
Water Rate ($/kgal)	50.00	Cost of Feed Source Water			

ENR - Engineering News Record Construction Cost Index published monthly by McGraw Hill in New York City (212-512-2000)

See http://www.enr.com

WATER ANALYSIS

Input analysis in Yellow cells

Links to this page are GREEN

Example

Component	Water Analysis	Units	MCL (mg/L)	Amount Over MCL	Valence Charges	Molecular Wt.	Equivalent Weight	Moles/ Liter	Equiv./ Liter	Ionic Strength	mg/L as $CaCO_3$	
METALS:	Agricultural Influence											
Aluminum		mg/L	0.05		3	26.97						
Antimony		mg/L	0.006		3	121.75						
Arsenic		mg/L	0.05		3	74.92						
Barium		mg/L	2		2	137.33						
Beryllium		mg/L	0.004		2	9.01						
Cadmium		mg/L	0.005		2	112.41						
Calcium	659.54	mg/L	---		2	40.08	20.04	1.65E-02	3.29E-02	6.58E-02	1645.56	0.66
Chromium, total		mg/L	0.1		2	52						
Copper		mg/L	1		2	63.55						
Iron		mg/L	0.3		2	55.85						
Lead		mg/L	0.015		2	207.2						
Magnesium	250.17	mg/L	---		2	24.3	12.15	1.03E-02	2.06E-02	4.12E-02	1029.51	0.25
Manganese		mg/L	0.05		2	54.94						
Mercury		mg/L	0.002		2	200.59						
Nickel		mg/L	---		2	58.71						
Potassium	32.98	mg/L	---		1	39.1	39.10	8.43E-04	8.43E-04	8.43E-04	84.34	0.03
Selenium		mg/L	0.05		4	78.96						
Silver		mg/L	0.1		1	197.87						
Sodium	579.94	mg/L	---		1	22.99	22.99	2.52E-02	2.52E-02	2.52E-02	2522.58	0.58
Strontium		mg/L	---		2	87.6						
Zinc		mg/L	5		2	65.38						
INORGANICS:												
Alkalinity-Bicarbonate	334.32		---		-1	61	61.00	5.48E-03	5.48E-03	5.48E-03	548.06	0.33
Alkalinity-Carbonate			---		-2	60						
Carbon Dioxide (aq)			---		0	44						
Chloride	1068.91	mg/L	250	**818.9106**	-1	35.45	35.45	3.02E-02	3.02E-02	3.02E-02	3015.26	1.07
Cyanide, free		mg/L	0.2									
Fluoride	3.53	mg/L	4		-1	19	19.00	1.86E-04	1.86E-04	1.86E-04	18.55	0.00
Nitrate (as N)			10		-1	14						
o-Phosphate			---		-3	95						
Sulfate	2046.85	mg/L	250	**1796.85**	-2	96	48.00	2.13E-02	4.26E-02	8.53E-02	2132.14	2.05
Silica	23.88											
pH		pH	6.5-8.5		1	1						
Solids (TDS)	4996.59	mg/L	500	**4496.589**								
Total Suspended Solids		mg/L	---									
Conductivity			---									
Temperature	31.00		---									

Cations Equiv./L	7.96E-02	C1	Acidity	548		45.25	Average MW
Anion Equiv/L	7.85E-02		Alkalinity	548		1.10E-01	Sum Moles/L
Ratio Cat/An	1.01	C2		-1097		1.37	average - chai
Anions Equiv./L - HCO3 & SO4	3.03E-02					1.52	average + cha
Sum TDS	5000.11					0.253	activity coeffic
						487.67	Osmotic Pres:

Cost reports for water treatment processes

Yellow colored cells are mandatory input cells

Links to this page are Yellow too.

	Design		Planned
Capacity (Plant Output):	12,491	m^3/day	3875178 m^3/yr
	3,300	kgal/day	1204500 kgal/yr

Process	Cost Parameter	Units		Construction Cost Total $1000	$/m^3 Cap	$/kgal Cap	Operating Cost $1000/yr	$/m^3	$/kgal
Desalination									
Microfiltration				$2449	$196	$742	$624	$0.16	$0.52
Microfilter system equipment	Memcor, 90M10C								
Number of microfilter	4								
Recovery	0.99								
Reverse Osmosis/Nanofiltration				$4664	$373	$1,413	$1514	$0.39	$1.26
Membrane Type	Film Tec, BW30-400								
Number of elements	288								
Operating Pressure	2610	kPa	378 lb/in^2						
NaCl Rejection	0.995								
Recovery	0.75								
Target Product (TDS mg/L)	500								
Blending? (Y or N)	Y								
Ratio Blend:Product %	9.4								
Ion Exchange			Base Cost	$138	$11	$42	$32	$0.01	$0.03
Cation Equivalents/L Resin	20								
$/m^3 Cation Exchange Resin	$1,607		45 ft^3	$55	$4	$17			
Cation Resin Volume:	68	m^3	2,424 ft^3						
To Remove Cation Equivalents/L:	5.43E-02								
Anion Equivalents /L Resin	11								
$/m^3 Anion Exchange Resin	$6,250		175 ft^3	$212	$17	$64			
Anion Resin Volume	68	m^3	2,424 ft^3						
To Remove Anion Equivalents /L:	7.85E-02								
Run Cycle (days)	1								
Ionics Electrodialysis Reversal				$4096	$328	$1,241	$1288	$0.33	$1.07
Staff Days/day	3								

RO&NF Input

Process Input	Calculates Blending			
Available Flow	158	L/s	2507	gpm
Target Flow	145	L/s	2292	gpm
Total Disolved Solids	4997	mg/L		
Target Disolved Solids	500	mg/L		
Mono-valent	0.75	Decimal		
Multi-valent	0.25	Decimal		
Average Molecular Mass	45.25	g/mol		
Allow Blending	Y	"Y"or "N"		
Recovery Rate	0.75	Decimal		
Product TDS	33	mg/L		
Product Flow	131.0	L/s	2076	gpm
Membrane Feed Flow	174.6	L/s	2768	gpm
Concentrate TDS	19886	mg/L		
Concentrate Volume	43.7	L/s	692	gpm
Bypass flow for blending	13.6	L/s	215	gpm
% blending	9.4	%		
Data from Membrane Manufacturer Specification			Calculates 'A'	
Type of membrane	Film Tec, BW30-400			
Membrane Diameter	20.32	(10.16 or 2	8.0	in
Productivity	40	m^3/day		
Area per module	37	m^2		
Operating pressure, P_{app}	1550	kPa	224.9	psi
Test solution TDS	2000	mg/L		
Avg. MW of TDS,	58.44	mg/mmole NaCl		
Chloride Rejection	0.995			
Sulfate Rejection	0.998			
Recovery Rate	15	%		
Temperature	25	°C		
NaCl dissociation constant	0.99			
C_f, conc. of salt in feed water	34	mole/m^3		
C_p, conc. of salt in product water	0.17	mole/m^3		
Cr, conc. of salt in reject	40	mole/m^3		
C_m, bulk conc.	37			
Osmotic pressure	184	kpa	27	psi
Net driving pressure, NDP_o	1366	kpa	198.3	psi
A, water transport coefficient	9.16E-12	$m^3 m^{-2} Pa^{-1} sec^{-1}$		
Determination of operating pressure				
User input pressure , NDP,	1366	kPa	198.2	psi
Ave Intrinsic Rejection	0.996			
Cf, conc. of salt in feed water	110	mole/m^3		
C_p, conc. of salt in product water	0.442	mole/m^3		
Cr, conc. of salt in reject	440	mole/m^3		
C_m, conc. of	275	mole/m^3		
Osmotic pressure, P_{osm}	1244	kpa	180	psi
Operating pressure, P_{app}	2610	kPa	379	psi

Colored cells are changeable here.
White cells are equations or taken from
the input, cost indices,and cost report worksheets.

Construction Cost Input				
	11,316	m^3/day	2,989,706	gpd
Bypass	1,174	m^3/day	310,294	gpd
Total Capacity	12,491	m^3/day	3,300,000	gpd
Module Productivity	40	m^3/day	10,566	gpd
Number of modules per vessel	6			
Max Vessels per Skid	36			
Number of Modules	288	# of modules		
Number of Pressure Vessels	48	for 2:1 array		
Number of RO Skids	2			
Recovery Rate	0.75	Decimal		
Chemical Feed Dosages				
Acid	0.00	mL Conc H_2SO_4/L		
Antiscalant	0	mg/L		
Disinfectant	0.0	mg/L		
Building Area	361	m^2	3887	ft2
Administrative Area	200	m^2	2153	ft^2
Odor Control?	n	Yes (Y) or No (N)		
Emergency Generatore Size	2.4	MW		
High Pressure Pump	VST			
Height Difference	10	m	32.81	ft
Pipe Diameter	0.06	m	2	in
Length of Pipe	10	m	32.8	ft
Efficiency	60			
Number of Pumps	5			
Differential Pressure	2610	kPa	379	psi
Capacity per Skid	0.087	m^3/s	1384	gpm
Size	605	hp		
Transfer Pumps	SST	*SST, VST or CSS		
Height Difference	2	m	6.56	ft
Pipe Diameter	0.07	m	0.23	ft
Length of Pipe	10	m	32.81	ft
Efficiency	60			
Number Transfer Pumps	5			
Pressure Differential	200	kPa	29.0	psi
Capacity per Pump	0.039	m^3/s	611.1	gpm
Size	14.1	hp		
Product Water Pump	SST	*SST, VST or CSS		
Height Difference	10	m	32.81	ft
Pipe Diameter	0.07	m	0.24	ft
Length of Pipe	20	m	65.62	ft
Efficiency	78			
Number Pumps	5			
Pressure Differential	101.3	kPa	14.7	psi
Capacity per Pump	0.029	m^3/s	458.3	gpm
Size	11	hp		

Operations & Maintenance Cost Input		
Chemical Costs		
Citric Acid	0.14	$/kg
Antiscalant	4.37	$/kg
Disinfectant	0.2	$/L
H_2PO_4	23.7	$/kg
NaOH	18	$/kg 50%
Membrane Life	3	Years
Cleaning Rate	4	per Year
Staff Days/day	3	

Estimating Construction Costs for BW-30-400 Membrane Treatment Plant								
Membranes	$	201,600			@	700	$/module	
RO Skids	$	240,000			@	5000	$/Vessel	
Building	$	388,662			@	1076	$\$/m^2$	$100/ft2
Electrical	$	264,914			With base of	614	$\$/m^3$	
Insturmentation & Controls	$	136,765			add $300,000 for top of the line DAC			
High Pressure Pumps	$	1,229,250		kW 903		6724122	kWhr/year	
Transfer Pumps	$	72,968		kW 53		392652	kWhr/year	
Product Water Pumps	$	62,300		kW 41		307893	kWhr/year	
Degasifiers	$	18,618		Total kW 997				
Odor Control	$	-						
Process Piping	$	214,637						
Yard Piping	$	122,463						
Chemical Feed w/ Pumps								
Acid	$	-				1	$/L storage for 30 days	
Antiscalant	$	-				1	$/L storage for 30 days	
Chlorine	$	-				1	$/L storage for 30 days	
Cartridge Filters	$	34,204	Maint Materials					
Membrane Cleaning Equip	$	20,000	Manf & Elect					
Contractor Engineering & Training	$	55,262	Labor					
Concentrate Treatment & Piping	$	33,537	Piping			13	$\$/m^3$	Concentrate
Generators	$	379,039	Electrical			2.4	MW	RO & Building
Sitework	$	200,578	Electrical		$	14.53	$\$/m^3$	
Total Direct Capital Costs	**$**	**3,674,797**						

Indirect Capital Costs			
Interest During Construction	183,740	5	% of Total
Contingencies	219,603	6	% of Total
A&E Fees, Proj. Management	439,138	12	% of Total
Working Capital	146,992	4	% of Total
Total Indirect Captial Cost	**989,473**		

Total Construction Cost	**$**	**4,664,271**

Cost per m^3/day capacity	**$**	**373**
Cost per gpd capacity	**$**	**1.41**

Estimating Operations & Maintenance Costs		
Electricity	$	519,727
Labor	$	308,177
Acid	$	-
Antiscalant	$	-
Chlorine	$	-
Membrane Replacement	$	60,306
Cleaning Chemicals	$	1,882
Cartridge Filters	$	108,252
Repairs and Replacement	$	16,489
Insurance	$	6,596
Lab fees	$	17,230
Total O & M Cost	**$**	**1,038,658**

Total Costs		
Capital Recovery	$	475,066
O&M	$	1,038,658
Annual cost	**$**	**1,513,724**
$\$/m^3$ Product	**$**	**0.39**
$/1000 gal Product	**$**	**1.48**
$/acre foot Product	**$**	**482**

Based on "Estimating the Cost of Membrane (RO or NF) Water Treatment Plants" By William B. Suratt, P.E., Camp Dresser & McKee Inc. Vero Beach Florida Presented at the AWWA Membrane Technology Conference, Reno, NV, 1995. also published as

"Estimating the cost of membrane water treatment plants." AWWA Proceedings

Membrane Technologies in the Water Industry. Orlando, Florida, March 10-13, 1991. pp631-647.

Bureau of Reclaimation Formulas for Pricing an EDR System
7/6/00

TDS		4997	mg/l
Capacity		3.30	mgd
Capital Cost	$	3,717,027	
Total Power		13.5	kWh/Kgal
Membrane Replacement	$	184,494	
WATER RECOVERY		85%	
Staff days/day		3	

All cost numbers are only within +,- 15%

Estimate Operations & Maintenance Costs		
Electricity	$	964,365
Labor	$	38,522
Membrane Replacement	$	184,494
Cleaning Chemicals	$	3,300
Cartridge Filters	$	39,994
Repairs and Replacement	$	18,585
Insurance	$	7,434
Lab fees	$	31,680
Total O & M Cost		**$1,288,374**

Total Costs	**$**	**3,717,027**
Capital Recovery	$	378,587
Annual cost	**$**	**1,666,961**
$/m^3 Product		$0.43
$/1000 gal Product		$1.63
$/acre foot Product		$530.61