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IMPROVING PERFORMANCE AND ECONOMICS OF RO SEAWATER DESALTING USING CAPILLARY MEMBRANE PRETREATMENT

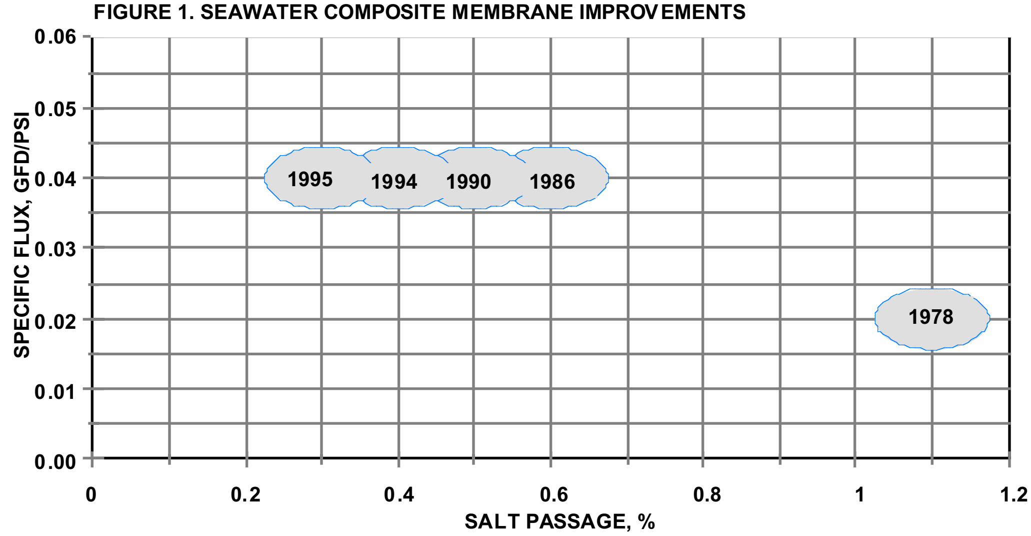
ABSTRACT

RO seawater systems that operate on a surface feed water originating from an open intake source require an extensive pretreatment process in order to control membrane fouling. Considerations of long term performance stability lead to a design concept of operation at a low permeate flux rate and low permeate recovery. In recent years the nominal performance of composite seawater membrane elements has improved significantly, and new effective water microfiltration technologies have been introduced commercially. These developments can be utilized to improve the quality of surface seawater feed to the level comparable to, or better than the water quality from the well water sources. These new developments enable a more advanced RO system design which should result in increased reliability and lower water cost. This paper will evaluate the applicability of new pretreatment technologies for seawater desalting and estimate their potential in improving the performance and economics of RO seawater systems. Results of operation of a hybrid UF/RO seawater system will be presented and evaluated.

Evolution of seawater membranes and operating parameters.

With introduction of composite polyamide membrane technology , the economics of the application of reverse osmosis to seawater desalting were significantly improved when compared to initial attempts to desalt seawater with cellulose acetate membranes. However, the early seawater composite membranes, introduced in 1978, which were based on aliphatic polyamide polymers, had relatively high salt passage. To produce potable quality water the RO systems equipped with the early membrane technology had to operate in a two pass configuration at a low recovery rate (1), usually in the range of 30% to 35%. With the development of the new generation of composite membranes, based on aromatic polyamide, the performances improved dramatically. Since 1986 (see Fig. 1), the salt rejection of subsequently introduced seawater membranes has been improving continuously without sacrificing water permeability. The latest offering of commercial seawater membranes has a nominal specific flux of twice the value of the 1978 technology and a salt passage of about four times lower. Currently, membrane elements with a nominal salt

FIGURE 1. SEAWATER COMPOSITE MEMBRANE IMPROVEMENTS



rejection of 99.7% and 6000 gpd (22.7 m3/day) nominal permeate flow are commercially available. The nominal membrane elements performance specifications do not translate directly to actual RO system performance, as the test conditions of single elements are significantly different than the operating parameters of an RO system. The system performance can be calculated based on intrinsic membrane parameters, installed membrane area, feed water composition and operating conditions (2).

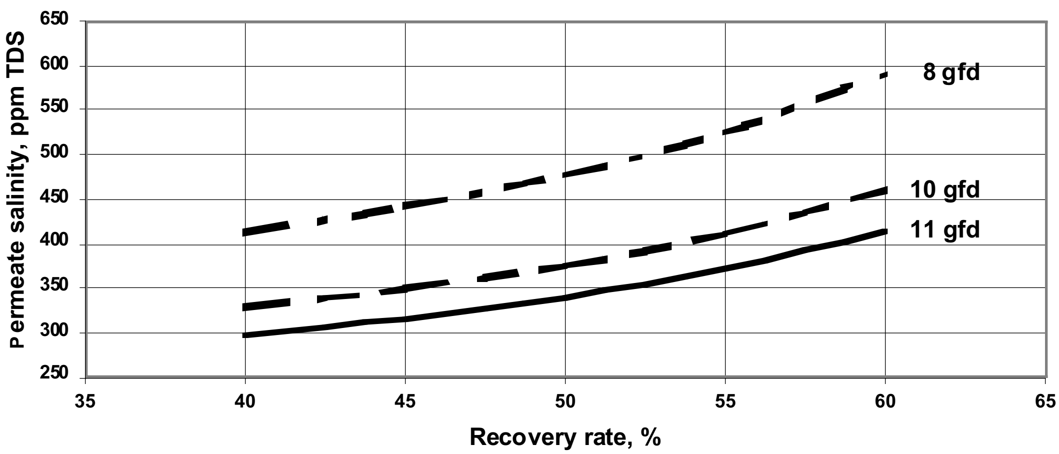
A lower salt passage of the membrane elements used (higher salt rejection) will result in lower permeate salinity. Therefore, increased salt rejection enables RO systems to operate at a higher recovery rate. For example, for Mediterranean seawater feed of about 38,000 ppm TDS salinity and water temperature in the range of 18 - 28 C, the RO systems are designed to operate at a recovery rate in the range of 40% - 45% and with an average permeate flux in the range of 7 - 8 gfd (11.9 - 13.5 l/m2-hr). At the above operating conditions the feed pressure is in the range of 800 - 1000 psi (55 - 70 bar) and permeate salinity is in the range of 300 - 500 ppm TDS.

For a given feed water salinity and salt rejection of the membrane elements used, the permeate salinity is a function of feed water temperature, recovery rate and permeate flux. An increase in feed water temperature results in an increased rate of salt and water diffusion across the membrane barrier at the rate of about 3% per degree Centigrade. Because RO plants usually operate at a constant flux rate, the changes of permeate salinity follow closely the changes in feed water temperature (2).

Permeate salinity is inversely proportional to the average permeate flux. Higher permeate flux increases the dilution of salt ions which passed the membrane, and therefore results in lower permeate salinity. The average permeate flux rate in seawater systems is maintained at relatively low values: 7 - 8 gfd for surface seawater feed and 10 gfd (16.8 l/m2-hr) for seawater from beach wells. The difference in flux rates between the two water source types results from better quality of the well water and therefore, a lower fouling rate of the membranes. These flux values are relatively low and only about 50% of the permeate flux values used in brackish RO systems. Attempts to operate seawater systems at higher flux rates have usually resulted in irreversible flux decline.

Until recently, the design recovery rate of new commercial seawater RO systems has been increased subsequently to the availability of membrane elements with increasingly higher salt rejection. So far, the maximum recovery in seawater RO systems has been mainly limited by the membrane salt rejection or the ability to produce permeate water of potable quality. Figure 2 displays permeate salinity as a function of recovery rate and permeate flux. The calculation were conducted for Mediterranean feed of salinity of 37,500 ppm TDS and feed temperature of 28 C for a recovery range of 40 - 60% and flux rate of 8 - 11 gfd. Nominal 99.6% salt rejection membrane elements were used. For calculations of permeate quality, the membrane salt passage was increased by 15%. This is to account for projected 5% per year salt passage increase during 3 years of an average membrane life (20% membrane element replacement per year). As expected, a higher recovery rate requires operation at an average flux rate above the standard value of 8 gfd. This is to maintain an acceptable permeate salinity, especially during the periods of high feed water temperature.

Figure 2. Projected permeate salinity for Mediterranean feed, 28C



Process economics

Recovery rate has a major impact on the economics of the seawater RO process. The size of all process equipment which is determined according to feed or concentrate flow will decrease with increased recovery rate. This applies to the size of the feed water supply system and power consumption of intake pumps. The size of all pretreatment equipment;

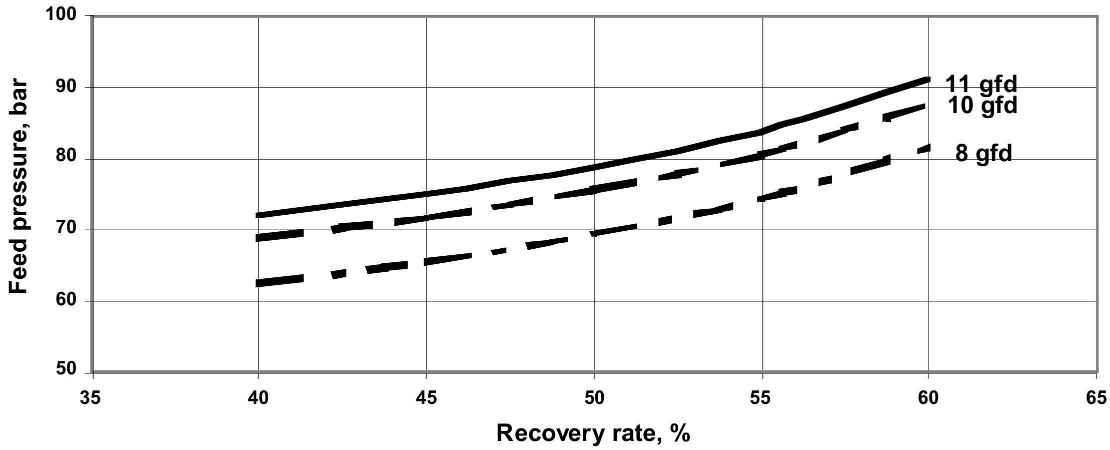
storage tank, booster pumps, filtration equipment and chemical dosing systems is determined according to the feed flow. The same considerations apply to sizing of concentrate piping and of the outfall facility. The design permeate flux rate affects the number of membrane elements installed, number of pressure vessels, manifold connections and size of membrane skid. The effect of the above parameters on investment cost will be examined on an example for a 6 mgd (22,700 m3/day) system operating on Mediterranean seawater from an open intake. The cost estimation of the conventional reference design is based mainly on the data developed by G. Leitner (3), P. Shields and I. Moch (4). Table 1 contains a comparison of equipment cost (including 35% indirect cost) of the basic design and a system operating at high recovery and high permeate flux, (HRF design). The basic design consists operation at 45% permeate recovery and 8 gfd flux rate. The HRF design consists operation at 55% recovery and flux rate of 11 gfd.

Table 1. Effect of recovery rate and permeate flux on cost of 6 mgd (22,700 m3/day) RO seawater system (in 1000 s US$).

Investment cost component	Reference design: 45% recovery, 8 gfd flux	HRF design: 55% recovery, 11 gfd flux	Equipment cost change, %
Intake and outfall	940	830	-11.7
Pretreatment	5,000	4,390	-12.2
Membranes	2,000	1,450	-27.5
Process equipment	16,050	13,650	-15.0
Product water treatment	400	400	0
Site development	670	640	-4.5
Total investment	25,060	21,360	-14.8
Specific investment, \$/m3-day	1,104	940	-14.8

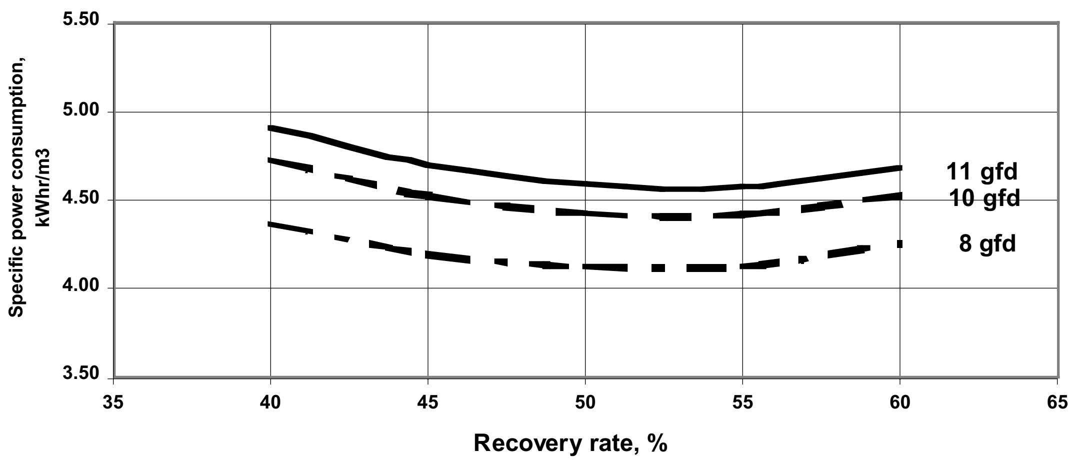
The combined effect of higher recovery and higher flux rate results in significant reduction of the investment cost components. These savings are transferred to the total water cost. However, the higher recovery rate and higher permeate flux require higher a feed pressure. Therefore, specific power consumption will be higher. Figure 3 contains the values of feed pressure at the feed water temperature of 18 C vs. recovery rate at a permeate flux rate in the range of 8 to 11 gfd. The increase in feed pressure with recovery rate at a given permeate flux is due to the increase in the average feed salinity and osmotic pressure. For pressure calculations, a 20% flux decline has been assumed due to fouling and compaction. Figure 4 contains values of specific power consumption corresponding to feed pressure values from Figure 3. The power consumption was calculated for a system equipped with a high efficiency pump and energy recovery turbine. For the pump and energy recovery turbine, efficiencies of 83% were assumed. For electrical motor efficiency a 93% value was used. Such high efficiency equipment is currently commercially available (5), and was offered and provided for a seawater unit of 2.1 mgd (8000 m3/day) capacity, which commenced operation in 1997. As shown in Figure 4, for a given flux rate, the minimum specific power consumption corresponds to a recovery rate of about 50%. Changing the design parameters from a 45% recovery rate and 8 gfd flux to a 55%

Figure 3. Projected feed pressure for Mediterranean feed, 18 C



recovery rate and 11 gfd results in increase of specific power consumption of the high pressure pump from 4.2 kWhr/m3 to 4.6 kWhr/m3. This difference of 0.4 kWhr/m3 at 18

Fig 4. Projected power consumption for Mediterranean feed, 18 C



C will decrease slightly at a higher feed water temperature. At a feed water temperature of 28 C (the high temperature limit for this evaluation) this difference will be about 0.3 kWhr/m3. Table 2 contains results of comparison of water cost components for operation at standard conditions with projected costs for operation at high flux, high recovery rate. For calculations of water cost the following economic parameters were used: interest rate: 8%, plant life: 20 years, replacement cost of 8 seawater membrane element: $800; maintenance: 3% of capital; power cost: $0.06/kWhr; plant load factor 95%. The cost of chemicals is based on the use of chlorine, inorganic flocculant, polymer, sodium bisulfite, and scale inhibitor in the pretreatment system. Operation at the high recovery rate and increased permeate flux improves process economics. However, maintaining stable membrane performance at the higher flux rate requires a significant improvement in feed water quality compared to the quality obtained from conventional pretreatment applied to

Table 2. Effect of recovery rate and permeate flux on total water cost in RO seawater system (in US$).

Water cost components	Reference design: 45% recovery, 8 gfd flux	HRF design: 55% recovery, 11 gfd flux	Water cost change, %
Capital cost	0.320	0.275	-4.0
Membrane replacement	0.050	0.037	-26.0
Maintenance	0.095	0.084	-12.0
Power consumption	0.252	0.276	+9.5
Chemicals & cartridge filters	0.060	0.050	-17.0
Labor	0.050	0.040	-20.0
Total water cost, $/m3	0.827	0.762	-7.9

surface seawater. The combined effect of lower investment and improved operating conditions is about 8% decrease of total water cost.

The conventional pretreatment process.

The objective of the pretreatment process is to reduce the concentration of fouling constituents in the feed water to the level that would provide stable, long term performance of membrane elements. In RO seawater systems membrane fouling can develop due to the presence in feed water of colloidal and particulate matter, dissolved organics, and as a result of biological growth in the RO system. The formation of inorganic scale, sometimes encountered in the tail position elements in brackish systems does not present a problem with the majority of seawater feeds. Precipitation of sparingly soluble salts from RO seawater concentrate is less likely to occur due to the relatively low recovery rate, high ionic strength, and low concentration of bicarbonate ion. Membrane fouling due to the first three fouling factors: e.i. particulate matter, organics and biogrowth, usually is most pronounced in the lead elements. It has been postulated based on results of recent research work on membrane fouling processes (6, 7) that colloidal fouling is accelerated by high ionic strength and permeate drag forces (permeate flux). The high ionic strength of a treated solution reduces mutual double layer repulsion between colloidal particles and between the particles and membrane surface. Permeate flux, which results in drag forces perpendicular to the membrane surface, forces colloidal particles and organic macromolecules into the membrane pores and the foulant layer on the membrane surface. In seawater applications, the combination of high ionic strength and the presence of colloidal particles and dissolved organics results in higher fouling rates at a lower permeate flux than encountered in the brackish applications. The fouling process affect membrane performance by affecting both water flux and salt passage. Under high fouling conditions permeate flux is reduced and a higher feed pressure is required to produce the design flow. Usually, there is a parallel increase in salt passage, resulting in a higher salinity of permeate. Advanced stages of particulate or biological fouling will result in blockage of feed channels and increased pressure drop.

The effectiveness of the operation of the pretreatment system and quality of feed water is measured in terms of the Silt Density Index (SDI). The SDI is a measure of filterability of RO feed water through a membrane filter of a defined porosity. Usually a filter with a nominal pore size of 0.45 micron is used. The value of SDI gives only a relative indication of feed water quality. There is no well defined relation between the SDI values of the feed water, the membrane fouling rate, or long-term membrane performance. The majority of membrane manufacturers specify an upper limit of SDI as 4 to 5. However, for stable membrane performance the average value of SDI should be below 3.

The conventional pretreatment system for surface seawater feed can be quite extensive. It may consist of breakpoint chlorination, up to a residual of 05 - 1.0 ppm,
followed by in-line coagulation and flocculation. Aggregated colloidal particles are removed in two stage pressure multimedia filters. The two stage filtration, roughing filtration followed by polishing filters, is an effective configuration for producing a more consistent feed quality during backwash steps and also during periods of seasonal deterioration of raw water quality. After media filters, and ahead of the cartridge filtration unit, scale inhibitor and sodium bisulfite is added to the feed water. The above configuration of the pretreatment system is effective in reducing SDI of the raw water from an unmeasurable values (complete plugging of the test filter) to a range of 2 - 3 SDI unit for the feed water. Regardless of the actual configuration of the pretreatment system, the above range of feed water SDI is very common for most of seawater plants receiving seawater from an open intake (8, 9, 10). However, the conventional pretreatment, consisting of media and cartridge filtration, does not represent a definite barrier to colloids and suspended particles. The quality of feed water produced, fluctuates significantly in respect of particulate matter. During filter backwashing, the filtration rates of filters remaining in operation increases, thus increasing the possibility of break through. Also, after the backwash, during the period of formation of filtration cake, large concentrations of colloidal particles are carried over with the filtration effluent. Apparently for these reasons the design average flux rate in seawater systems operating on surface water, has to be limited to about 8 gfd range.

The membrane pretreatment process.

The use of membranes as a definite barrier in the pretreatment process has been proposed in the past (10, 11). Ultrafiltration (UF) and microfiltration (MF) membranes have the ability to produce feed water of significantly better quality than the conventional pretreatment process based on media and cartridge filtration. However, the conventional, spiral wound configuration of ultrafiltration membrane elements was not suitable for the treatment of highly fouling surface water. UF elements could not operate at high flux rates without severe fouling of membrane surfaces and plugging of feed channels. High cross flow feed velocities, required to reduce concentration polarization, resulted in high power consumption. Membrane cleaning, frequently required, was cumbersome and not very effective in restoring permeate flux. New microfitration and ultrafiltration technology offered recently is based on a large ID capillary membrane configuration. The capillary bore is 0.7 - 0.9 mm diameter. Membrane material consists of polypropylene, sulfonated polyether sulfone or cellulose acetate. In some elements design the feed - permeate flow is outside in, others has inside out direction.

There are two common novel properties of the new commercial capillary equipment;

1. A frequent, short duration, automatically sequenced flushing (or backflushing in some models) of the capillary fibers, which enables to maintain stable permeate flux rates with little off-line time.
2. The ability to operate at a very low cross flow velocity, or even in a direct flow (dead end) mode.

The off-line time due to pulse cleaning is very short, comparable the to off line time of conventional filters due to filter backwashing. The frequent pulse cleaning results in a stable permeate flux rates. The feed pressure is in the range of 1 to 2 bar. Operation at low feed pressure and low cross flow or in a direct filtration mode results in high recovery rates and very low power consumption, of about 0.1 kWhr/m3 of filtrate. The membrane type is either microfiltration (nominal pore size 0.2 micron) or ultrafltration (molecular weight cut off 100,000 - 200,000 Dalton). The dimensions of capillary ultrafiltration modules are in the range of 100 - 130 cm long and 20 - 32 cm in diameter. In actual field operation, a single module can produce 30 - 150 m3/day of filtrate. This new capillary technology has been developed to treat potable water, which originates from surface sources. Compared to conventional technology, it offers modular design, high capacity from a small foot print, no need for continuous handling and dosing of chemicals, and limited labor requirements. The major advantage, however, is inherent to membrane technology: the existence of a membrane barrier between feed and permeate which enables a several log reduction of colloidal particles and pathogens.

This technology has been extensively tested and a large number of systems, mainly based on microfiltration membranes, are already in operations. Following the success in potable water applications, the capillary technology has been proposed and tested as a potential pretreatment for RO systems operating on highly fouling water. Application of the capillary technology as a pretreatment to seawater systems seems to be very advantageous as well. A limited number of tests on the operation of capillary units with seawater indicate that very good quality RO feed can be produced consistently (13). The cost of the capillary membrane pretreatment is estimated to be similar to cost of the extensive conventional pretreatment which is usually required for the surface seawater. Table 3 summarizes operating and economic parameters of the conventional and capillary membrane pretreatment for seawater RO systems. The use of capillary technology will simplify the pretreatment system and reduce the use of chemicals. It will eliminate the need of a continuous presence of free chlorine, and produce feed water with very a low concentration of colloidal particles. There no sufficient data to decide which type of capillary technology; microfiltration or ultrafiltration, will be more effective in reducing foulants concentrations in RO seawater applications. In addition to particulates, the ultrafiltration membranes will also retain to some extent dissolved organics (TOC) present in the seawater. Future field results should show to what extent this property is important in seawater applications. Membrane pretreatment will go in parallel to the recent trend of increasing packing density of spiral wound elements i.e. increase of membrane area and permeate capacity per element. This is achieved by optimization of the element configuration, which also include decrease of the height of feed channels. The potential for blockage of the thinner feed channels in the new spiral element design will definitely be smaller in operation with high quality feed water containing very low concentrations of particulate matter. Moreover, reduction of the fouling rate will reduce the frequency of chemical cleaning, and should result in a significant increase in the useful membrane life. This in turn will reduce cost of chemicals, cleaning associated labor and membrane replacement cost.

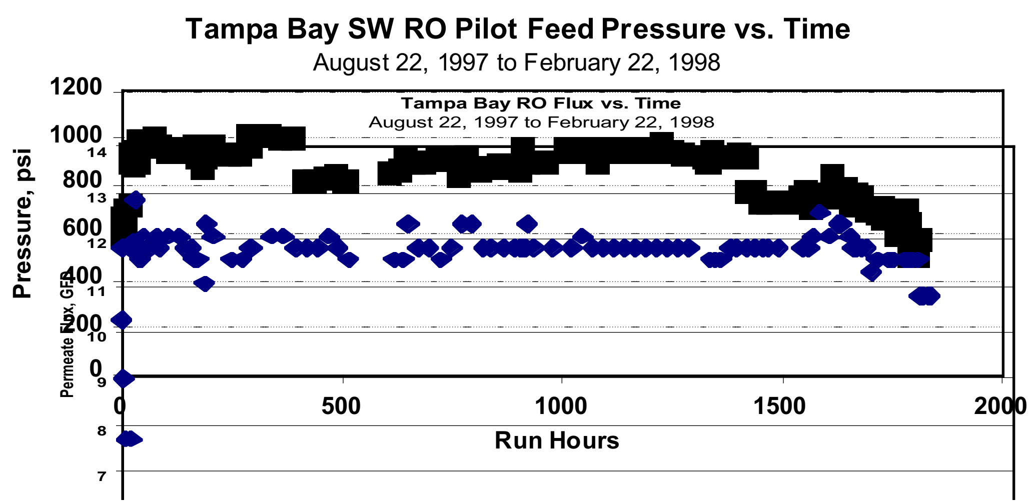
Table 3. Comparison of conventional pretreatment and capillary membrane technology.

System configuration	Conventional pretreatment	Advanced pretreatment
Feed water source	Open intake	Open intake
Strainer	Coarse screen	Self cleaning microscreen filter: 35 micron
Raw water chlorination	3 ppm	Not required
In line coagulation, continuos dosing	$FeCl_3$ 5 ppm Polymer 0.2 ppm	Not required
Static mixer	Yes	No
Colloids filtration equipment	Two stage multimedia pressure filters	Single stage capillary membrane ultrafiltration: 150,000 - 200,000 Dalton cut-off
Backwashing mode & frequency	Air scour and water backwash every eight hours. Off line time 15 min (3.1%).	Water pulse backwash every 15 min. Off line time 30 sec. (3.3%)
Chemical cleaning	Not required	Every 4 hr soak with 50 ppm NaOCl solution for 10 min. Off line time 0.4%

Filtration rate, l/m^2hr (gfd)	First stage filters: 6 (3.5) Second stage filters: 10 (6)	100 (60)
Maximum driving pressure, bar	First stage filters: 0.5 Second stage filters: 0.8	2 bar
Power consumption, kWhr/m3 effluent	0.07	0.10
Backwash water losses, %	4 (2.5% + 1.5%)	5
Cartridge filters rating, micron	5 - 15	Not required
Effluent quality, SDI	2 - 3	< 1
Dechlorination, NaHSO3 ppm	3	Not required
Pretreatment investment cost, $\$/m^3$-day permeate	100 - 250	150 - 300

Operation of hybrid UF/RO seawater pilot unit at Tampa FL.

The UF unit commenced operation in August 1997. The UF unit operates on surface seawater from Tampa Bay. The Pilot unit consists of one 8 capillary element. The UF unit operates in dead end and partial cross flow filtration mode. The UF unit serves as a pretreatment for the feed to the RO unit downstream. The utilization of capillary UF as a pretreatment step enables operation of RO membranes at high recovery and high permeate flux rate. The RO unit operate at recovery rate of 65% and flux rate of 12 gfd . Figures 5 and 6 include results of permeate flux and feed pressure during pilot operations. The results are indicative of stable membrane performance at a significantly higher values of permeate flux and recovery rate then would be customary to operate RO seawater unit using conventional pretreatment .



Summary.

An increase in recovery rate and permeate flux in seawater systems can improve the economics of the desalting process. Implementation of high recovery, high flux operation requires better quality of the feed water. New capillary membrane technology used as a pretreatment step has the potential to produce feed water quality which will enable to operate seawater membranes at a higher flux rate. The new technology has demonstrated reliable operation at variety of operating conditions. It is cost competitive with conventional pretreatment technology, and will result in higher reliability and better overall economics of reverse osmosis seawater desalting. The combined savings due to lower investment and operating cost and ability to optimize system operating conditions due to better feed water quality should result in about a 10% reduction in total water cost.

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