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THE DESALTING AND WATER TREATMENT MEMBRANE MANUAL: A GUIDE TO MEMBRANES FOR MUNICIPAL WATER TREATMENT (2ND EDITION)

Water Treatment Technology Program Report No. 29

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U.S. DEPARTMENT OF THE INTERIOR
Bureau of Reclamation
Technical Service Center
Water Treatment Engineering and Research Group

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The Desalting and Water Treatment Membrane Manual: A Guide to Membranes for Municipal Water Treatment (2nd Edition)

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UNITED STATES DEPARTMENT OF INTERIOR ★ **BUREAU OF RECLAMATION**

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Disclaimer

SI Metric Conversions

From	To	Multiply by
ft	m	‘0.3048
in	m	‘0.0254
ft^2	m^2	**‘0.09290304**
kgal (U.S.)	m^3	3.785 412
Mgal (U.S.)	m^3	3,785.412
acre-ft	m^3	1,233.489
lb/in^2 (psi)	kPa	6.894 757
°F	°C	[1] °C=(° F-32)/1.8

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[1] Exact conversion.

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Chapter 1
INTRODUCTION

1.1 Recent Developments in the Membrane Industry

Since the first Desalting and Water Treatment Membrane Manual was written in 1992, there have been significant advancements in the development of membrane technology. Every aspect of the technology has changed except, of course, for the basic principles of membrane separations. There are now more manufacturers producing many more types of membrane out of different materials, in different configurations. The latest catalogs feature modules with greater membrane area, and modules that are 40 or 60 inches long. Many manufacturers now have a product line with a range of productivity and rejection characteristics. Manufacturers of ceramic and stainless steel membranes, that were priced for high dollar, low quantity separations in the early '90s, are now trying to lower their costs to attract large-scale applications. Microfiltration (MF) and ultrafiltration (UF) have also been reclaimed from the dairy and paint industries and have taken their places in water treatment plants all over the world.

1.1.1 Wider Application

While in the early 1990s most water treatment membrane applications were for brackish water treatment, there are now more sites where membranes are being used or considered for removing specific contaminants such as manganese, radium, fluoride, nitrate, or color. This change in applications is a sign that communities are beginning to think more about their "bang for the buck." If one must spend $2 million to remove 0.5 mg/L of manganese with a greensand filter and still end up with poor tasting water, or $3 million to produce good tasting water, which process should be endorsed? Communities striving to attract development should be considering the latter.

Table 1 .1 lists several undesirable water contaminants, the conventional solutions for them, and corresponding membrane processes that can do the job. There are many variations on these conventional processes that could be included, but the ones listed are sufficient to illustrate that there is a membrane process alternative available to address most drinking water problems.

1.1.2 Systems Integration

At the 1997 American Water Works Association (AWWA) Membrane Technology Conference it seemed the question of conventional verses membrane treatment had been settled. Presentations there were concerned with which membrane process to use with which membrane, under what conditions, and in what order. There was a session on Integrated

Table 1 .1 .-Conventional and membrane process solutions to common water problems

Constituent of concern	Conventional process	Membrane process
Turbidity Suspended solids Biological contamination	Coagulation/flocculation Media filtration Disinfection	Microfiltration
Color Odor Volatile organics	Activated carbon Cl, + media filtration aeration	Ultrafiltration
Hardness Sulfates Manganese Iron Heavy metals	Lime softening ion exchange Oxidation, filtration Ion exchange Coagulation/flocculation	Nanofiltration
Total dissolved solids Nitrate	Distillation Ion exchange	Reverse osmosis Electrodialysis

Membrane Systems with studies on whether MF or UF should serve as pretreatment for reverse osmosis (RO) or nanofiltration (NF) for treating surface water. Membrane optimization was discussed in several of the sessions. Should the productivity of each stage be balanced or not? Does it make a difference if recovery is increased by adding membrane area, recirculating concentrate, or by lowering feed flow?

1.13 Selection Criteria

There were quite a few papers on membrane selection studies. Most compared all the membranes at one permeation rate. Is this valid? In one of our screening tests, we tested each membrane according to its manufacturers suggested operating pressure and feed flow (Boegli, Chapman Wilbert, 1997). How should membranes from different manufacturers be compared in a screening test? Should they be compared at the same pressure or the same permeation rate? Or should the optimum performance for each membrane be found through experimentation and then compared on that basis?

1.1.4 Module Integrity

Module integrity is another area of present concern. Communities would like to use their membrane systems to ensure protection against viral and bacterial contamination. If a membrane rejects salt, one would hope that it would also exclude viruses and bacteria, but does it? Test data are discouraging on that point. During a pilot study in Avondale, Arizona, bacteria were found in NF permeate when there was a high bacteria count in the feed water (Jurenka, Chapman Wilbert, 1996). Another study of virus removal efficiency of MF, UF, and RO found that RO could effect a 3-4.8 log removal while UF effected greater than 5.4 log removal (Kruithof, et al., 1997).

1.2 Basics of Membrane Technology

1.2.1 What is a Membrane?

A membrane is a film. A semi-permeable membrane is a very thin film that allows some types of matter to pass through while leaving others behind. Some membranes are porous and separate materials based on size compared to the size of the pores. Others are dense films with no apparent pores that separate matter based on differences in diffusion rates through the membrane.

Membranes are divided into indistinct classes based on the size of the materials they retain. Table 1.2 gives an overview of the various types. MF membrane is very porous; it allows water, dissolved salts, colloidal materials, and particles that are smaller than the pores to pass through. On the other end of the spectrum, RO membrane is a dense film with no pores, only spaces in its polymeric structure that are large enough to allow water and other small, uncharged molecules to pass through.

Table 1.2.—Characteristics of the major types of water treatment membranes

	ED	MF	UF	NF	RO
Materials retained	Water, micro-organisms, uncharged molecules, suspended solids	Particles larger than pore size	Molecules larger than the molecular mass cutoff	>95% of multi-valent ions, 25-90% of mono-valent ions, molecules and particles over 300 Daltons	>95% of all ions, most molecules and particles over 200
Materials transported	Dissolved salts	Dissolved salts, small particles	Small molecules and ions	Mono-valent ions and very small molecules.	Very small uncharged molecules
Water[1] permeation $(m^3m^{-2}d^{-1})$	Practically none	[2]800 - 4,000	0.4 - 2.5	1.0	0.8

[1] Pure water at normal operating pressure. Actual values are extremely condition dependent.

[2] Depending on micron rating.

There are three primary mechanisms for the separation and transport of water and solutes across a membrane: sieving, convection, and solution diffusion. Sieving in membranes is the same as sifting sand and stones through a wire mesh. Molecules or particles that are larger than the membrane pores are retained within the membrane if it is a cartridge type filter or in the concentrate stream if it is a cross flow design. Convection is the mechanism whereby solute is carried through the membrane with the solvent. Diffusion is the transport of solute through the membrane once it has dissolved into the membrane polymer. The concentration gradient across the membrane drives the diffusion of solute to the low concentration side (Allgeier and Summers, 1995). Figure 1.1 illustrates the differences in these mechanisms and their controlling factors.

1.2.2 Sieving

The pore size distribution of the membrane and the particle shape and size distribution of the solution are the important factors in the sieving mechanism of solute rejection. See the chapter on MF for a more detailed discussion.

1.2.3 The Solution Diffusion Model (briefly)

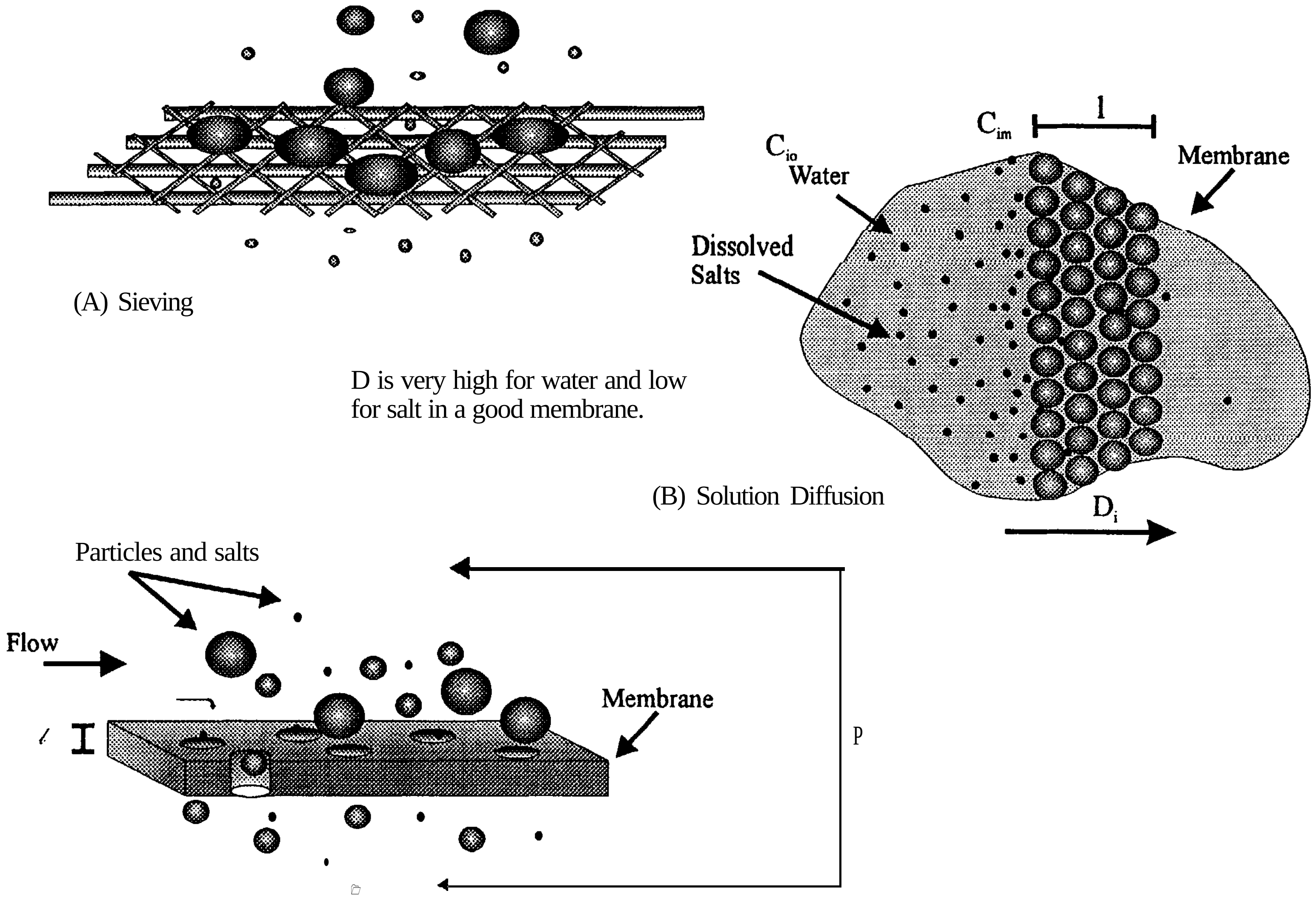
There are three important aspects of a membrane that control the solution diffusion process and, thus, the membranes performance:

1. The solubility and diffusion rate of the solvent in the membrane (in this case, water),
2. The solubility and diffusion rate of the solute in the membrane (in this case, salts or other impurities of water), and
3. The thickness of the membrane

In the solution-diffusion model, material dissolves in the membrane and then diffuses to the other side down a concentration gradient (see Wijmans, Baker, 1995 for more detailed discussion and further references). In this way, the permeation of a component "*i*"of the solution is described:

$$J_i = \frac{D_i(c_{io(m)} - c_{il(m)})}{l} \qquad \text{Eq. 1.1}$$

Where D_i is the diffusion coefficient made up of the solubility of component "*i*" in the membrane and the diffusivity of "*i*" through the membrane. The term in parentheses is the difference in concentration across the membrane, and l is the thickness of the membrane. To have a high water permeation rate, the membrane should be able to absorb water, move it through quickly, and be very thin. Yet, the membrane must also have a low salt permeation rate. Since the water is pouring through our "perfect" membrane, the salt concentration



(A) Sieving

D is very high for water and low for salt in a good membrane.

(B) Solution Diffusion

(C) Pore Flow Model.

Figure 1.1.-Models for various types of membrane separation.

gradient and thickness are helping to enhance salt permeation. The diffusion coefficient must come to the rescue. If our membrane has a minuscule salt diffusion rate, we will still have a perfect membrane.

1.2.4 Pore Flow Model (even more briefly)

Permeation through membranes that do have pores is described by the pore flow model which looks similar to the solution diffusion model but is completely different.

$$J_i = \frac{k\ \Delta p}{l} \qquad \text{Eq. 1.2}$$

Where J, is the permeation of component "*i*," k is the Darcy's Law coefficient that has to do with the number of pores large enough to accommodate "*i*," Ap is the difference in pressure across the membrane, and "*l*" is the membrane thickness.

1.2.5 Basic System Design

Each membrane process has three process streams: the feed stream, that may have undergone some type of pretreatment; the product stream; and a concentrate stream. Using the term "stream" is convenient, but in some applications the concentrate is actually a solid, such as in some dairy applications and in sludge de-watering. Figure 1.2 illustrates the general design of a membrane system. The feed stream is exposed to the semi-permeable membrane which, under the influence of the appropriate driving force, separates the product from the concentrate.

1.2.6 Membrane Configurations

Membranes can be made from many different materials and can be formed into a wide variety of configurations. The most common are depth filters, plate and frame, spiral wound, pleated, tubular, and hollow fiber. Figures 1.3a and 1.3b show the differences between the various configurations. No configuration is better under all circumstances; there are advantages and disadvantages inherent in each of them. One must select the best for a particular situation.

- Depth filters are used in MF to trap coarser particles. They are dense, thick walled cylinders made from spun polymer strands. They are formed such that the outside has a more open structure than the inside so that as water passes from the outside in, larger particles are trapped first.

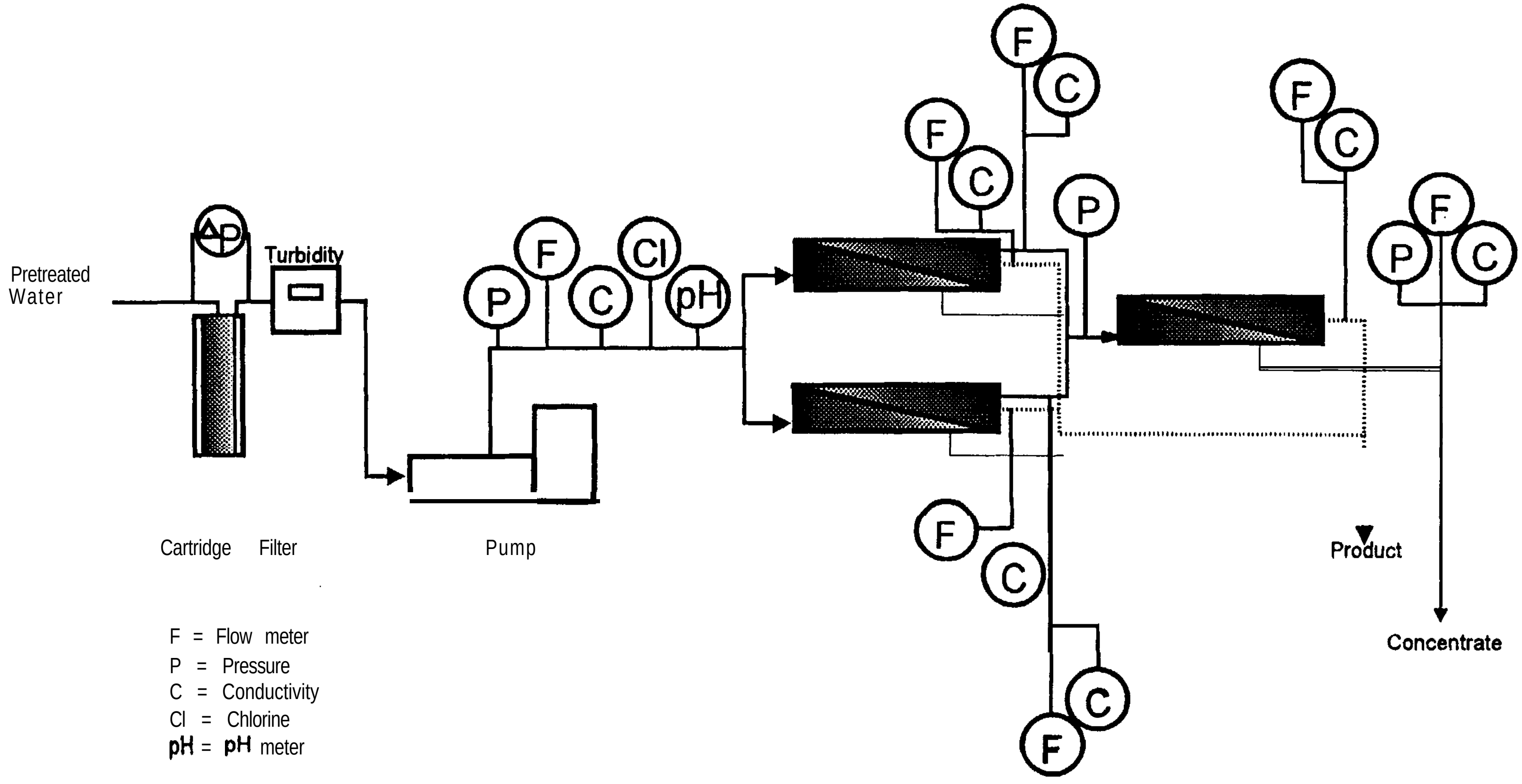


Figure 1.2.—General design of a membrane system.

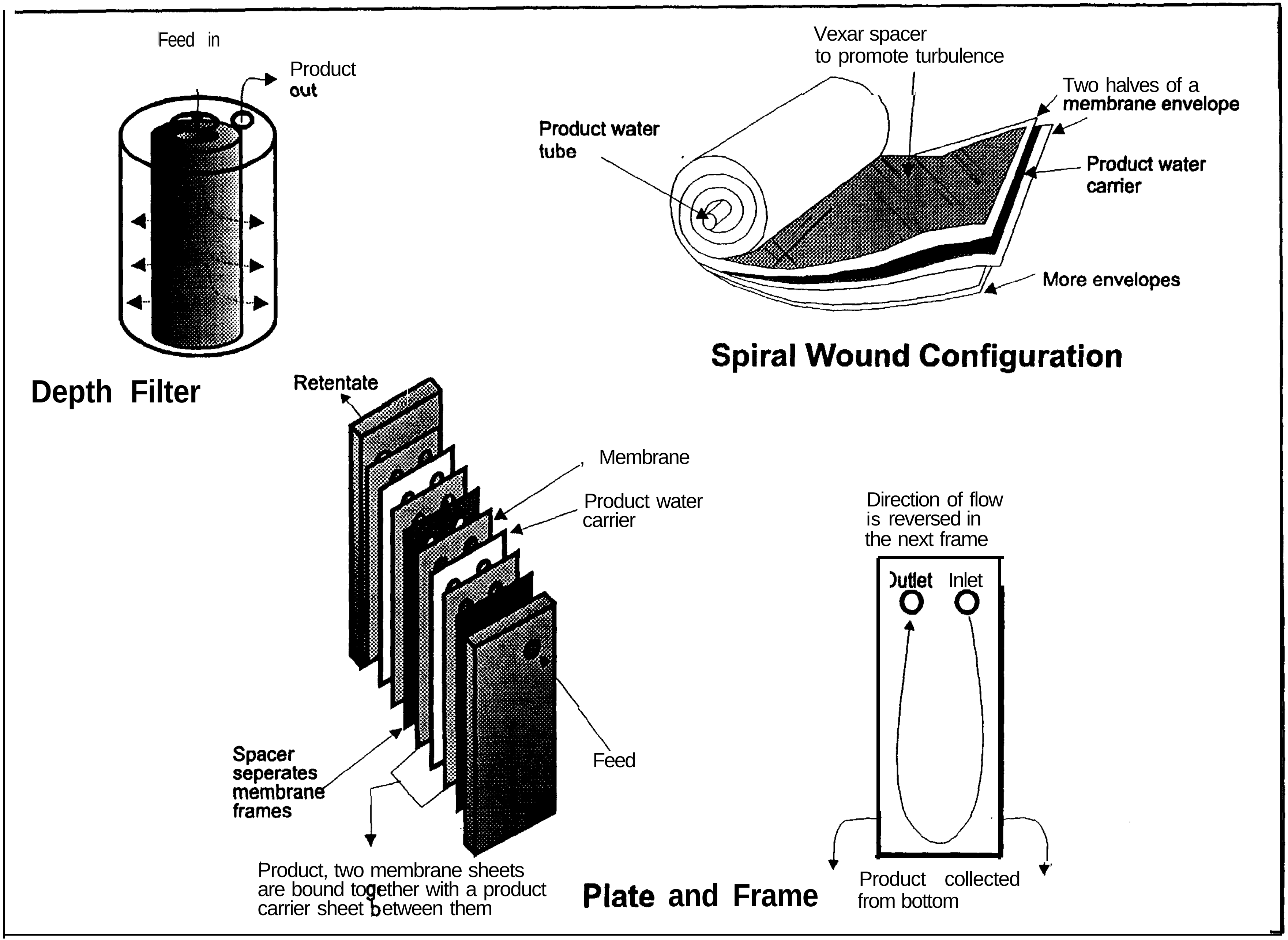
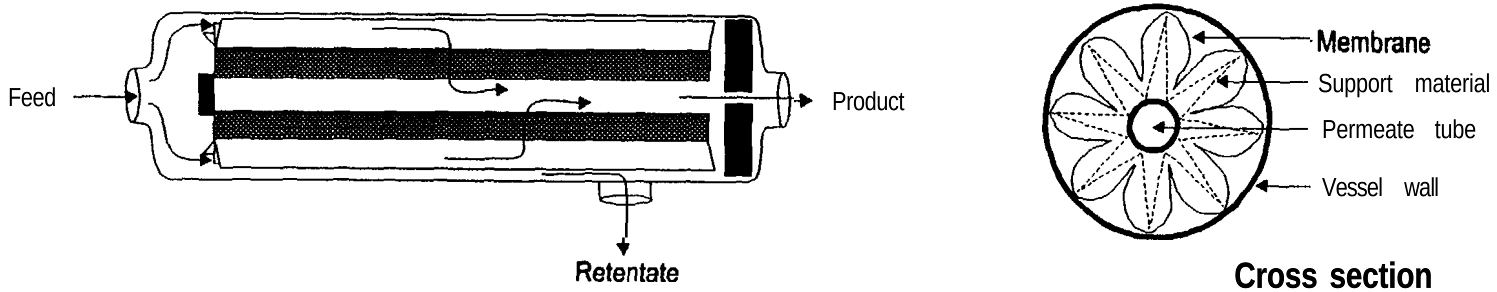
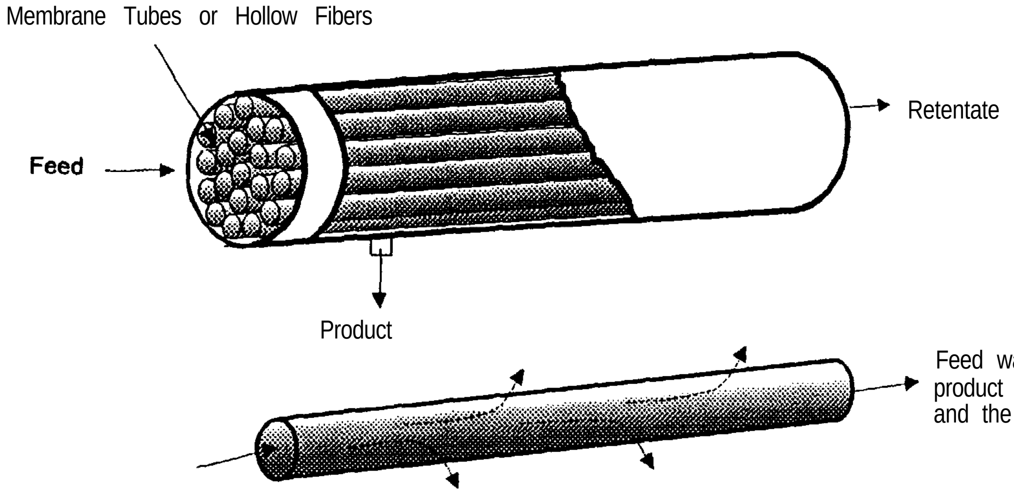


Figure 1.3a.—General membrane configurations.



Pleated Membrane Cartridge



Feed water flows through the inside of the tube, product water permeates through the membrane, and the retentate exits through the other end

Tubular or Hollow Fiber Module

Figure 1.3b.—Generalized membrane configurations.

- Plate and frame configurations are used in electrodialysis and in high solids content micro- and ultra-filtration because the units can be taken apart and cleaned very thoroughly.

- The spiral wound configuration, which is used in the whole range of filtration processes, provides a greater surface area per module than pleated or tubular configurations in the same processes, but is much more difficult to clean.

- Pleated membrane cartridges are a cross between a solid depth filter and a spiral wound membrane. The thinner, pleated, membrane material allows higher flux rates than a depth filter. Because the membrane is supported on both sides, back-flushing can be used to help extend membrane life.

- The hollow fiber configuration, used in RO and NF, has the highest surface area to volume ratio attainable, but requires a higher level of pretreatment than the other configurations.

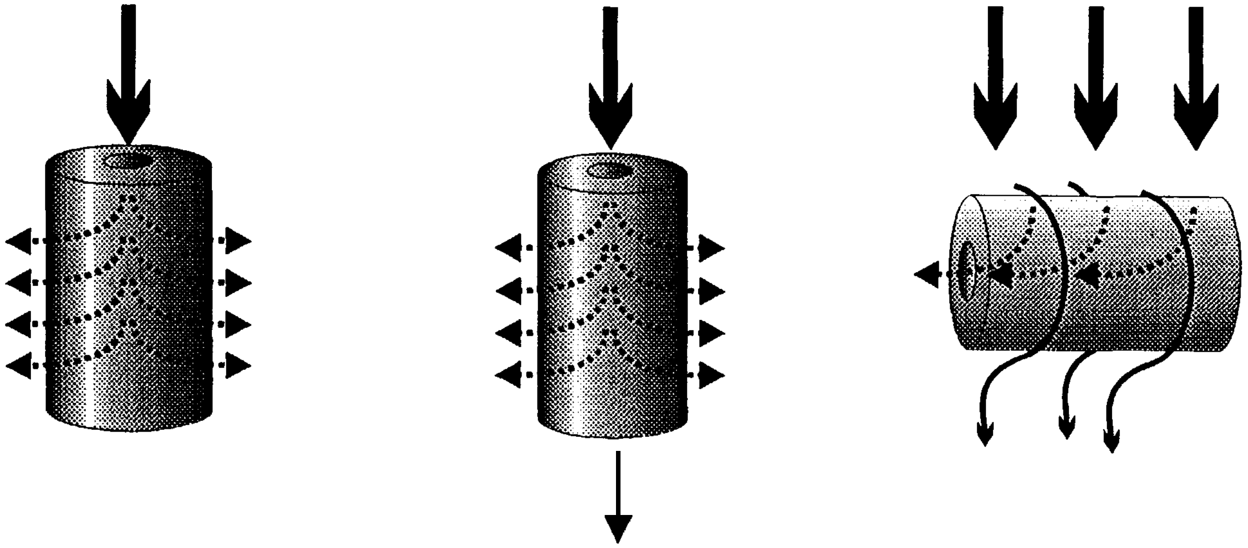
- Tubular configurations can be used for MF, UF, and sometimes NF membranes. They are easy to clean but have a low surface area to volume ratio. Tubular configurations are used for high flow/high solids content feed streams.

- A more recent configuration, called transverse-flow tubular, uses small diameter tubular membrane with the tube ends potted into the sides of the cylindrical vessels to form a pattern like the webbing of a tennis racket. The feed water flows at a right angle, or transverse, to the tubes. Product water permeates to the inside of the tubes and is collected outside the vessel. The advantage of this configuration is that the tubes are flexible enough to vibrate as the water flows over them. The motion and flow pattern helps to wash foulants from the membrane surface.

- Another variation on the tubular membrane configuration has the membrane strands submerged in the source water. Rather than pumping water into the filter, the filtrate is pumped out of the source from the interior of the tubular membranes under a vacuum. If you imagine sucking on a foam straw that is blocked off at the end, you will see just how this system works. Blasts of air can be used to vibrate the tubes periodically to prevent build up on the outside of the tubes.

1.2.7 Dead-End, Cross flow, and Transverse Flow Operation

Dead-end, cross flow, and transverse flow operation refer to the direction of flow across the membrane. Figure 1.4 diagrams the differences between them. In dead-end filtration, there is no liquid waste stream; all feed water passes through the membrane. Over time, particles build up in the membrane structure or on the surface, limiting water passage. As the filter cake accumulates, the AP builds until the maximum is reached, and then the membrane must be replaced or cleaned.



Dead end flow: All water passes through the membrane, particles are trapped within the membrane structure.

Cross flow: Feed water passes parallel to the membrane surface, product water permeates through the membrane, the concentrated stream helps carry particles out of the system.

Transverse flow: Feed water meets the membrane surface at right angles, product water permeates through to the inside of the tube, concentrate washes over the outside of the tube removing particle build-up.

Figure **1.4.—Dead-end,** cross flow, and transverse flow.

In cross flow operation, the feed stream flows parallel to the membrane surface, limiting filter cake thickness and density. A part of the stream permeates through the membrane, leaving a zone of high particle concentration at the surface. Particles and/or solutes are drawn back into the bulk feed stream by the flow of lower concentration water past the surface.

Transverse flow operation is used with tubular membranes configured such that the feed stream flows past them at right angles. The product stream permeates to the interior of the tubes. The higher turbulence across the membrane in this configuration enhances filter cake disruption and thereby maintains a higher productivity rate than cross flow or dead-end filtration operated under equivalent conditions.

1.3 About This Book

Answers to the questions raised in section 1.2 need to be found in the next few years if membrane technology is to reach its full potential. They will be explored more fully in the following chapters on each form of membrane filtration. A brief review of each process is presented with a discussion of their similarities and differences, main operational parameters, and most important aspects to consider in choosing a membrane for your particular application. Examples of novel uses and methods of process optimization are discussed. Since specific cleaning procedures are required to satisfy warranty conditions for each manufacturer, the chapter on cleaning will provide a basic discussion of cleaning philosophy rather than a reproduction of cleaning recommendations. The process design section describes simple methods for designing membrane systems, determining membrane area, and predicting performance. Cost and concentrate disposal options are covered in chapters 8 and 9. The product listing has been updated with performance statistics calculated from the manufacturers test data in chapter 10. An analysis is included to help determine the functional differences between the new module configurations.

Chapter 2
MICROFILTRATION

Microfiltration as a process has received considerable attention in the United States in recent years. Much of this has been under the impetus of legislation which has increased the requirements on drinking water. In addition, improved methods of removing the fouling layer from microfiltration membranes have considerably improved the economics of the process. Highly varied equipment is available commercially. Existing microfiltration plants are now providing operating experience. And a number of demonstration or pilot plants are being used to provide comparison to alternative processes and to generate design data. Research on a number of different applications relating to drinking water, including pretreatment for desalting plants and watershed protection, is being carried out.

2.1 Definition

When we speak of microfiltration, we are talking about filtration through a coherent medium with a nominal pore size range from slightly below 0.1 µm to slightly above 1 µm. This size range refers to the pore size of the medium itself, not to that of any cake or fouling layer which may accumulate on the medium. Thus, in terms of pore size, microfiltration fills in the gap between ultrafiltration and granular media filtration. In terms of characteristic particle size, this range covers the lower portion of the conventional clays and the upper half of the range for humic acids. This is smaller than the size range for bacteria, algae and cysts, and larger than that of viruses. Sizes of various materials of concern are shown in figure 2.1.

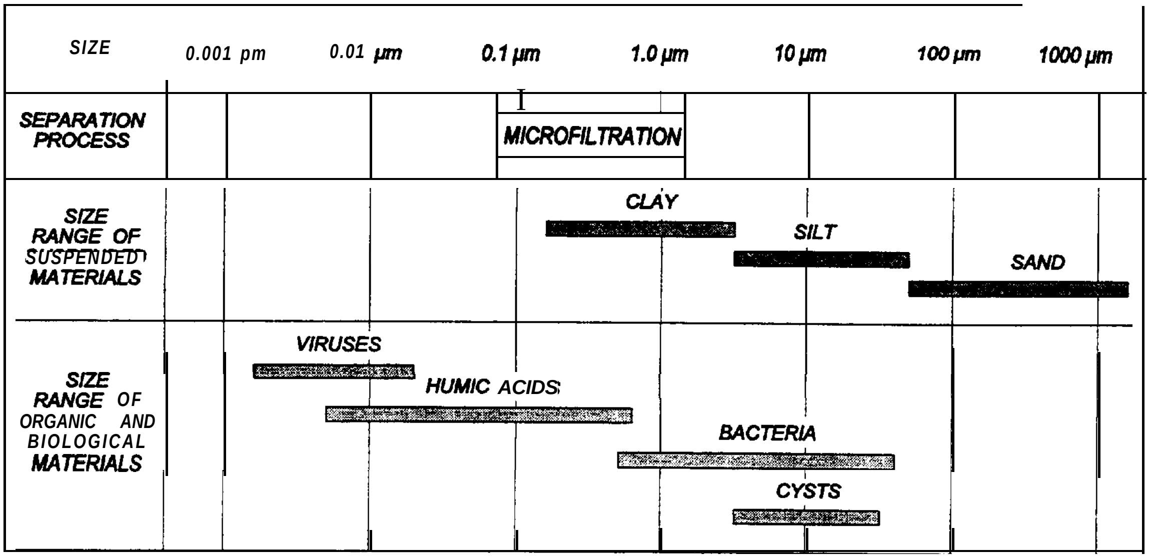


Figure 2.1 .-Comparison of size range of various materials.

Over the years, a number of industrial applications have developed for microfiltration. However, this paper covers only drinking water applications. The field of drinking water supplies both a blessing and a curse to any treatment process. The blessing is that the potential market is both volumetrically huge and worldwide. The curse is that the process must be inexpensive.

2.2 Background

One of the principal reasons behind the considerable interest in microfiltration in the United States has been recent regulations concerning drinking water. The 1986 Safe Drinking Water Act Amendments required new drinking water regulations and treatment requirements. The Surface Water Treatment Rule requires that disinfection or a combination of filtration and disinfection be employed to achieve 99.99 percent (4 log) removal of viruses and 99.9 percent (3 log) removal or deactivation of *Giardia* cysts. Turbidity standards require that 95 percent of the water samples collected each month have a turbidity of less than 0.5 NTU (Nephelometric Turbidity Units) and that filtered water turbidity never exceed 5.0 NTU. In addition, the Disinfection Byproduct Rule targets trihalomethanes and haloacetic acids.

Interestingly, in the EPA (Environmental Protection Agency) manuals on Microorganism Removal for Small Water Systems (SME-Martin Inc. Microorganism..., 1983) and Turbidity Removal for Small Public Water Systems (SME-Martin Inc. Turbidity..., 1983) there is *no* mention of membrane filtration. The first manual mentions disinfection with chlorine, chloramines, chlorine dioxide, ozone, and ultraviolet radiation along with the non-treatment alternatives of watershed management, source substitution, and regionalization of supply. The second refers to conventional flocculation, sedimentation, and filtration.

2.3 Microfiltration Membranes

Membranes for microfiltration, as defined above, have been available and in use for decades. The 0.45 micron membrane used for silt density index measurements are microfiltration membranes. Nucleopore membranes have been available since the early 1960s (Silk and Barnes, 1955; Fleischer, *et al.,* 1963; Fleischer, *et al.,* 1969). However, only in recent years has there been the kind of commercial interest that we now see. In order for microfiltration to become a commercial process, we need membrane apparatus with three characteristics:

1) Produce water at a reasonable cost,

2) Be essentially free of leaks and have good seals around pressurized units, and

3) Be capable of being cleaned.

2.4 Microfiltration Systems

There are four major suppliers of microfiltration equipment. Characteristics of equipment are shown in tables 2.1 and 2.2.

Table 2.1 .-Characteristics of commercial microfiltration eauioment

	Memtec	Dow	U.S. filter	Xenon
Pore size	0.2 µm	0.1 to 0.2 µm	0.8µm	0.2µm
Material	Polypropylene	PVDF	Ceramic	Polymeric
Configuration	Hollow fiber	Flat sheet	Multichannel	Hollow fiber
Flux	67 GFD	30-50 GFD	400 GFD	50-63 GFD
Flow pattern	Dead end	Crossflow	Crossflow	Crossflow
Trans-membrane pressure	3-15psi	2 to 15 psi	Up to 20 psi	8 to 12 psi
Chlorine sensitivity	Can only tolerate a combined chlorine residual	3mgR free chlorine residual desired	Tolerates chlorine residual	Tolerates chlorine residual
Chemical addition	Sodium thiosulfate for dechlorination		Aluminum chlorohydrate for flocculation	
Cleaning	Water/air backwash every 15 minutes. Citric acid used for inorganic fouling.	Periodic backwashing with permeate. Every 1 to 3 weeks, clean with sodium hypochlorite or Biz	Recirculation of 2% caustic, chlorine and possibly sulfuric acid	Continuous air scour. Relaxation or permeate back-flush cleaning.

An interesting aspect of the Memcor equipment is that it permits an automatic test for membrane and seal integrity. Operation is periodically interrupted, and a pressure of air at 15 to 20 psi is applied. Too rapid a decline in pressure discloses the presence of a leak.

The problem with estimates like those above is that they are based on comparatively little data. As a result, there is considerable uncertainty in these numbers. Because of the recentness of this application, there are few data on the useful life of microfiltration membranes in this service.

An excellent review of the overall status of microfiltration is provided by the NWRI/OCWD (National Water Research Institute/Orange County Water District) symposium (1994) held in

Table 2.2.—Manufacturer's estimate of system costs for units about 1 million gallons per dav capacitv

	Memtec	Dow	U.S. filter
Capital, $/daily gallon	0.77	0.67	0.94
O&M, $/1,000 gallons	0.16	0.40	0.14
Energy, kWh/1,000 gallons	0.8	5.0	Not provided

Irvine, California. A recent paper presents results of a survey of microfiltration plants worldwide (Adham et *al.,* 1996). This study indicates that in 1996, based on the results of their survey, there were a total of 40 microfiltration plants worldwide, with a total capacity of 12 MGD (45,000 cubic meters/day). Over 90 percent of these plants were supplied by Memtec. For plants with a capacity of 0.1 MGD or more, the installed membrane system unit cost was between $0.50 and $2.00 per (gallon/day). The plant capital cost is about twice that. Based on about 290 responses, the operating and maintenance cost ranges from $0.20 to $0.25/thousand gallons.

Microfiltration can typically produce water of satisfactory turbidity with feed waters exceeding 100 NTU. *Giardia* removal as high as 6 logs is reported. Bacteria removal is satisfactorily *high. Cryptosporidia* is satisfactory. Virus removal of 0.5 log (68 percent). Disinfection byproduct removal is about 10 percent.

The major difficulty for the technology in California, where there is a large potential market, is the desire to treat wastewater and the necessity to meet the CDHS (California Department of Health Services) regulations. For instance, at Grange County Water Factory 21, they treat wastewater and inject it into a drinking water aquifer to prevent seawater intrusion. Their treatment must have multiple barriers that must remove at least 4 logs of microbiological species. They have found that microfiltration has been good at bacterial removal, but in order to get the virus and cryptosporidium removal, ultrafiltration is needed. Interestingly, cellulose acetate RO membranes have a better removal than thin film composite membranes, but ultrafiltration has been shown to give a better removal than RO. One hypothesis is that sealing the product stream from the feed stream is more of a problem in RO than UF, as the equipment is configured. Many studies have been done to determine the different membrane removal efficiencies.

2.5 Operating Plants

San Jose, California (Yoo *et al.,* 1994; Yoo *et al.,* 1995). Faced with a requirement to upgrade or abandon the existing Saratoga 5 MGD diatomaceous earth treatment plant, the San Jose Water Company installed a microfiltration plant of this capacity. The major alternatives considered were diatomaceous earth filtration, two-stage filtration, microfiltration, and ultrafiltration. These were looked at on the basis of cost, regulatory approval, operability, and environmental and design constraints. Cost estimates for these processes are shown in figure 2.2.

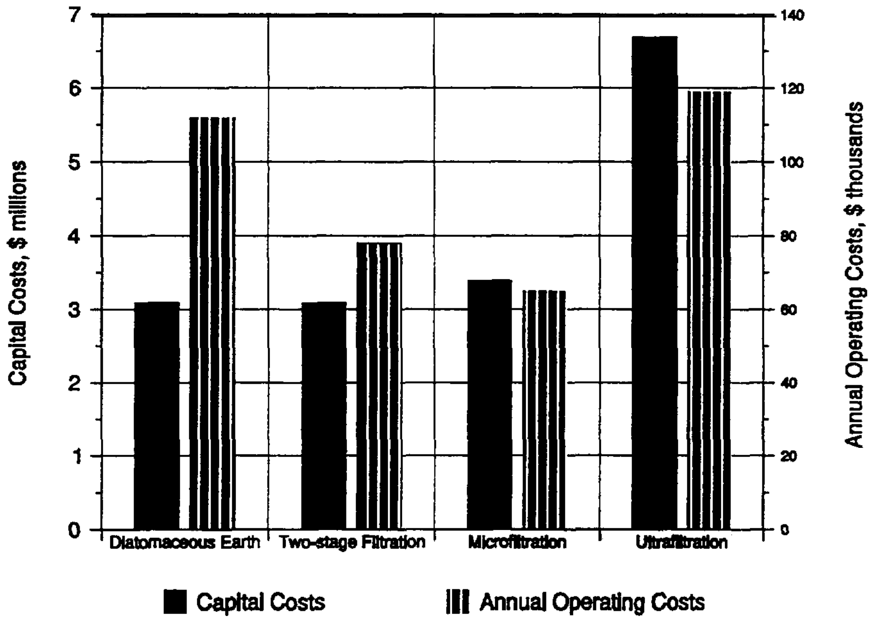


Figure 2.2.—Alternatives for the Saratoga Water Treatment Plant.

Microfiltration equipment was manufactured by Memtec Ltd. consisting of six factory assembled units, each composed of 90 modules, each of which contains 10 square meters of membrane surface. Capital cost of the microfiltration equipment alone was $1.5 million which equates to $0.30 per (gallon&y). Design flux is 86 GFD. It has successfully operated since February 1994. Gere (1997) reports that in its fourth year, the plant continues to operate reliably and to provide quality product water. Operating costs for the first year came to 30.9 cents per thousand gallons, which includes a prorated cost for membrane replacement based on six year membrane life. No membranes have been replaced yet. Periodically, the membranes require chemical cleaning. Cleaning interval varies from 2 to 6 weeks, depending on the quality of the feedwater.

Toms Brook, Virginia, has installed a plant with a capacity of 130,000 GPD (McClain, 1994). This treats part of the water being supplied to the communities of Toms Brook and Maurertown. The water source being treated by microfiltration is a new well, Toms Brook

Well No 2. The fairly porous karst soil allows groundwater from this well to be influenced by surface infiltration, so the Virginia Health Department indicated that they considered this a surface water in terms of treatment regulation.

Because microfiltration is a new water treatment method, a pilot study was required to demonstrate its efficacy. The pilot study, which was run with a 43,000 gallon per day unit, ran successfully for more than 2 months. During this test, the turbidity was artificially elevated to challenge the membranes. Parameters monitored during the test included turbidity, particle counts, coliform and membrane integrity. Removal of Giardia, Cryptosporidia, bacteria, and some removal of viruses are attributed to the microfiltration membrane. However, post-chlorination is the primary means of removal of viruses.

The final plant consists of three units identical to the one tested with the option to install two further units. It contains a 200 pm bag prefilter. After microfiltration, the water is chlorinated and sufficient contact time is allowed before distribution. The plant is sufficiently automated that it needs to be checked by the operator only at the beginning and end of a shift. Thus, additional personnel were not required by addition of the filtration unit.

2.6 Demonstration or Pilot Plants

Scottsdale Arizona. Scottsdale is currently testing microfiltration for aquifer recharge and for surface water treatment (Dueker and Hemken, 1994). They have a 720,000 gallon per day Memcor unit. They are also investigating microfiltration as a pretreatment for reverse osmosis units.

Orange County, California. Water Factory 21 tested a 2,800 GPD unit in 1992 (Sudak and Dunnivan, 1994). Pleased with the results, they subsequently tested a 43,000 GPD unit. Both units were from Memtec. Indications were that the MF would allow them to increase RO flux rates from 10.3 to between 12 and 13 GFD. They are now embarking on a demonstration project of 800,000 GPD.

Colorado River Aqueduct (Coffey et *al.*, 1994). The Metropolitan Water District performed a very thorough test from August 1991 to March 1992 on a Memcor pilot unit containing 4 square meters of membrane. Based on favorable results, they requested that California approve the microfiltration process for this application. The CDHS credited microfiltration *with 3* logs of *Giardia* and 0.5 log (68.4 percent) reduction of virus. Based on these results, MWD procured microfiltration units for five of their desert stations.

Barrow, Alaska. A 40 GPM combined microfiltration and nanofiltration plant is described by Lozier *et al.* (1997). The feed water is cold, contains high quantities of disinfection byproduct precursors, and microbial contaminants, *Giardia* and *Cryptosporidium*. For the first year of operation, the plant product exceeded all requirements proposed for water treatment and removal of disinfection byproducts. The water treatment facility plans to convert its existing conventional filtration and reverse osmosis system to the combined membrane process.

2.7 Potential Projects

New ***York City.*** The New York City water supply serves half of the population of New York State, or approximately 9,000,000 people. This system, containing 19 reservoirs, provides 1.3 billion gallons (72,000,000 cubic meters) of water daily. This works out to 144 gallons (0.54 m^3) per person, not an unusual usage for the United States. The Ashokan Reservoir, for example, has a capacity of 123 billion gallons (6.8 billion cubic meters) and a residence time of 8 months. The principal purification method is simply long residence time.

Once called the "Champagne of drinking water," this water is being threatened by the effects of development and agricultural practices in the watershed. Testing of the water shows a number of potential threats to public health: cryptosporidia, *Giardia lamblia,* and E. coli.

The Croton reservoir system supplies 10 percent of this water. It receives effluent from 53 sewage treatment systems in addition to a number of septic systems and surface storm runoff. Estimated capital cost of a filtration plant is $600 to $700 million and $40 million annually for operation. This works out to $5/(gallon/day) and $0.84/kilogallon. The Catskill and Delaware systems which supply 90 percent of the water are less threatened. If water quality drops below Federal standards, the city will have to construct a filtration system. Such a plant would cost approximately $5 billion investment cost and $500 million per year to operate. This works out to $4.20/(gallon/day) and $1.20/kilogallon.

The system has so far obtained "avoidance" agreements extending to the year 2002. An extensive program of watershed protection and monitoring of species like Cryptosporidium has been undertaken (Ashendorff, 1997). This includes a recently signed memorandum of agreement for a cooperative effort on the part of all stakeholders. EPA's Expert Panel (Okun, 1997) has concluded that disinfection alone will not provide adequate protection against cryptosporidiosis for the users of this water. Both watershed protection and filtration will be necessary for this water supply.

It is highly speculative whether microfiltration fits into this picture. At this time, there is no "barrier," as the water quality regulators express it, between the reservoirs and the taps in New York City. This is a highly politicized issue and the best technical answer may not be the best political answer. A variety of non-structural solutions, like land acquisition, restrictions on land use, and tightened agricultural regulation have been proposed, and some are being implemented. There will probably be a filtration plant on the Croton system. MF would almost certainly meet or exceed their requirements. Where application of microfiltration seems most likely is in the treatment of some of the more seriously contaminated waters that currently flow into the reservoir system. Draft regulations for sewage treatment plants located within the watershed have been revised to include microfiltration. Kachalsky (1994) presents the results of tests run on effluent from the City's Brewster Treatment Plant. Four microfiltration units and three ultrafiltration units were tested. Microfiltration was recommended to achieve the required reduction of targeted organisms.

Others. It is estimated that over 300 million gallons per day of microfiltration and ultrafiltration projects are on the drawing boards for startup in the next 4 years within the United States. Table 2.3 is a summary of MF plants and plant expansions planned for the near future. Among the plants being discussed are: 5 MGD by Los Angeles Water and Power at Terminal Island and within 3 to 4 years expanding to 15 MGD; Orange County Water District will be scaling up to 5 MGD of MF or UF, eventually going to 15 MGD; San Diego is planning 24 MGD by 2000; West Basin Water District in California is starting with 2.5 MGD and will be expanding to a total of 5 MGD before 2000 to provide treated water for industrial cooling; Sacramento is studying 22 MGD of MF or UF treated water from the American River using Ranney collectors; Tampa Bay, Florida, is studying the pretreatment of 25 MGD for reverse osmosis; San Francisco is studying a 24 MGD project; Tucson, New Orleans, and a city in Wisconsin are studying projects that total 140 MGD; and Minneapolis is studying membrane treatment of their water supply.

Table 2.3.—MF plants and plant expansions proposed for the near future

City	Capacity	Remark
Los Angeles (CA) water and power	5 MGD	Expanding to 15 MGD
Orange County Water District, CA	5 MGD	Expanding to 15 MGD
West Basin Water District, CA	2.5 MGD	Expanding to 5 MGD for industrial cooling
San Diego, CA	24 MGD	By year 2000
Sacramento, CA	22 MGD	
Tampa Bay, FL	25 MGD	For RO pretreatment
San Francisco, CA	24 MGD	
Tucson, AZ, New Orleans, LA, and a city in Wisconsin	Total of 140 MGD	
Minneapolis		Water supply treatment

2.8 Research and Development Efforts

The plants and demonstrations suggest that microfiltration is an economically valid process for water treatment where the desired results are *Giardia* and bacterial removal at a level of 3 to 3.5 log. Virus removal is very modest; the best estimate is about 0.5 log. However, if there is not a great quantity of THM formation potential, the combination of microfiltration with chlorination can be a satisfactory process.

Freeman (1994) has investigated treating backwash water from conventional filtration via microfiltration. Cost estimates from this study were $0.67/gallon per day for MF and $0.33/gallon per day for a plate settler. These were considered close enough to encourage testing.

Vickers (1994; 1995) has studied use of microfiltration in conjunction with coagulation processes. The objective of this process is the removal of organic matter, particularly disinfection by-products. An advantage to MF or UF is that the coagulant and process can be optimized for organic removal instead of particulate removal.

Jacangelo *et al.* (1995) have run extensive tests on the mechanism of cryptosporidium, giardia, and MS2 virus removal by MF and UF. They have concluded, *inter alia*, that initially removal is dominated by sieving action of the membrane, but that late in a cycle, cake formation dominates in removal. Over a number of cycles, the irreversible fouling layer dominates over cake formation.

Adham *et al.* (1995) have evaluated methods of assessing membrane integrity for several low-pressure membrane technologies. Methods tested include bubble-point testing, air-pressure-hold, sonic sensing, particle counting, particle monitoring, and turbidity monitoring. The air-pressure-hold test, though not continuous, was found to be most effective. Particle counting was most sensitive of the continuous methods.

Pretreatment for Water reuse and RO Pretreatment. Lozier and McKim (1994) carried out an evaluation of the suitability of microfiltration for RO pretreatment. They concluded that filtrate from some microfiltration equipment exceeds all industry criteria for RO feedwater. However, the removal of metals and reduction of silica which occurs in lime softening does not occur with MF. They found that MF provides from 2 to 4 log reduction of bacterial viruses when used with alum coagulation. Jolis and Hirano (1994) noted that following microfiltration they found negligible fouling or solids deposition on the sleeves of *W* lamps used for sterilization. Their tests on operation of an RO system following a MF unit were still in progress after 5 months.

Chapter 3
ULTRAFILTRATION

3.1 Basics of Ultrafiltration

Ultrafiltration (UF) is midway between MF and NF in degree of separation. Separation is accomplished through sieving just as with MF, but while MF membrane retains particles and allows macro molecules and colloids to pass through, UF membrane retains most molecules and particles larger than its molecular mass cutoff (MMCO). Colloidal silica, protein, humic substances, pyrogens, and viruses would be retained by a UF membrane but dissolved salts, sugar, and mid-sized organic molecules would not.

UF is a relatively new process for municipal water treatment. It has been used more extensively in industrial processes for concentrating milk, protein, and cells; recovering process materials such as paint, colloidal metals, and dyes; and recovering oily waste water. In the last 5 years, the application of UF in drinking water treatment has expanded profoundly due to changes in drinking water quality regulations. The Surface Water Treatment Rule (SWTR) of 1986 requires 3 log removal or inactivation of *Giardia* cysts and a 4 log removal of enteric viruses. An enhanced Surface Water Treatment rule may increase the required reduction in ***Giardia*** and ***Cryptosporidium*** to 6 logs for some source waters. At the same time, the Disinfectants/ Disinfection By-Product Rule (DBPR) will reduce the allowable concentration of disinfection by-products, but drinking water must still contain a residual disinfectant. It is difficult to meet all of these requirements with conventional media filtration and coagulation/clarification processes. These processes typically rely on disinfection to inactivate any remaining organisms, but they do not completely remove the disinfection by-product precursors (naturally occurring humic and fulvic acids). So, when the disinfectant is added, they will be able to meet the SWTR, but not the DBPR! UF membranes do remove DBP precursors and can easily meet 6 log reduction in organisms ***Giardia*** and ***Cryptosporidium*** if the MMCO is low enough.

UF membrane comes in all of the configurations shown in figure 1.3 (Spiral Wound, Tubular, Compound Tubular, Flat Sheet) except hollow fine fiber. If the feed water suspended solids content is low, spiral wound modules are best since they have the highest area/volume ratio of all UF type configurations. But if the feed water does contain a high concentration of suspended solids, one of the tubular configurations will be easier to maintain.

3.2 Membrane Characterization

There are four important aspects necessary to describe UF membrane: configuration, composition, molecular mass cutoff, and clean water permeability. Of these four, the only one that is well described by manufacturers is the configuration. They are fairly good about telling you the dimensions and whether the membrane is in a spiral wound, tubular, compound tubular, or hollow fiber configuration. They will also give you an idea of the composition such as PS (polysulphone), CA (cellulose di-tri acetate), PVF (polyvinylfloridine), or ceramic. They may further assist the customer by pointing out that one has high fouling resistance and the other is good for oily waste water. Maybe that is the best one can hope for. Manufacturers do provide detailed information on the last two items; unfortunately, it mostly serves to confuse matters when trying to compare one membrane to another, especially among manufacturers.

3.2.1 Molecular Mass Cutoff

UF membrane is described chiefly by its MMCO which is more often called the MWCO (molecular weight cutoff). MWCO is not entirely accurate since molecules are always described by their mass rather than their weight. Weight is the mass of an object under the influence of Earth's gravity. A 100 lb weight will weigh much less than 100 lb on the Moon, but it will have a mass of 45.45 kg no matter what planet it is taken to. This may seem trivial, but some people are fanatical about being exactly correct in language. Whether it is called MMCO or MWCO it is still determined in the same general manner, though there is no standard method. The membrane is challenged with substances of known molecular mass such as latex spheres, dextran, protein, or other organics. Retention rate is determined for a range of molecule sizes; then the MMCO is defined as the molecular mass of molecules retained at an arbitrary level. Most manufacturers use 90 percent.

There are problems with this method of characterization and also with the interpretation of the results. Different companies challenge their membranes with different substances at different concentrations and then use different methods for determining the retention rate. Who can say whether a membrane with a MMCO of 100,000 Daltons determined using 2,000 mg/L of a "polymer marker" will perform similarly to one with the same MMCO determined using polyethylene glycol at 1,000 mg/L? Standard methods would help clear up this confusion, but the MMCO statistic is still only an approximation. Membrane pores are not homogenous or evenly spaced. Some will be much larger than the MMCO and others much smaller. Figure 3.1 is a generic molecular mass versus retention curve for a membrane with a sharp MMCO, or narrow pore size distribution, and one with an indistinct MMCO, or wide pore size distribution.

Another problem with assigning a MMCO is that many organic molecules are adsorbed onto the membrane surface forming a cake layer which reduces the effective pore size. The rate of adsorption depends on the organic molecules present, their concentration, the pH, flow velocity, pressure, temperature, ionic strength, and membrane material. As material builds up at the membrane surface, the pores are closed in from the sides. Figure 3.2 illustrates the difference in retention of a membrane before and after cake formation.

The Seventh Edition Perry's Handbook gives a rule of thumb for estimating UF separation abilities. The molecular mass should be a factor of 10 greater than the MMCO for a good separation using UF membrane (Perry and Green, 1997). In practice, the customer must test

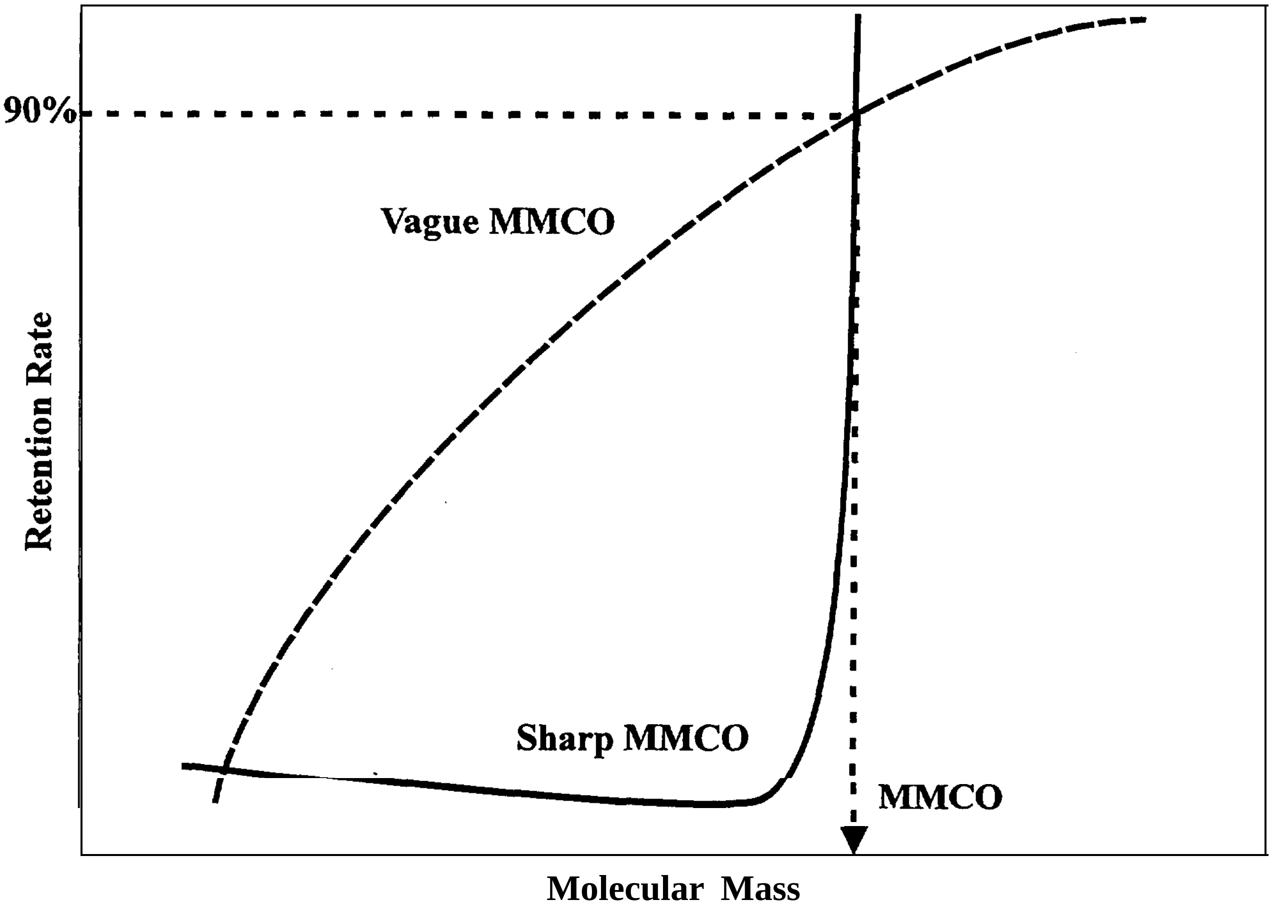


Figure 3.1 .-Comparison of molecular mass and retention rates for membranes with narrow and wide pore size distributions.

a range of MMCOs and membrane materials, since the pores of one material may differ from those of another, to determine which membrane will be adequate for the source water and treatment objectives.

3.2.2 Clean Water Permeability

What is clean water? I am sure that the number of definitions for clean water is only limited by the number of UF membrane manufacturers. Some add sodium chloride, some use RO permeate or local tap water. In reality, UF is not normally used for treating "clean" water. The reason one wants to know the clean water permeability is to get an idea of the membrane module productivity to estimate how many to purchase. But the module productivity has nothing to do with the clean water permeability! It depends on the pressure, feed flow rate, recovery rate, fouling rate, backwash frequency, and temperature and probably we'll find that the phase of the moon has a major impact when we *really* understand the process. Manufacturers measure the clean water permeability so that they can tell you something about the membrane. Its easier and more direct than giving lots of pilot test data from different water sources that still may be inapplicable to your particular situation and may

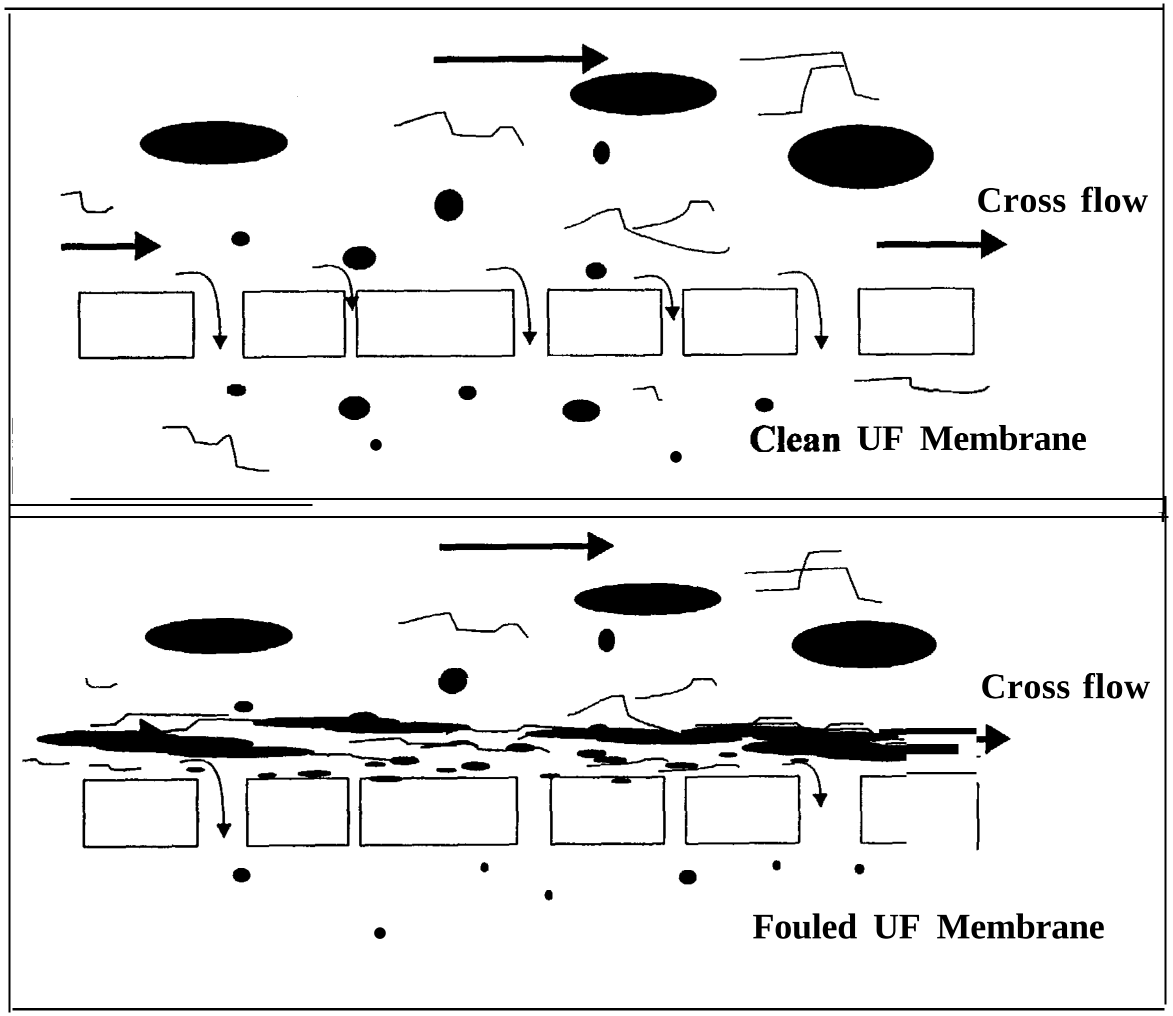


Figure 3.2.—Difference in retention between clean and fouled UF membrane.

lead you to believe their product is not as good as another. The trouble is that UF is so profoundly dependent on the feed water quality that it is complicated to predict how a UF membrane will perform. The only way to get the best membrane for the job is to pilot test a few different types and MMCOs. Luckily, there are more and more documented pilot tests available now that may involve testing a feed water similar to the one you need to treat. The AWWA Membrane Conference Proceedings from the last few years have several case studies. The manufacturers will also have documented case studies and will be able to help you choose a few good membranes to test.

3.3 Process Design Considerations

UF is most always operated in a cross-flow mode. There are references to operating in dead-end mode (see Chan et. al., 1997), but these systems are designed to handle a wide range in water quality. They are automated to switch to cross-flow mode when turbidity or some other monitored parameter exceeds the setpoint. Figure 3.3 compares the averages of productivity for all seawater (SEA) and brackish RO, NF, UF, and MF membranes listed in chapter 10. UF membrane "clean" water permeability is about twice that of NF membrane and 2.5 times that of brackish water RO membrane. In practice, however, these high flow rates are only realized with feed water of the same quality as is needed for RO and NF treatment. UF is not normally used on such clean water, though. With high turbidity water (in excess of about 15 NTU in the Chan study), concentration polarization becomes a limiting factor in UF productivity (for a perspicuous discussion of concentration polarization and UF see' Perry and Green, 1997, p 22-53).

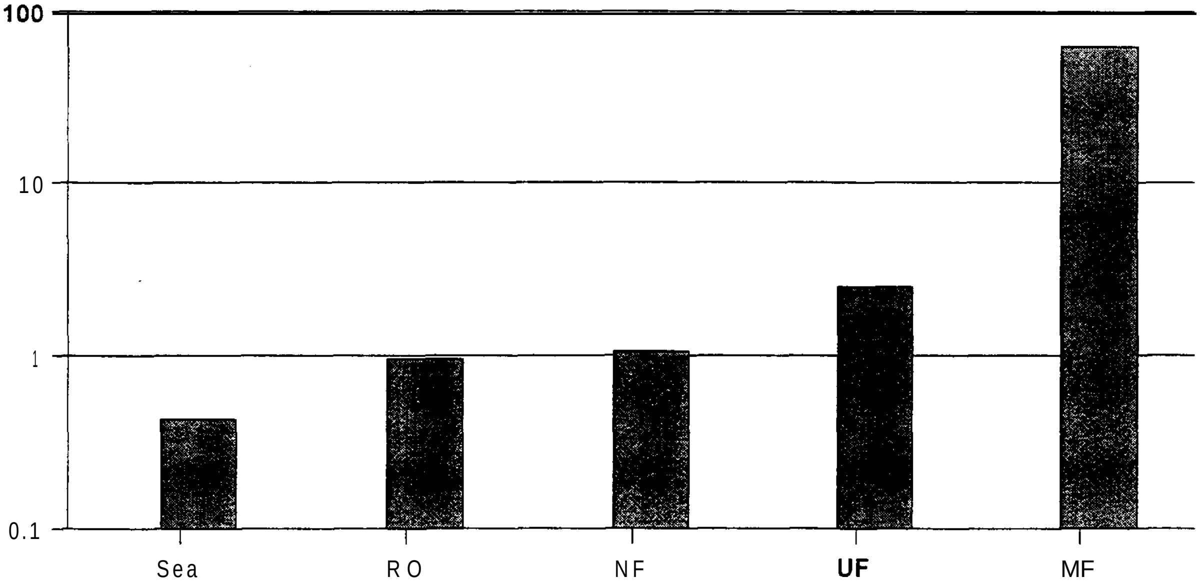


Figure 3.3.—Comparison between membrane productivity for seawater and brackish water RO, UF, and MF membrane as tested by the manufacturers. Productivity units are $m^3m^{-2}day^{-1}$.

Concentration polarization is the build-up of material at the membrane surface. In MF, this material is blown out periodically in the backwash. In RO and NF, it is minimized by extensive pretreatment, high flow rates, and by using feed channel spacers designed to promote turbulence. UF systems incorporate a combination of these strategies. The systems are operated in cross-flow mode at high shear rates and low pressure; feed channel spacers are designed to promote turbulence; and the systems are periodically backwashed. High recovery is achieved by recirculating the concentrate stream to the feed tank, back into the feed line, or into another stage of membranes. Figure 3.4 illustrates the three UF process designs with pre-filters, permeate backwash, and concentrate recycle.

3.4 Process Instrumentation and Evaluation

There are three purposes for instrumentation in UF treatment processes. These are to monitor:

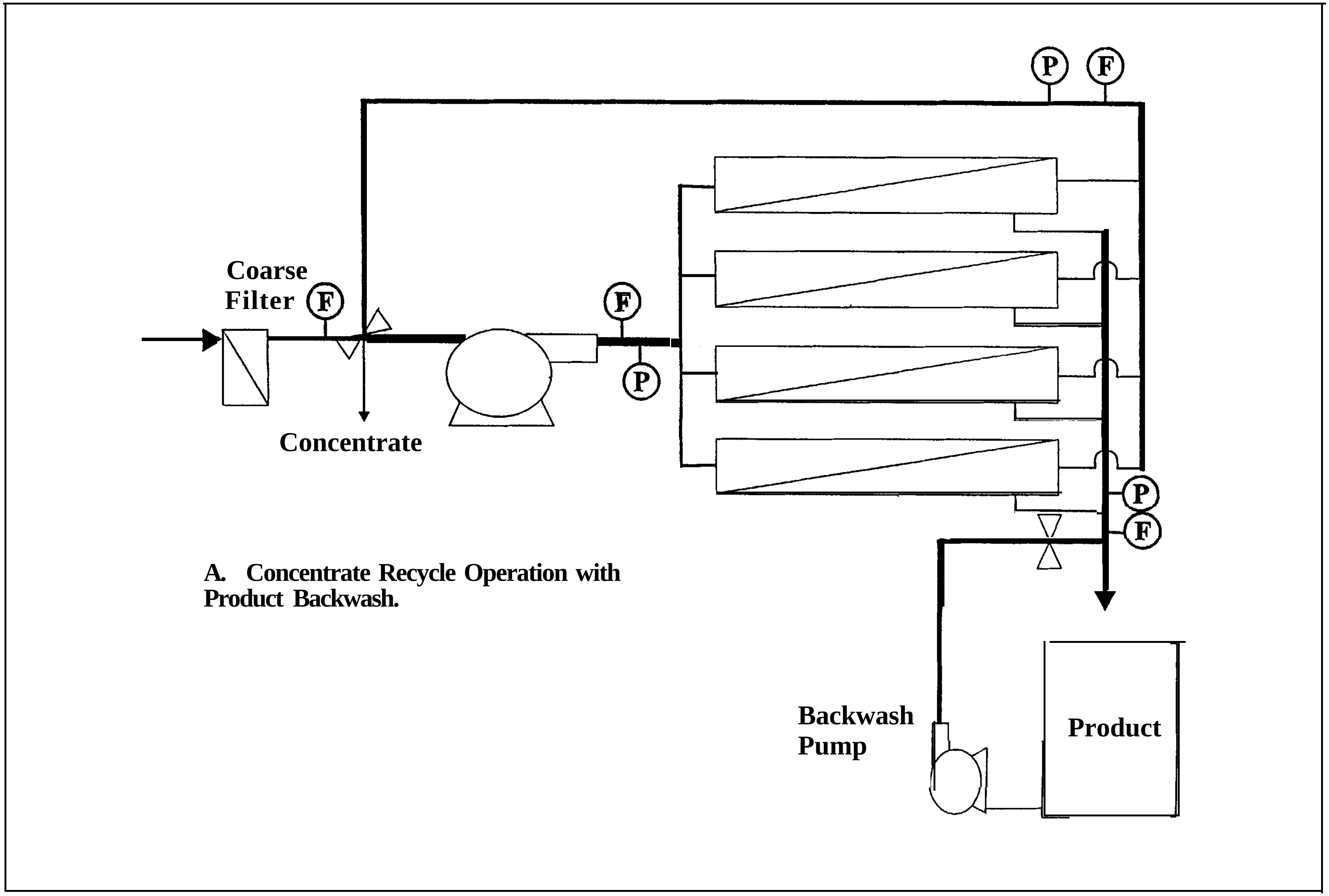
- Hydraulic performance,
- Retention performance, and
- General water quality.

Hydraulic monitoring requires measurement of the feed and permeate pressure, feed, permeate, recycle and backwash flows, and temperature of the feed stream. Retention monitoring requirements depend on the source water and treatment objectives. Operators should be able to tell, with a glance at the monitoring instrumentation, whether the product water is within specifications or not. Surface water treated for drinking purposes should be monitored for turbidity and particle count of the feed and permeate. General water quality monitoring for UF is the same required of any drinking water process to provide documentation that the water meets Safe Drinking Water Standards. This should include at least conductivity, pH, and chlorine residual. Table 3.1 summarizes monitoring equipment for a UF water treatment system by purpose.

UF is evaluated in a number of ways depending on the treatment objectives. Table 3.2 lists performance parameters used in various UF pilot studies reported recently. These are explained in greater detail following the table. Some, such as the net permeate volume per unit energy or average power consumption would only be monitored during a pilot study to optimize system operation. Others, such as the heterotrophic plate count (HPC), *Giardia* and *Cryptosporidum* enumeration, Total Coliform (TC), and Bacteria Virus Assay (BVA) would be required of any drinking water treatment plant. The rest are necessary to monitor performance of a UF system.

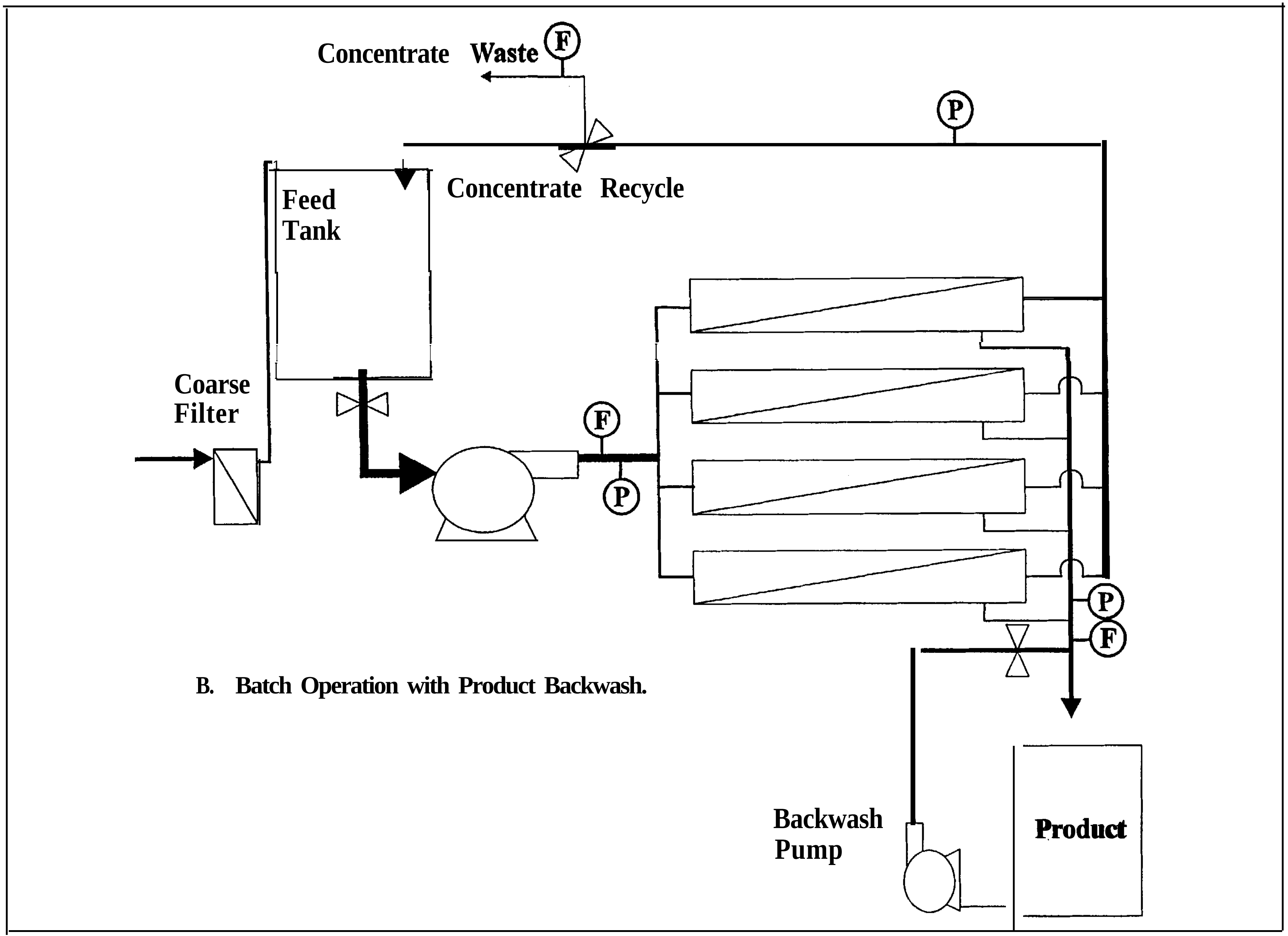
Transmembrane Pressure (TMP)

TMP should be monitored continuously or at least twice a day. In automated systems, the backwash cycle is initiated either on a timed basis or by a specific increase in the TMP. It is calculated as follows from three pressure measurements.



A. Concentrate Recycle Operation with Product Backwash.

Figure 3.4a.—Concentrate recycle operation with product backwash.



B. Batch Operation with Product Backwash.

Figure 3.4b.—Batch operation with product backwash.

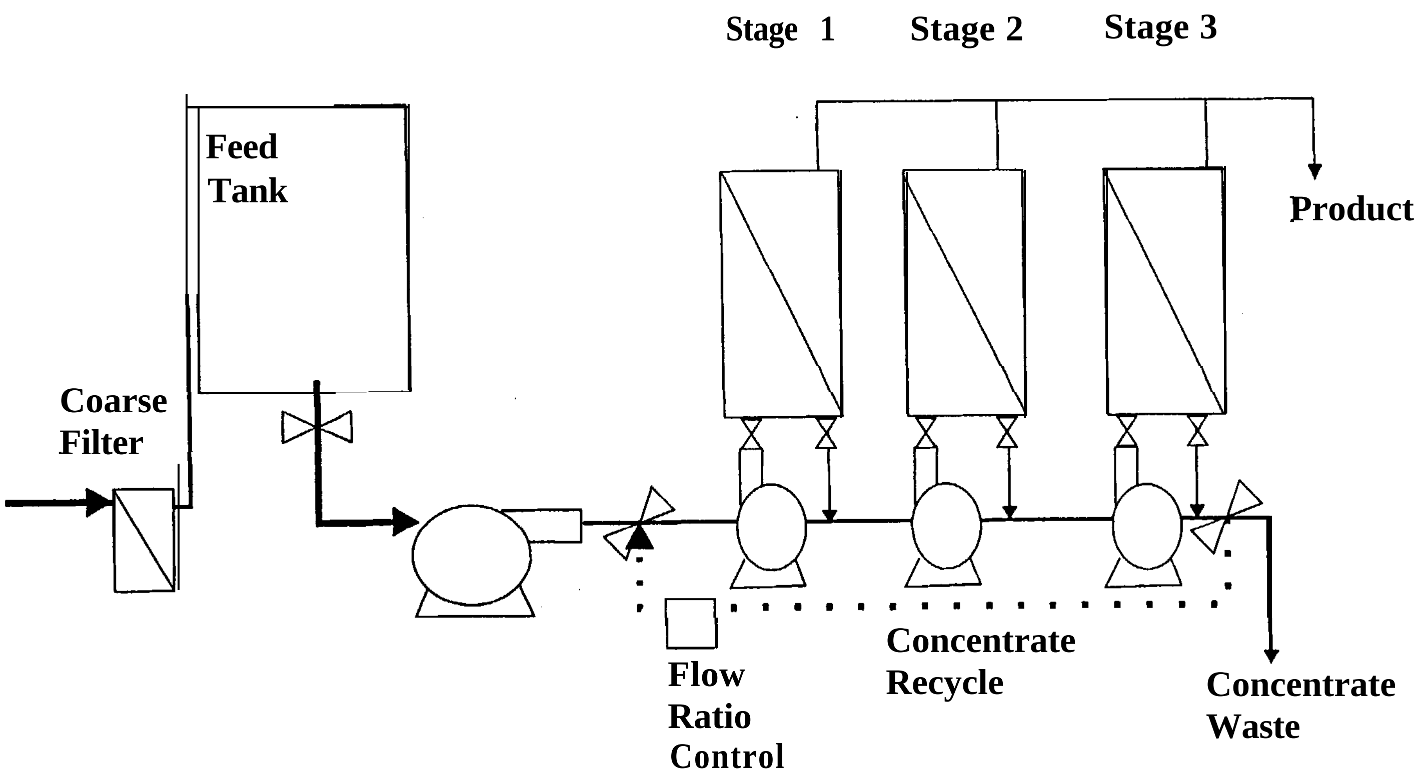


Figure **3.4c.—Stages** in series. Each stage flow and TMP are controlled separately. At a minimum feed, individual product, and concentrate recycle flows and pressure across each stage must be monitored.

$$TMP = 0.5(P_f + P_r) - P_p \qquad \text{Eq. 3.1}$$

Where: P_f = Feed pressure
P_r = Recycle pressure
P_p = Permeate pressure

Table 3.1 .-UF instrumentation requirements

Function	Parameter	Feed	Permeate	Recycle	Concentrate	Backwash
Hydraulics	Pressure	X	X		X	
	Flow	X	X	X		X
	Temperature	X				
Retention	Turbidity	X	X			
	Particle Count	X	X			
Water quality	Conductivity		X			
	pH		X			
	Chlorine Residual		X			

Table 3.2—UF performance evaluation parameters used in recent case studies

Objective	Performance parameters	Reference
Optimizing operation using backwash time and powedered activiated carbon (PAC) addition as variables	Net permeate vol/unit energy Net permeate vol/average power requirement	Marriott, Clark and Lainé, 1997.
Removal of color and TOC from surface water w/ and w/o chemical addition.	Water quality, TMP, and run length.	Scanlan et. al., 1997.
Testing pretreatment for RO in recovering waste water from petrochemical plant.	Permeation rate and turbidity.	Wong et. al., 1997.
Determining effect of charge and coagulant dose on NOM Removal and fouling rate.	Permeation rate, UV@254 and TOC.	Fu and Dempsey, 1997.
Color, TOC, iron, bacteria removal.	Water quality, TMP, permeation rate	Chan et. al., 1997.
Microbial removal and integrity verification.	Permeation rate, particle count, turbidity, SDI, TMP, TOC, UV@254 HPC, TC, BVA, ***Ctypto,*** and ***Giardia*** enumeration.	Jacangelo, Adham and Lainé, 1997.

Net Permeate Flow (Q_{Pnet})

Q_{Pnet} is used to evaluate performance using different backwash frequencies and durations. It accounts for product water used during backwash and cleaning. It should be calculated from actual flow rates, not the temperature normalized flow.

$$Q_{Pnet} = Q_P - Q_{BW} - Q_C \qquad \text{Eq. 3.2}$$

Where: Q_p = Total product flow ($Lhr^{-1}m^{-2}$)
Q_{BW} = Product used for backwashing ($Lhr^{-1}m^{-2}$)
Q_C = Product used for cleaning ($Lhr^{-1}m^{-2}$)

Normalized Permeate Flow (Q_{Pn})

Permeate flow may be normalized for temperature to allow comparisons over time with large temperature fluctuations. This is not as critical as with NF and RO processes, though, because UF is not operated at its full permeation capacity. Manufacturers generally provide a temperature normalization equation specific for each type of membrane they produce that adjusts for the change in viscosity of water and changes in the membrane with temperature. The following equation corrects for changes due to viscosity only (Streeter and Wiley, 1985).

$$Q_{Pn} = \frac{Q_{Pnet} * e^{-0.0239(T-20)}}{Area} \qquad \text{Eq. 3.3}$$

Where: Q_{Pn} = Permeate flow ($Lhr^{-1}m^{-2}$) adjusted for viscosity at temperature T
Q_{Pnet} = Net Permeate flow ($Lhr^{-1}m^{-2}$)
T = Temperature (“C)
Area = Membrane area (m^2)

Net Permeate/Net Energy (Q_{Pn}/E_n) and Net Permeate/Average Power Consumed (Q_{Pn}/W_{avg})

Q_{Pn}/E_n and Q_{Pn}/W_{avg} are parameters used by Marriott and associates (1997) to optimize a UF process using backwash frequency and duration, permeation rate, and PAC addition as variables. The energy and power terms, as defined in the reference, are given below. Q_{Pn} is as described above.

$$E_n = E_d + E_{bf} = \sum P_d Q_d \Delta t_d + \sum P_{bf} Q_{bf} \Delta t_{bf} \qquad \text{Eq. 3.4}$$

$$W_{avg} = \frac{E_n}{t_d + t_{bf}} \qquad \text{Eq. 3.5}$$

Where: E = Pump energy (joules)
P = Pressure (Pascal)

Q = Flow rate during time At (m^3/s)
At = Time period (sec)
W_{avg} = Average power consumption (watts)
t = Total time during each cycle (sec)

Summations are of one second intervals (or any other convenient time interval) over the duration of the test. Subscript "n" means "net," "bf" refers to the backwash cycle, and "d" refers to production cycle .

Using compound parameters such as these helps to normalize cost/benefit tradeoffs for different scenarios. The volume per energy term is based on the total energy required. An efficient process would have a high Q_{pn}/E_n value. The volume per average power term is based on the power consumption averaged over the same time period. If the process is efficient and timely, then the Q_{pn}/W_{avg} value would also be high. Operating with a very conservative permeation rate and backwash cycle would be reflected in a high Q_{pn}/E_n value and relatively low Q_{pn}/W_{avg} value. Even though the system would operate smoothly, it would be not be operating to its full potential-

3.5 Integrity Testing Methods

If a UF membrane is the main barrier between pathogenic organisms and the public, it is reassuring to be able to monitor its integrity. On-line monitoring would be the most desirable, but it is difficult to make a direct correlation between measurable parameters and the presence of pathogens. Jacangelo and associates (1997) reviewed a number of methods classified as direct or indirect methods, though the distinction between the two classes is rather gray. Direct methods monitor for a change that is directly traceable to a flaw in the membrane module and/or system. Examples of direct monitoring are air pressure hold and bubble point testing. Indirect methods monitor changes that infer a flaw in the membrane and/or system such as particle counting and turbidity monitoring.

The objective of the Jacangelo study was to determine the ability of UF and MF membrane to remove microbes. One spiral wound, one tubular, and four hollow fiber membranes were compared on five different source waters. The membranes were subjected to pin pricks and cut fibers during the integrity testing described in Jacangelo's report. Only the direct methods of air pressure hold and bubble point testing were able to detect the induced flaws in the spiral wound membrane. The problem is that a spiral wound membrane has a

Table **3.3.—Integrity** testing and monitoring techniques for UF membrane

Technique	Description	Comments
Air pressure hold test	Permeate side of the membrane is drained and pressurized with air. Pressure is monitored for 10 minutes.	System must be shut down.
Sonic sensing	Sound wave sensors are attached to the membrane module during the air pressure hold test to listen for leaking bubbles.	System must be shut down. Extraneous noise could interfere with the test.
Bubble point test	Mainly used for determining MF pore size. UF membrane pores are too small. The method can be used to detect broken fibers in a bundle, though.	Module must be taken offline. Cannot detect leaks in the system piping and gasketing. Only good for fibers.
Particle counting	Light scattering used to enumerate particles in one or more size ranges.	Low upper limit on particle counts: feed particles are too numerous to count. Not sensitive enough for small flaws. Need correlations between particle size and bacteria/virus/spore concentrations.
Particle monitoring	Measures fluctuations in light intensity, output is an index of relative water quality. Less complicated and expensive than particle counters.	High upper limit, but feed may still exceed limit. Not sensitive enough for small flaws. Need correlations between particle detection and bacteria/virus/spore concentrations.
Turbidity monitoring	Measures degree of light scattering by suspended solids.	Not sensitive enough for small flaws. Need correlations between turbidity and bacteria/virus/spore concentrations.
Silt density index	Measure of fouling potential. Calculated from the difference in time required to filter 100 ml at constant pressure through a **0.45µm** filter at the start and end of 15 minutes.	No correlation between SDI and presence of bacteria/virus/spores. Not sensitive enough to detect small flaws.

(Based on Jacangelo et. al. 1997).

far greater surface area than the hollow fiber and tubular modules tested. In this way the spiral wound module is a good model for a large membrane system of any configuration. A small flaw in a large system will be difficult to detect by any of these methods.

3.6 Concentrate Issues

UF concentrate consists entirely of backwash and cleaning water. The volume of waste depends on the frequency of backwashing and cleaning. A survey of full size UF and MF plants revealed a range in recovery rates of 70 to 99 percent. Extremely high recoveries, without additional treatment of the backwash, can result in build up of pathogenic organisms in the system and is not recommended for drinking water treatment. Scanlan and associates (1997) had trouble with accumulation of algae in their backwash solution. They had saved it, though, for batch treatment at the end of raw water testing. Algae growth can be prevented by keeping sunlight out of the system. If only a portion of the backwash is constantly recycled to the feed, the problem of pathogen build-up and algae growth should be minimized. Since the mineral make up of the backwash water is similar to the feed source, it can be discharged to a settling pond, disinfected, and then recirculated to the feed stream or discharged to the source. Some plants use the backwash for irrigation (Jacangelo, 1997). Its high natural organic matter content makes it an excellent fertilizer.

Cleaning solutions need to be neutralized before disposal. Depending on the solution used, they could be discharged to the sewer or disposed of with the backwash. Typical cleaning solutions are sodium hypochlorite, sodium hydroxide, hydrogen peroxide, or detergents.

3.7 UF Case Studies

Examining case studies of an unfamiliar process is a good way to illustrate how the process is used and what kind of problems may come up when testing the process. The following are studies reported in the AWWA Membrane Technology Conference Proceedings from 1995 and 1997.

3.7.1 Removing TOC and Color

Scanlan and associates compared MF and UF with conventional sedimentation and dissolved air floatation (DAF) for treating mountain runoff with low turbidity, moderate alkalinity, but high color and TOC to produce drinking water for the city of Cheyenne, Wyoming. Their objective was to remove particles, disinfection bi-product precursors (DBPs), color, and soluble manganese with minimal chemical addition. Table 3.4 shows the various treatment combinations.

A Memcor, 0.2 - 0.5 micron, MF system with automated backwash using air and water was used. Between tests, the membranes were cleaned with citric acid and caustic. The Koch,

Table 3.4.—Treatment alternatives for Cheyenne, Wyoming

1	2	3	4
Alum or Ferric	Alum or Ferric		
Flocculation	Flocculation	Strainer	Alum, Ferric or PAC
Tube settler	Solids floatation	200µm bag filter	Strainer
Dual media filter	Dual media filter	UF	MF

0.05 µm, UF system was backwashed at 30 minute intervals with a water/hypochlorous acid/caustic solution and cleaned between tests with a proprietary caustic solution. Because of the system configurations, the two systems were operated on different principles. Constant permeate flow was maintained on the MF system, while constant pressure was maintained on the UF system. It would be best if both systems could be operated in the same manner, because when the pressure is increased to maintain a given permeation rate, fouling becomes more difficult to remove with ordinary backwash cycles. One must work with what one has though; since they could not compare TMP changes or energy usage differences meaningfully, they compared run length and volume and quality of water produced.

During the test, the water was analyzed for alkalinity, turbidity, color, chlorine, manganese, TOC, total trihalomethane (TTHM), and particle counts in the range of 2 to 15 µm, the approximate size of *Cryptosporidium* and *Giardia*. Their findings are listed below.

- All the treatments reduced turbidity to 0.15 NTU or less.
- All treatments did a satisfactory job of reducing color, though increasing ferric sulfate doses resulted in increased color removal.
- MF and UF had 2.6 and 3.6 log removal of particles, respectively, compared to 1.5 to 2.2 log removal with sedimentation and DAF.
- UF and MF removed significant amounts of manganese without chemical addition. Sedimentation and DAF removed manganese to the same extent when alum was used as coagulant.
- UF and MF were not able to remove TOC to the target 45 percent level. Even with 50 mg/L coagulant or PAC, MF TOC removal increased to only 36 percent. DAF and sedimentation were able to remove 44 percent TOC with the same coagulant dose and 49 percent with 70 mg/L.
- MF had longer run times than UF. The UF system was unable to maintain the design permeation rate, but did perform consistently at 1/4 the design rate.

In this case, the UF membrane was unable to perform as desired. but if they had tested a UF with a much smaller MMCO (0.05 µm ≈ 100,000 Daltons). or used PAC pretreatment with the

UF, they could have come to completely different conclusions. Here. the objective of attaining high water quality with minimal chemical addition will not be met because the apparent winners of the test are the conventional sedimentation or DAF processes with 50 to 70 mg/L alum or ferric sulfate.

3.7.2 Enhancing UF Organic Removal With PAC

In another example, a group at the University of Illinois piloted a UF system in conjunction with a floc blanket reactor (FBR) and PAC to remove organic contaminants (Schirnmoller, 1995). Organic molecules are adsorbed onto PAC at a rate dependent on the concentration of organic compounds and the retention time. As organic concentration increases in the UF concentrate stream, the PAC can be more fully utilized. The extended retention time in the FBR also helps improve PAC utilization. Figure 3.5 illustrates the configuration of such a system. The PAC is injected just ahead of the UF system. PAC that remains in the reject stream is recycled back to the UF. When the membranes are backwashed, the PAC that has built up on the surface is recycled to the FBR. Solids are removed from the FBR as needed. It turns out that the PAC in the UF feed stream actually slows the rate of flux decline. This could be due to the scouring effect of the PAC particles, and/or the PAC could improve the permeability of the filter cake layer by making it less compressible.

Groundwater from Urbana, Illinois, was used as feed water for this study. It had 2.5 mg/L TOC and turbidity from 0.1 - 1 .O NTU. The FBR effluent was treated with hollow fiber UF membrane with 100,000 Dalton MMCO. The concentrate was recycled to the UF feed tank. After reaching steady state performance, the FBR removed about 1 mg/L, and the UF system removed another 0.75 mg/L for a final reduction from 2.5 to 0.75 mg/L, an overall 70 percent reduction in TOC.

3.7.3 Testing UF as Pretreatment for RO

Wong and associates (1997) describe a pilot test to recover waste water from a petrochemical plant. This plant has three waste streams: the cooling tower blowdown (CTB) and the organic and inorganic waste water treatment system effluent (OWWTS and IWWTS respectively). Since the IWWTS has very high TDS, the OWWTS and CTB were selected for recovery and reuse. An elaborate pretreatment system was designed using potassium permanganate oxidation and greensand filtration to precipitate and remove manganese, iron, and cobalt, followed by a GAC and cartridge filter to remove organics and particulates. UV irradiation was used before the membrane systems for disinfection. They conservatively tested and optimized each process in turn all the way up the line. The GAC filter was taken out of line to see if it was really necessary. but that caused an increase in the UF TMP so the GAC was returned to the process stream.

The UF system used was designed for a waste water treatment plant. but not such a treatment plant as this! The feed to the UF system was of such high quality that they where able to replace the permeate flow restricting valve with a partially open globe valve thereby

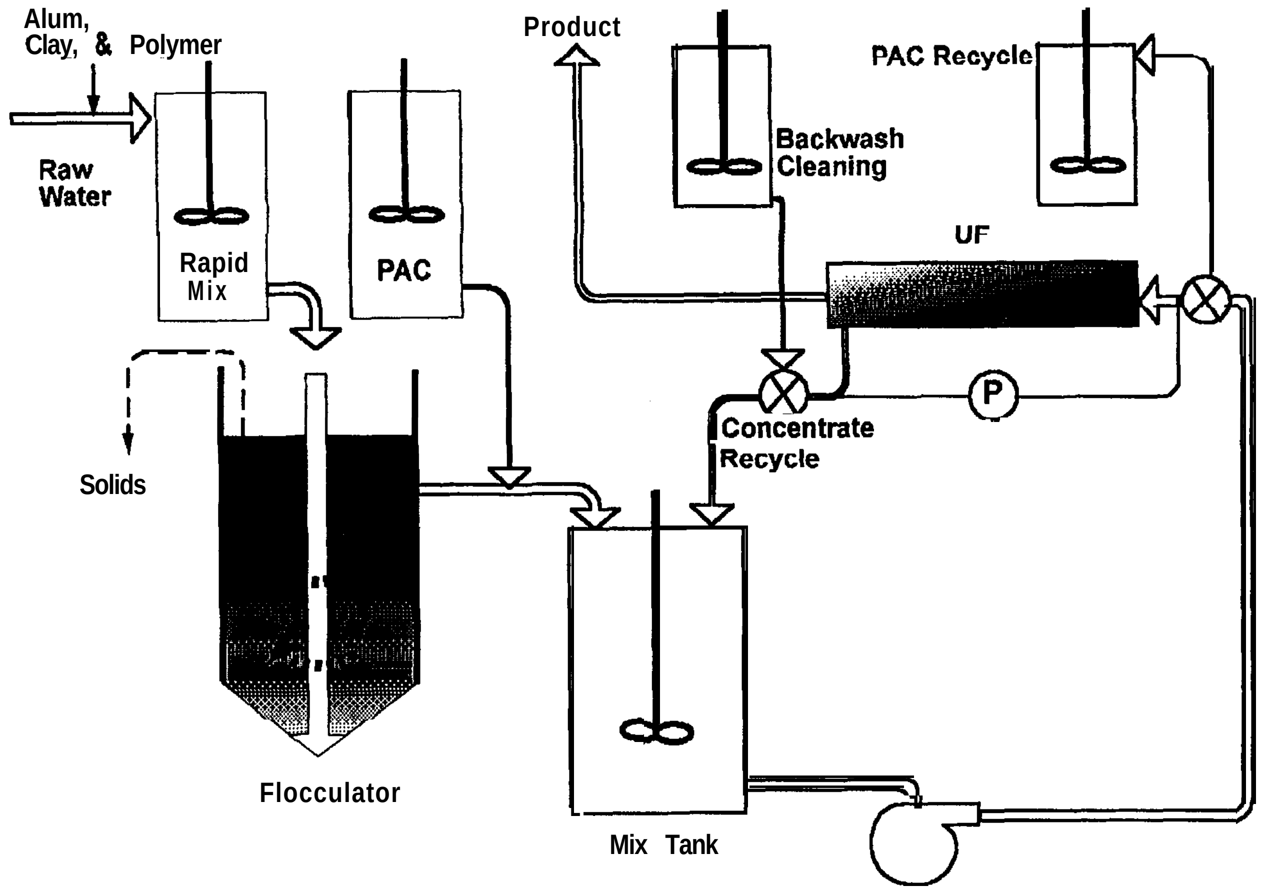


Figure 3.5.—Schematic for a Floc Blanket Reactor-UF system for removal of organic contaminants.

increasing productivity by 45 percent. This in turn improved the RO system performance by 17.5 percent through providing an optimum feed flow rate. The system operated consistently for the entire 2000 hour test. The only adverse effect on the UF system was the formation of a black layer of material on the membrane surface that could not be removed with the usual NaOCl cleaning. It was removed successfully with HCl and $NaHSO_3$, though.

This system was conservatively designed with the idea that components could always be excluded from the final design if they proved unnecessary, which is always better than finding the design won't work because proper pretreatment is not in place. UF systems do not need to have such stringent pretreatment systems, but there are trade offs. The UF system would probably have operated satisfactorily at a reduced permeation rate without the GAC, but then more membrane area would be needed to meet the RO system feed flow requirements. In this case, it was more economical to keep the GAC in the design and operate at a higher UF permeation rate.

3.8 Conclusions and Research Needs

From the case studies referenced here, it seems that applications of UF for waste water recovery, microbial removal insurance, and pretreatment for RO and NF are growing in numbers and interest. Still, there are shortfalls in three major areas:

- Membrane Characterization - Standard methods need to be established for evaluating UF membrane retention characteristics. Preferably, a method similar to that described above will be adopted where a range of sizes of differently shaped molecules would be tested and the whole retention versus molecular mass curve for each shape would be used to describe the membrane.

- Membrane Specifications - UF membrane specifications are diverse and confusing. Standard methods for evaluating and reporting permeation rates need to be developed. The standard method should include specifications for operating pressure, flow rate, temperature, and water composition.

- Integrity Monitoring - There does not seem to be a reliable on-line method for integrity monitoring. If a method cannot be found, then correlations should be developed between microbial activity and a parameter that can be measured reliably on-line.

Chapter 4
NANOFILTRATION AND REVERSE OSMOSIS

4.1 Process Fundamentals

Nanofiltration (NF) and reverse osmosis (RO) are unlike the previous filtration processes discussed. MF and UF use pressure merely to provide convective flow of liquid through the membrane. NF and RO require hydraulic pressure as well to overcome the osmotic pressure of the feed solution. Osmosis is a natural process whereby water is transported through a semi-permeable membrane from a solution of low concentration, to one of high concentration. Plants use this phenomena to draw in water from soil. Reverse osmosis occurs when pressure in excess of the osmotic pressure is exerted on the high concentration side of the membrane inducing pure water to diffuse through the membrane to the low concentration side, as depicted in figure 4.1.

The osmotic pressure of a solution, π, is the pressure that is exactly enough to prevent osmosis from occurring. It is approximated by the following equation:

$$\pi = \gamma InRT \qquad \text{Eq. 4.1}$$

Where:

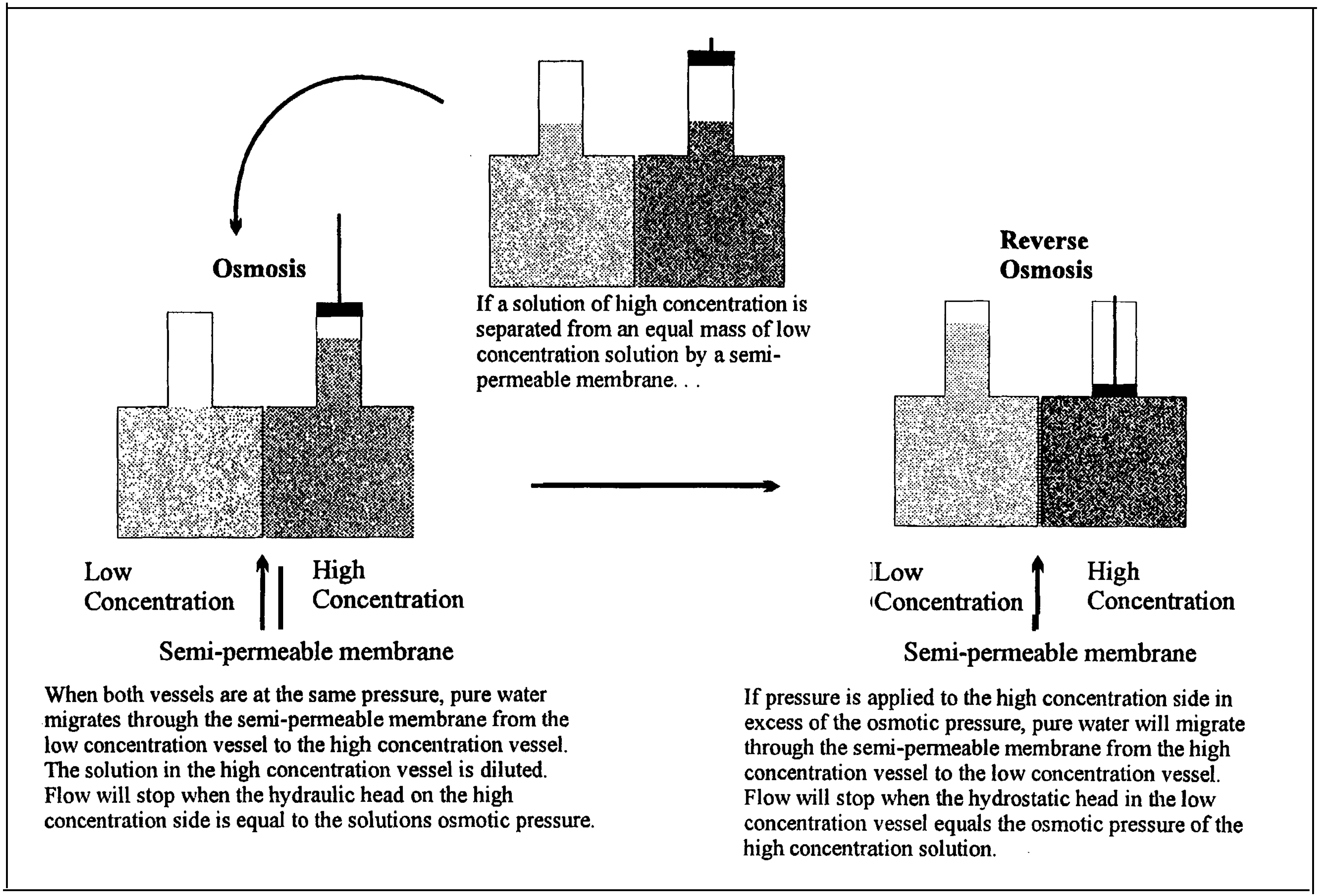
n = number of moles of ions per mole of solute,
R = the universal gas constant, 8.314 m^3 Pa°K^{-1} mole-‘,
T = °K,
I = ionic strength of the solution,
γ = activity coefficient of the solution.

$$I = \frac{1}{2}\sum (c_i\ Z_i^2) \qquad \text{Eq. 4.2}$$

c_i = concentration of the i^{th} species,
Z_i = charge of the i^{th} species.

$$\log \gamma = A\ Z_+ Z_- \left[\frac{\sqrt{I}}{1+\sqrt{I}} - B\ I\right] \qquad \text{Eq. 4.3}$$

A = 0.5091 for water at 298.15°K,
B = 0.2 for monovalent species,
Z_+, Z_- are the charges on the anion and cation.



If a solution of high concentration is separated from an equal mass of low concentration solution by a semi-permeable membrane. . .

When both vessels are at the same pressure, pure water migrates through the semi-permeable membrane from the low concentration vessel to the high concentration vessel. The solution in the high concentration vessel is diluted. Flow will stop when the hydraulic head on the high concentration side is equal to the solutions osmotic pressure.

If pressure is applied to the high concentration side in excess of the osmotic pressure, pure water will migrate through the semi-permeable membrane from the high concentration vessel to the low concentration vessel. Flow will stop when the hydrostatic head in the low concentration vessel equals the osmotic pressure of the high concentration solution.

Figure 4.1.—Osmosis and reverse osmosis.

A rough approximation is:

$$\pi\ (lb/in^2) \approx 0.01 * TDS\ mg/L \qquad \text{Eq. 4.4}$$

Non-ionized solutes, such as sugar or other organic molecules, have only one mole of particles per mole of molecules and therefore have a fraction of the influence on osmotic pressure that solutes which dissociate in solution, such as NaCl, have. A solution with one gram of NaCl per liter would have an activity coefficient of 0.877, ionic strength of 0.017, 2 moles of ions per mole of solute, and an osmotic pressure of about 74 kPa (10.8 lb/in’).

The key descriptive parameters for RO and NF membrane are their water and salt transport coefficients. The transport coefficients determine the amount of water or salt that passes through, or permeates, a unit area of membrane over a given range of salinity, pressure, temperature, and pH. Water transport is often called the “flux” and the salt transport is usually discussed in terms of the salt rejection rather than the salt passage. But flux and rejection are dependent on the operating conditions, while the transport coefficients are supposed to be intrinsic qualities of the membrane. Actually, the transport coefficients are also dependent on the operating conditions, but over the normal range of operation, the effects are minimal. Generally, it is assumed that the transport coefficients can be used to compare membranes operated under fairly similar conditions.

The water transport coeffkient, normally denoted as “A,” measures the permeation, or flow, of water through a membrane for a unit of applied pressure.

$$A = \frac{(\Delta P - \Delta\pi)}{F_w} \qquad \text{Eq. 4.5}$$

Where:

A = water transport coefficient in m $sec^{-1}Pa^{-1}$,

F_w = water flux, or volume of liquid passing through a unit area of membrane expressed as $m^3\ m^{-2}\ sec^{-1}$

A P = difference in pressure across the membrane in Pa,

A n = difference in osmotic pressure across the membrane in Pa.

The salt transport coefficient is the rate at which dissolved salt passes through the membrane.

$$B = \frac{(C_c - C_p)}{C_c\ V_c\ Area} \qquad \text{Eq. 4.6}$$

Where:

C_p = concentration of salt in the product stream in mol/L,
C_c = concentration of salt in the concentrate stream in mol/L,
V_c = volume of concentrate produced per unit time in m^3/sec,
Area = active area of the membrane in m^2,
C_c-C_p = difference in salt concentration across the membrane in mol/L,
B = salt transport coefficient for the membrane in m/s.

Salt rejection is the percentage of salt that does not pass through the membrane:

$$R = \frac{(C_f - C_p)}{C_f} 100 \qquad \text{Eq. 4.7}$$

Where:

C_p = concentration of salt in product water in mol/m^3,
C_f = concentration of salt in feed water in mol/m^3.

4.2 Module Configuration

The most common commercial configurations for RO and NF elements are spiral wound and hollow fine fiber (HFF). They are similar to those shown in figure 1.3. The main difference between HFF elements and those depicted is that HFF fibers are like strands of hair. HFF modules contain the largest amount of surface area per unit volume of any configuration; there are over 840 m^2 in a 20 x 122 cm module compared to 37 m^2 in a similar spiral wound element. The flux for HFF membrane material is only about one tenth that of spiral wound membrane, but because of the tremendous amount of active surface area per module, HFF modules have three times the productivity of spiral wound elements. For this reason, a typical spiral wound system for treating brackish water usually has two parallel sets of pressure vessels, containing 6 elements each, in series with two or three stages of vessels, while HFF systems are generally configured with parallel vessels of one or two elements each and have only one stage. Figure 4.2 shows typical spiral wound and HFF membrane systems.

4.3 Pretreatment

RO and NF are intended only to remove dissolved salts. Ideally, the feed water to these types of systems should have turbidity levels under 1 NTU and silt density index (SDI) level less than 5. HFF modules have even higher standards of pretreatment; they require under 0.5 NTU and SDI under 3. In addition, thin film composite membranes are not tolerant of free chlorine or other oxidants, such as ozone, in the feed water. Generally, manufacturers recommend less than 000 mg/L- hours of contact with free chlorine. This means that the membranes are warranted for up to 1000 hours of contact with water containing 1 mg/L free

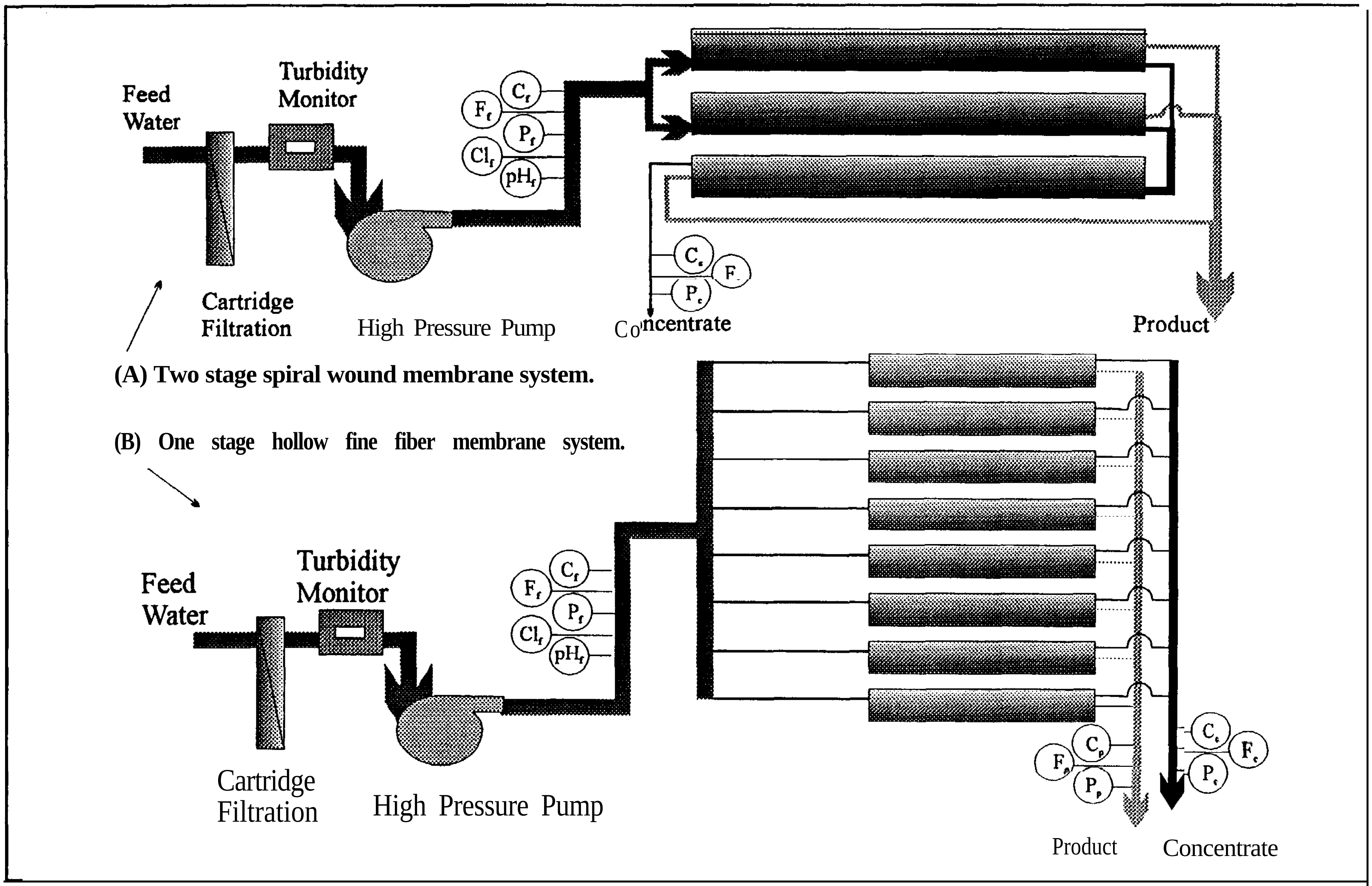


Figure 4.2.-(A) Spiral wound membrane systems require 2 stages of 6 element pressure vessels to attain a 75 percent recovery rate.
(B) Hollow fine fiber systems are designed with 1 element per pressure vessel and only 1 vessel in series.
With this 1 stage configuration they are capable of a 50 percent recovery.

chlorine, or 200 hours with 5 mg/L. Biological contamination and large organic molecules should also be removed in the pretreatment system as they tend to accumulate on (or foul) the membrane surface and cause a decline in performance. Of course, such pristine conditions are difficult to attain, but should be strived for in designing a pretreatment system. At the very least, there should be some method of disinfection and particle filtration. There are systems that incorporate chemical precipitation methods for clarifying the RO feed water, but often the added chemicals cause fouling problems in the membrane system. The real beauty of the membrane separation process is that it can be tailored to remove only what is necessary without adding more undesirable chemicals to the product water. The membrane processes described above can provide excellent pretreatment for NF and RO. Aside from possible clarification chemicals, there are a few chemicals that are commonly used to enhance NF and RO membrane performance and longevity. Acid is generally added to adjust the feed pH for cellulose acetate (CA) membranes. Anti-scalants may also be added to prevent scaling (precipitation of slightly soluble salts) in the concentrate stream.

4.4 Monitoring Performance

Temperature, pressure, and feed concentration generally determine the productivity of an NF or RO system. The membrane specifications in table 4.1 give the manufacturers limits for these parameters. These three parameters appear in the osmotic pressure equation, indicating their obvious importance in the process, but they also affect the membrane itself. You may be able to imagine what happens to a plastic wrapper in the microwave oven. Membrane in hot water behaves in the same way. It becomes loose and deformed, or even perforated, by the mesh spacer material, and the glue lines weaken. Water and salts are transported at a higher rate. Excessively high pressure can also cause the membrane to be deformed by the mesh spacer. Any fouling material on the membrane surface becomes compacted under pressure, resulting in a less permeable and more persistent layer. If the feed concentration should become higher with time, with no change in other design parameters, there could be problems with scaling or compaction, leading to possible loss in productivity. For all these reasons, it is important that the following parameters be monitored with instrumentation on all NF and RO systems, if possible:

- Feed, concentrate, and product conductivity for each stage.
- Feed, interstage, and concentrate pressure.
- Feed temperature.
- Feed pH, turbidity, and free chlorine concentration.

Conductivity is used as an estimate of concentration for real-time performance tracking. However, since performance is dependent on a number of factors, it is inadequate to track any single parameter. Instead, normalized permeate flow (NPF) and the AP across each pressure vessel are used to track performance. NPF is the permeate flow normalized for feed concentration, temperature, and applied transmembrane pressure.

Table 4.1.—Typical NF and RO performance specifications

Manufacturer:	A	B	B	B"	C
Process	Seawater RO	Brackish RO	Brackish RO	Brackish RO	NF
Configuration	Spiral Wound	Spiral Wound	Hollow Fine Fiber	Spiral Wound	Spiral Wound
Membrane polymer	Polyamide	Composite Polyamide	Aramid	Cellulose Acetate Blend	Composite Polyamide
Area	27.67 m^2 (300 ft^2)	37.16 m^2 (400 ft^2)	845.4 m^2 (9100 ft^2)	49 m^2 (528 ft^2)	37 m^2 (400 ft^2)
Minimum salt rejection	99.6%	99.0%	95.0%	98%	70% NaCl, 95% MgSO
Permeate flow	19 m^3/day (5 kgal/day)	45.4 m^3/day (12 kgal/day)	140 m^3/day (37 kgal/day)	39.7 m^3/day (10.5 kgal/day	47 m^3/day (12.5 kgal/day)
Max applied pressure	6.3 MPa (1200 lb/in)	4.16 MPa (600 lb/in^2)	4.14 MPa (600 lb/in*)	4.14 MPa (600 lb/in^2)	1.7 MPa (250 lb/in^2)
Max feed flow	Process dependent	264 L/min (75 gal/min)	176 L/min (46.5 gal/min)	137.8 L/min (36.4 gal/min)	265 L/min (70 gal/min)
Max operating temperature	45 °C (113 “F)	45 °C (113 “F)	40 °C (104 “F)	40 °C (104 “F)	35 °C (95” F)
Feed water pH range	4 - 11	3-10	4-11	4-6	3-9
Max feedwater turbidity	1.0 NTU	1.0 NTU	1 .O NTU	1 .O NTU	1 .O NTU
Max chlorine concentration	0	<0.1 mg/L	0	1.0 mg/L	<0.1 mg/L
Single element recovery	17%	15%	50%	20%	15%
Max pressure drop per element	Process dependent	69 kPa (10 lb/in)	69 kPa (10 lb/in^2)	104 kPa (15 lb/in*)	Process dependent

Data from manufacturer's literature.

$$NPF = F_p \frac{NDP_{today}}{NDP_{init}} TCF \qquad \text{Eq. 4.8}$$

$$NDP = P_a - P_o - P_p \qquad \text{Eq. 4.9}$$

Where:

F_p = product flow,
NDP_{init} = net driving pressure at startup,
NDP_{today} = current net driving pressure,
P_a = applied pressure,
P_p = osmotic plus the hydraulic back pressure of the permeate,
P_o = osmotic pressure of the feed water calculated as above,
TCF = temperature correction factor calculated from an equation supplied by the manufacturer. It has the effect of decreasing the NPF above 25 °C and increasing it below 25 °C. At 25 °C, the TCF is equal to one.

Whenever there is a 10 percent change in NPF or AP, or a significant change in any of the individual operating parameters, some corrective action, such as rinsing, cleaning, or system maintenance, is required. Sudden changes during normal operating mode usually indicate a mechanical problem with the system. Sudden increases in AP can indicate a blockage in one of the pipes or in the lead end element. Sudden decreases usually mean that one of the O-rings which prevent feed water from leaking around the membrane (and thus avoiding treatment) has deteriorated. In any case, it is something that needs to be addressed right away. In the author’s experience, most of the parameters that are monitored stay fairly constant until the system needs attention (backwashing media filters, replacing cartridge filters, replacing O-rings, etc.). Changes that indicate a problem with the membrane may be gradual, such as a slow creep in AP or steady increase in product flow. Table 4.2 lists the changes that can occur, possible causes, and solutions.

It is possible to spend ones entire life in fascination with the various ways RO membranes can be ruined; those listed in table 4.2 merely scratch the surface. More detail can be found in chapter 8 on membrane cleaning and preservation. Remember, pretreatment, close monitoring, and timely cleaning are the keys to longevity in membrane systems.

4.4.1 Effect of Operation Parameters on Performance

Membrane performance depends on the chemical make up of the feed water, but given a standard feed water, performance varies with pressure, water temperature, level of water recovery, and the oxidation potential of the feed water. The first three of these factors are

Table 4.2.—Troubleshooting guide for NF and RO systems

Symptom	Possible problems	Solutions
Sudden ↑ AP, no change in rejection.	Clogged cartridge filter Blocked pipes, or front end of membrane element	Replace cartridge filter, Check flows and front end of membrane.
Sudden ↓ AP, ↓ rejection	O-ring or brine seal failure, cracked permeate tube	Isolate point of failure by probing permeate tube with conductivity sensor, then replace defective part.
Gradual ı NPF 1st stage may be a slight ↑ rejection	Biological or particulate fouling	Clean 1st stage with high pH, high temperature solution.
Gradual I NPF 2nd stage with I 2nd stage rejection	Scaling	Clean 2nd stage with low pH solution at standard operating temperature.
Gradual ↑ NPF 1st stage with I 1st stage rejection	Lead end degradation caused by reaction of free chlorine with transition metals, or advanced biofouling	Check dechlorination system, check for sources of iron directly ahead of membrane system. Can sacrifice 1st element to protect rest of system. If problem is advanced biofouling, replace element and check prefiltration and disinfection system.
Gradual ↑ NPF 2nd stage, with ↓ 2nd stage rejection	Advanced scaling	Replace element and check pretreatment system. May need NF.

related to the feed water composition. The last is related to the material used in the membrane. Figure 4.3 illustrates how applied pressure, feed water temperature, and water recovery affect RO membrane flux and water quality. The graphs are generalized curves intended to show trends; they are not based on actual data points.

Effect of Pressure

The graph in figure 4.3 and the equation for "A" show that water flux is directly proportional to applied pressure. After a certain point, though, higher pressures cause problems that result in a decline in water flux. At high pressure, the feed water is forced against the membrane at a higher velocity, forcing a greater number of foulants in the feed stream to interact at the membrane surface. Material that might flow out of the system at lower pressure gets impacted onto the membrane surface and hung up in the spacer at higher pressure. The added layer causes a decline in water flux and may also aid in deterioration of the membrane material and provide protection to bacterial colonies.

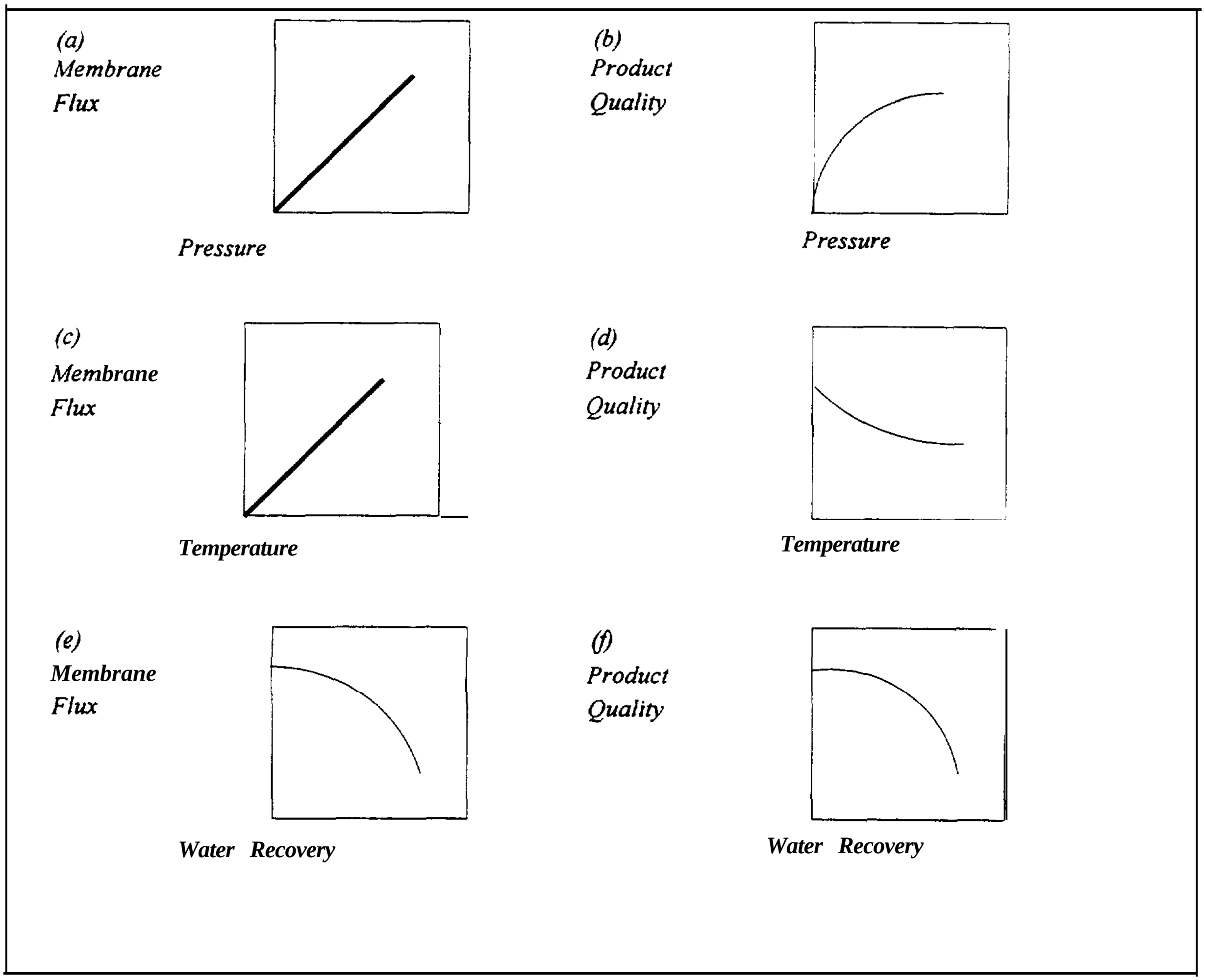


Figure 4.3.—The effects of applied pressure, feed temperature, and water recovery on membrane flux and product water quality of RO membranes. (From Riley, 1990.)

Scaling problems increase at high pressures. If the water flux is greater than the system is designed for, concentration polarization can cause precipitation of slightly soluble salts at the membrane surface. Concentration polarization is the normal phenomena of salt build up at the surface of a semi-permeable membrane. As the concentration builds, the local osmotic pressure increases. Under normal operating conditions, ions diffise away from the membrane surface and the bulk flow carries them out of the system. But at 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.

Membrane compaction and deformation are caused in part by excessive pressure. In the first stages of compaction, the membrane structure is compressed, restricting the flow of water and salt. The result is a decrease in water flux accompanied by an increase in product quality. If the condition continues, the membrane can be forced into the mesh of the product water carrier, causing minute tears to form (Kaakinen and Moody, 1985). An irreversible stage of compaction begins, characterized by an increase in water flux as well as an increase in salt passage. The time factor for compaction depends on other operating parameters. High temperatures, high pH, and the presence of oxidants will speed up the process.

Effect of Temperature

The effect of temperature on RO and NF membranes is the result of an increase in enthalpy of the system. Bonds within the membrane matrix are more relaxed, and salt molecules are more active at higher temperatures (≥25 “C). Water passes through the membrane with lower applied pressure than is required at lower temperatures. To a certain extent, salt flux is also increased as temperature is increased, but the effect is not important at normal operating temperatures.

Membrane manufacturers provide a table or formula for determining TCF (temperature correction factors). The TCF is equal to 1 .OO at 25 °C and is proportional to the change in pressure needed to maintain the 25 °C flux rate. Some manufacturers divide by the TCF to equate flux at one temperature to flux at 25 °C, in which case the TCF is less than 1 .OO for temperatures below 25 °C and greater than 1 .OO above 25 °C. Others choose to multiply flux by the TCF to correct for temperature so their TCFs are inverted. Just remember that at low temperatures the membrane flux will be lower, and to compare it to a high temperature flux, one must increase the low temperature flux value.

Most cellulosic and thin film composite membranes listed in chapter 10 have maximum temperature limits of 40 to 45 °C, which should be adequate for most surface and groundwater sources. Streams at higher ambient temperatures should be used in a heat transfer process before treatment. Excessive heat in the RO system can cause a variety of problems. Carbonate scaling is more likely at higher temperatures, membrane compaction is enhanced, and fouling caused by increased water flux is also more likely. Most membranes can handle feed water at temperatures as low as 1 °C without problem. The membrane matrix becomes more rigid at low temperatures and water flux decreases.

Effect of pH

The pH of the feed water can affect the membrane structure and the scale formation potential of the concentrate stream. Cellulosic membranes have a narrow operational pH range of 4 to 6. Some membranes have pH ranges as narrow as 5.5 to 6.0 (see tables 10.2 through 10.6). If exposed to a 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. Ceramics and metallics are unaffected by pH.

The pH of the feed water may need adjustment to control scaling of the concentrate conveyance system. For example, silica solubility increases dramatically above pH 7.7, and at higher temperatures. A silica scaling problem could be controlled by either raising the pH or the temperature of the feed water. Calcium carbonate, on the other hand, is more soluble at low temperatures and at a pH less than 8.0. A carbonate scaling problem can be relieved by lowering the pH, temperature or adding anti-scalants. However, if the concentrate is saturated in both silica and carbonate, changes in temperature or pH can cause one or the other to precipitate. Care must be taken to find the best condition to prevent scaling.

Tolerance to Oxidation

Oxidants are added to water supplies to control biological growth, to improve taste and odor, to remove iron and manganese, and to speed the decomposition of vegetable and animal matter (Weber, 1972). In the past, chlorine has been used very reliably. The CA membranes are chlorine tolerant. Non-cellulosic thin film composite membranes are not tolerant to oxidation; yet some method of biological treatment is still needed. Systems that use chlorine with thin film composite membranes require dechlorination just ahead of the RO unit.

Ozone, an aggressive oxidant, is useful for color removal, taste and odor removal, disinfection, iron and manganese removal, phenol oxidation, and cyanide oxidation (Weber, 1972). Ozone is bubbled into the feed water allowing sufficient time for reaction. Ultraviolet light is used to remove excess ozone before the feed water enters the membrane modules (Zoccolante, 1990). UV irradiation also can be used as a disinfectant, but is only active while the water is exposed to the light. Organisms that survive exposure can recontaminate the system. When used with ozone, however, irradiation has a synergistic effect.

4.5 Post-treatment

Water produced from RO or NF processes is not exceptionally good for drinking. Depending on the level of desalination, it may be corrosive to distribution lines. It is necessary to add back some of the hardness by adding lime, which is another reason for using NF if possible. Adding lime will also raise the pH, which is necessary if CA membranes were used. Water is practically de-aerated in the RO and NF processes, so some aeration is needed to improve flavor. Finally, chlorine is added to maintain the required residual disinfection prior to distribution.

4.6 Applications

The use of RO membranes on a large scale for water treatment originally began in the 1960s using CA membrane to remove salts from brackish water or seawater. The applications evident in the 1990s include the traditional potable water applications, industrial and waste water treatment, boiler feed water purification, irrigation return flow remediation, and production of ultrapure water for the pharmaceutical and electronics industry. Commercial applications have grown in diversity as listed in table 4.3.

Several manufacturers in the U.S. were surveyed to find out how their membranes are being used. A compilation of two of the top manufacturers (Dupont and Fluid Systems) indicated that the they sold close to 5 1 percent of the systems to prepare potable water and 49 percent of the systems for “other purposes.”

The information provided by these two companies has been compiled and organized in table 4.4 and figure 4.4. A complete list of the information is provided in appendix A. This is not intended as an extensive survey of manufacturers; however, this listing of over

Table 4.3.—RO and NF applications

Application	Purpose
Potable water	
Municipalities, cities, and towns	Drinking water, bathing, laundry
Resorts, hotels, and motels	Drinking water, bathing, laundry
Offshore drilling and production platforms	Drinking water, bathing, laundry
Mobile home parks	Drinking water, bathing, laundry
Realty developments	Drinking water, bathing, laundry
Island communities	Drinking water, bathing, laundry
Industrial applications	
Rinsing electronic components	High quality, uncontaminated components
Boiler makeup	Increases heat transfer by reducing scaling
Process water	Reduces waste, improves quality of process
High-purity water	Formulations, rinsing and cleaning metals, polymerization reactions
Electrodeposition paint processing	Recycle stream reduces waste water
Special applications	
Pharmaceuticals	Create a pure product
Medical	Hemodialysis, separating proteins
Concentrating cheese whey and juice	dehydration
Laboratories	Reduce TDS
Car washes	Prevent spotting
Small cooling towers	Reduce contaminant
Humidifiers	Prevent scale
Flower growers	Better yields and quality
Water bottlers	Reduce salinity
Ice rinks	Smoother ice
Fish farms and hydroponic farming	Consistent quality
Photographic wastes	
Campgrounds	Potable
Soft (bottled) drink manufacturers	Remove organics, better taste
Ice manufacturers	Reduce TDS
Private home systems	Improve quality
Landscaping gardens, golf courses	Reduce TDS, improved plant growth
Removes alcohol	Preserves flavor

Source: Dupont's Guide to Products and Properties, Amjad, 1993.

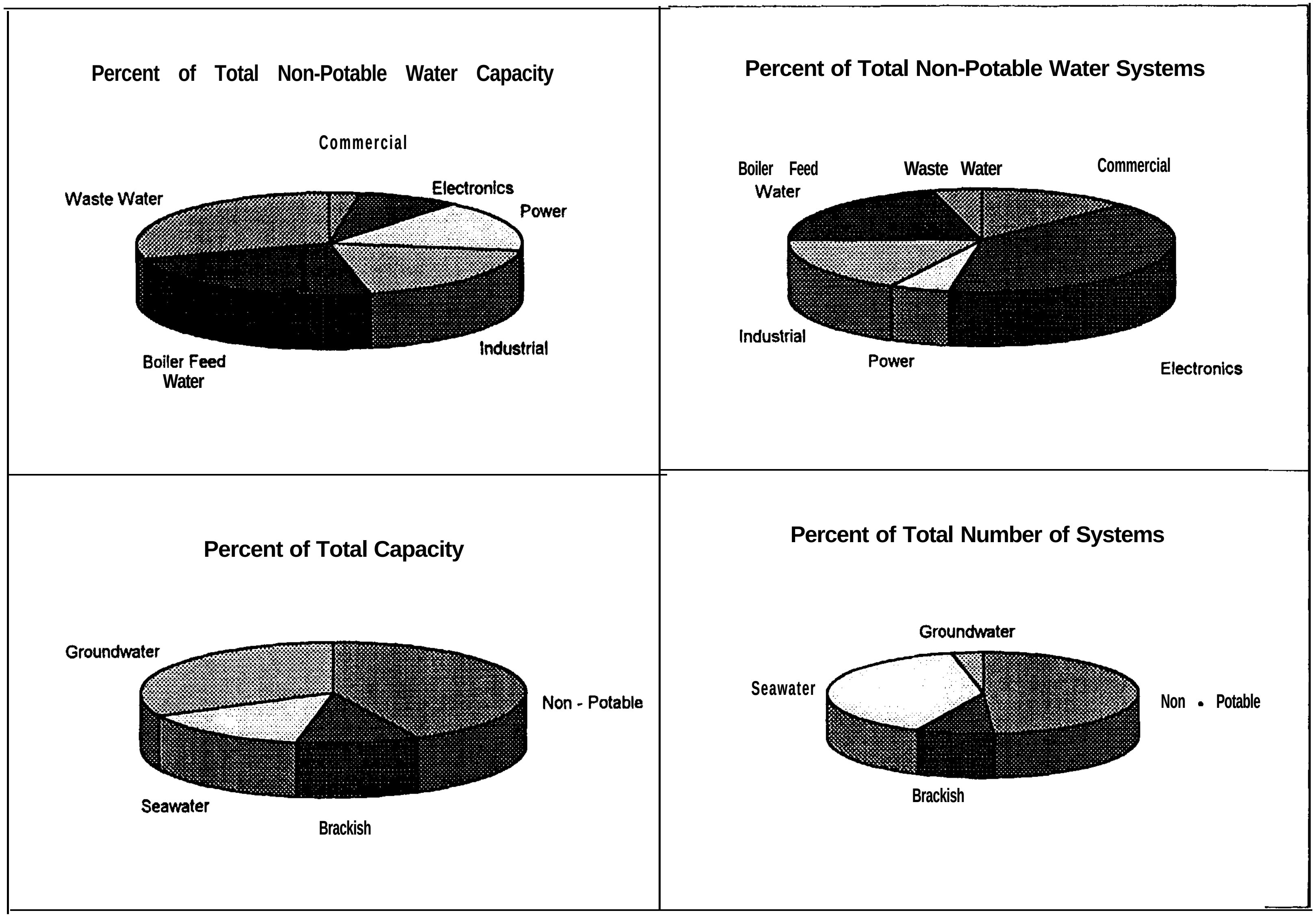
Table 4.4 Capacities and number of systems

Application	Average capacity (1,000 gal/d)	Number of systems	Capacity (% of total)	Number of systems (% of total)
Other applications:				
Commercial	99	21	1.1	6.4
Electronics	330	64	3.7	19.4
Power	609	8	6.9	2.4
Industrial	722	28	8.2	8.5
Boiler feed water	890	33	10.1	10.0
Waste water	1,119	7	12.7	2.1
Subtotal			42.7	48.8
Potable water:				
Potable - Brackish	910	27	10.3	8.2
Potable - Seawater	1,311	131	14.8	39.7
Potable - Groundwater	2,850	11	32.2	3.3
Subtotal			57.3	51.2
Total	8,840	330	100.0	100.0

300 installed applications is assumed to be a reflection of the variety of applications for these processes. Many smaller firms exist and may be seeing different trends in their sales. More detailed information which fully analyzes the world market may be available from the Wangnick Consulting report from Germany (listed in the bibliography).

Several broad conclusions can be made regarding this compilation. Three of the top four capacity applications are associated with the production of potable water. Treatment of groundwater uses the highest amount of RO membranes at 32 percent of the total capacity. After potable water, the largest applications are wastewater, boiler feed, industrial, power, electronics, and commercial applications. Potable sea water had the largest number of systems put in place with nearly 40 percent of all the systems. Next was the electronics field at 19 percent, then boiler feed water at 10 percent.

Manufacturers are sorting out the market niches for RO membranes. One manufacturer has dropped development of new applications and no longer sells spiral wound CA membranes. They now concentrate on their most profitable hollow fine fiber membranes for sea or brackish water purification sold primarily in Europe and the Middle East. Another manufacturer indicated that the current driving force of expansion in the industry was probably municipal water treatment.



Percent of Total Non-Potable Water Capacity

Percent of Total Non-Potable Water Systems

Percent of Total Capacity

Percent of Total Number of Systems

Figure 4.4.—Distribution of membrane applications.

The other applications serve as a flourishing testing ground for new approaches to problems. A review of the recent articles in the Chemical Abstracts indicates that many countries, in particular India, Korea, and Japan, are highly involved in RO research. Surveying the literature provided by the manufacturers points out that global application of this technology was strong in the early 1980s and continues to expand throughout the 1990s.

4.7 NF or RO?

The RO and NF membranes available today are very similar. It seems that, rather than marketing "RO" and "NF" membranes, manufacturers are producing different membranes to cover a wide range of rejection characteristics. For this discussion, any membrane with greater than 95 percent rejection of sodium chloride is considered an "RO" membrane, those with between 45 and 95 percent rejection of sodium chloride are considered to be "NF" membranes. So, how does one choose which to use for a particular situation? The following questions can be answered with a good water analysis and will help with the decision.

- **What is the TDS?** If the TDS is over about 1,500 mg/L, RO will probably be best.

- **What is the target product water quality?** If a reduction in TDS greater than ~ 90 percent is needed, RO will be necessary.

- **What is the percentage of multi-valent ions?** If the TDS is made up of mostly multivalent ions, NF may be best.

- **Which constituents exceed primary or secondary drinking water standards?** RO permeate can often be blended with water from another source, but if there are contaminants that will exceed drinking water standards even when blended with RO permeate, the cost of treating the blend water must be considered. If the cost is high and/or there is a high percentage of multivalent ions, then NF may be a better choice.

- **Are there size and cost restrictions?** Systems can be designed to minimize any parameter, but there are tradeoffs.

- **What are the concentrate disposal options** ? In some locations with only marginal water quality, surface disposal may be possible if the concentrate TDS is not too high. What is "too high" depends on the local soils, amount of precipitation, and type of plants to be grown in the area. In other locations, concentrate volume will be the most important factor.

The options for tailoring a membrane system to meet the specific objectives for cost and concentrate stream characteristics at a fictitious site are explored in figure 4.5. Salinity of the feed water is 1800 mg/L TDS, 28 percent of which is multi-valent. The source water also exceeds the secondary drinking water limit for manganese (0.05 mg/L) by 0.5 mg/L. NF membranes with three different rejection characteristics are compared with RO using three

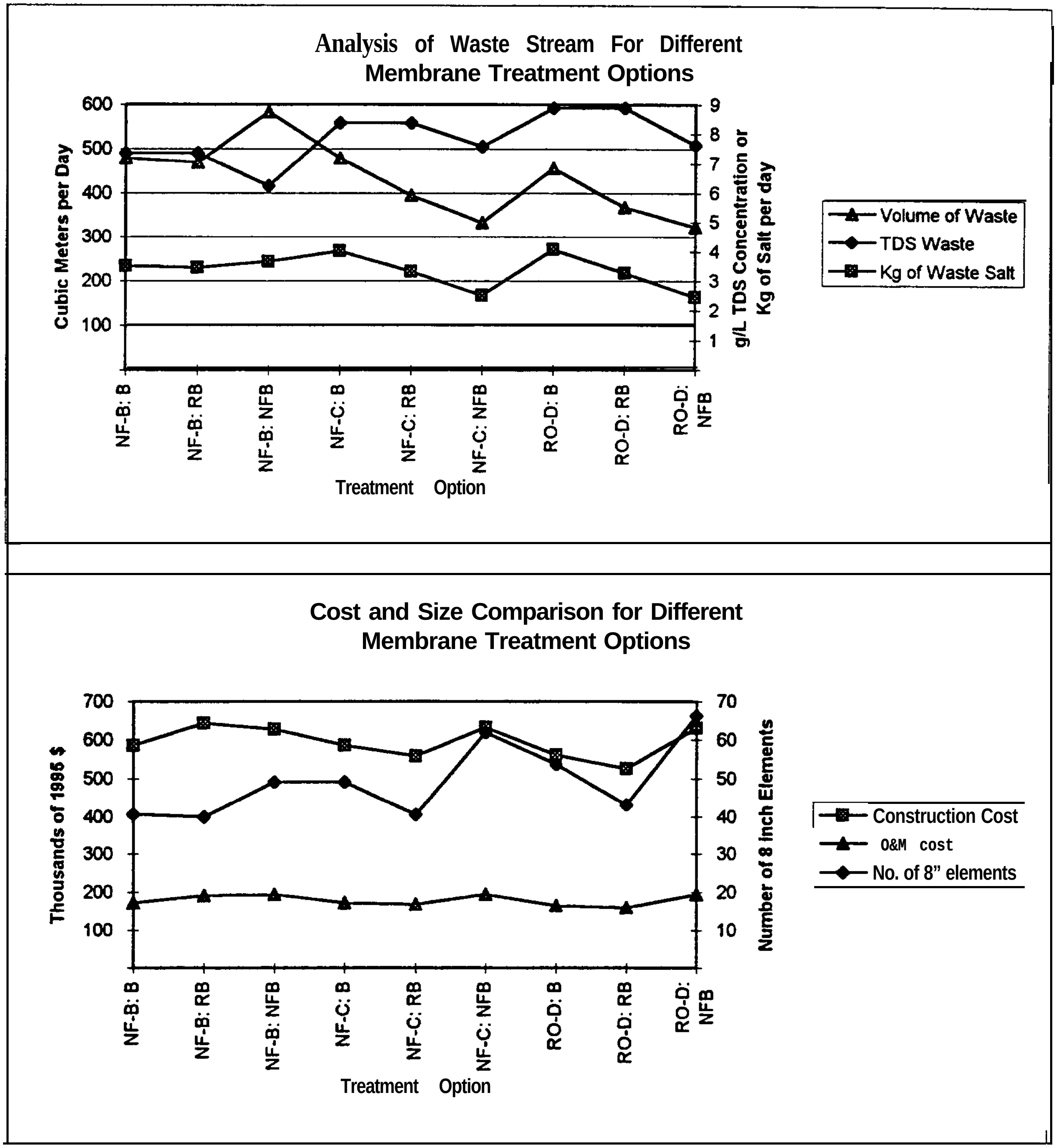


Figure 4.5.—Comparison of cost statistics for various membrane treatment options. Costs are based on producing 2,000 m^3/day with a target TDS of 500 mg/L. The raw water is assumed to have 1,800 mg/L TDS, 28 percent multi-valent, with 0.55 mg/L Mn (secondary standard is 0.05 mg/L). B is blend with raw water. RB is remediated blend, meaning the Mn in the blend water has been removed using green sand. NFB is blend with NF-A product water which would have 788 mg/L TDS and less than 0.05 mg/L Mn.

different blending options to assess differences in the concentrate stream volume, concentration, total mass of salt produced, construction cost, O&M cost, and plant size. The membrane characteristics used in this comparison are listed in table 4.5.

Table 4.5.—Membrane characteristics used in treatment option comparison. Assumptions: 80 percent recovery; feed water has 1800 mg/L TDS; 28 percent multi-valent ions, 0.55 mg/L manganese

	NF-A	NF-B	NF-C	RO-D
Element productivity m^3/day	28.3	47	39	34
Mono-valent rejection	40%	70%[1]	90%[1]	98%[1]
Multi-valent rejection	98%	95%	95%	99.5%

[1] Rejections are estimates based on laboratory tests.

The options of using selective NF membranes for treatment of water to be blended with product from membrane NF-C, or RO-D, produce the lowest total concentrate volume. If the concentrate stream had to be disposed of in an evaporation pond or brine concentrator, low volume would be an important consideration because it controls the cost of disposal. However, these two options also require the largest number of elements and are among the most expensive to build.

Because of the low blend ratio, membrane NF-B with NF-A blend option has the lowest concentrate TDS, but the highest volume. In this scenario, the product of membrane NF-B has a concentration of about 400 mg/L, just slightly below the standard of 500 mg/L. Treatment of the blend water with membrane NF-A allows a 23 percent blend rate, but then the concentrate stream is augmented with that of the NF-A system which is also operated at 80 percent recovery. However, if the concentrate stream TDS is low enough that it can be used for irrigation, swimming pools, or discharged to surface water, the higher construction cost and larger plant size would be out weighed by the lower disposal cost.

4.7.1 RO and NF Cost Comparison

There are two key interrelated cost controlling parameters for RO and NF. Both of these, the operating pressure and the membrane capacity, are dependent on the difference in osmotic pressure (or TDS) between the feed and product water. Figures 4.6- 4.8 show the change in cost for three types of NF and brackish water RO treatment systems with increasing salinity and volume. Figure 4.6 shows change in cost with total capacity, including blend volume. Feed water is assumed to have a TDS of 1,800 mg/L for all membrane types except NF-A, which is assumed to have 1,100 mg/L. NF-A would not be able to produce the target TDS of 500 mg/L from such a high salinity water. Capacity is directly related to membrane area, which explains the linear aspect of the capacity-cost relationship. The hierarchy of the curves is a function of blend ratio. If costs for plants with equivalent membrane area were compared, the hierarchy would be determined by operating pressure.

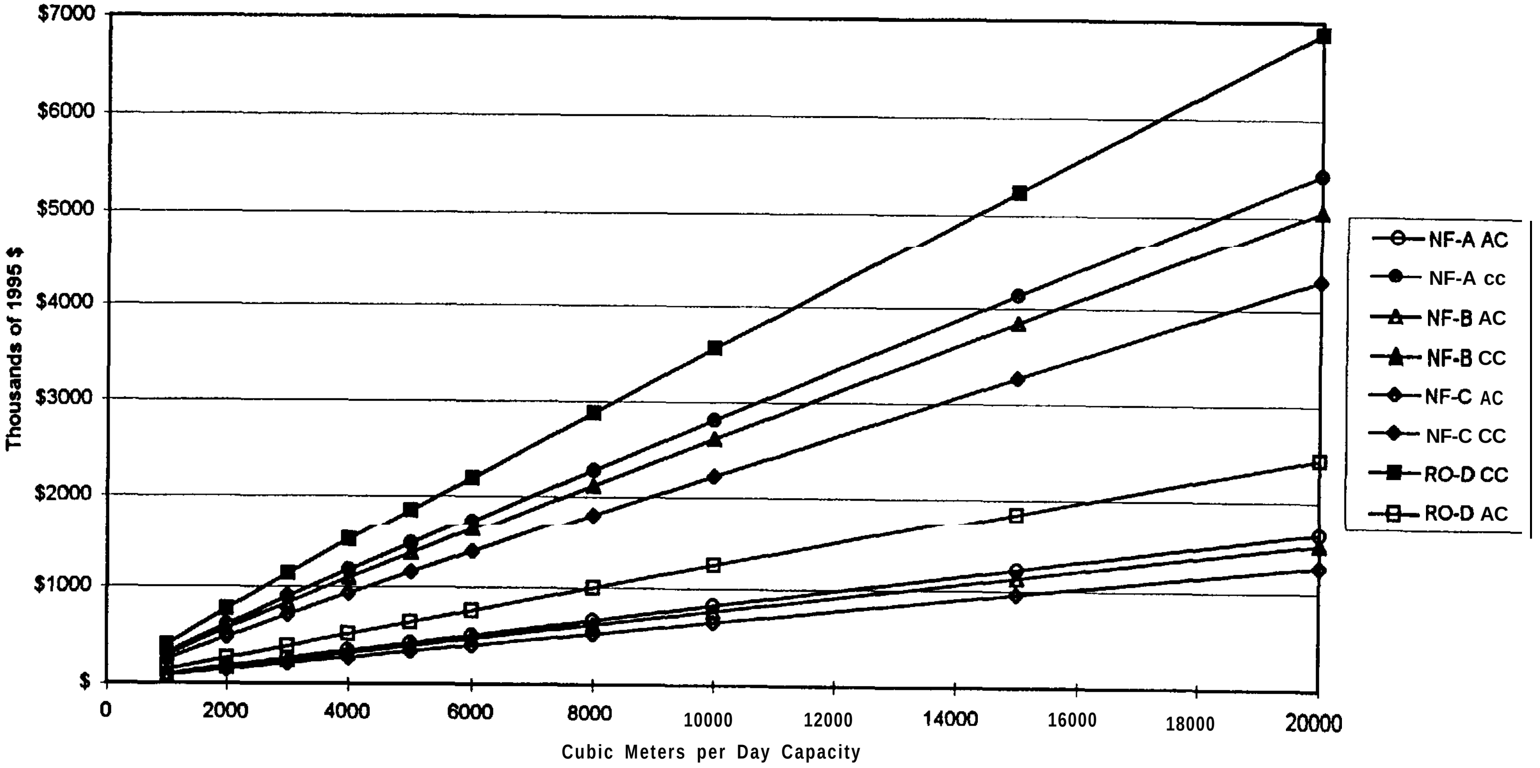


Figure 4.6.—Costs for plants to produce 500 mg/L TDS drinking water including blend water. Multi-valent ion concentration is 28 percent. Interest rate is 7 percent for 15 years. Plant is assumed to operate 85 percent of the year. Brackish water RO costs jump at 1,100 and 2,700 mg/L because the high rejection rate produces a large osmotic pressure differential which must be overcome with a higher operating pressure.

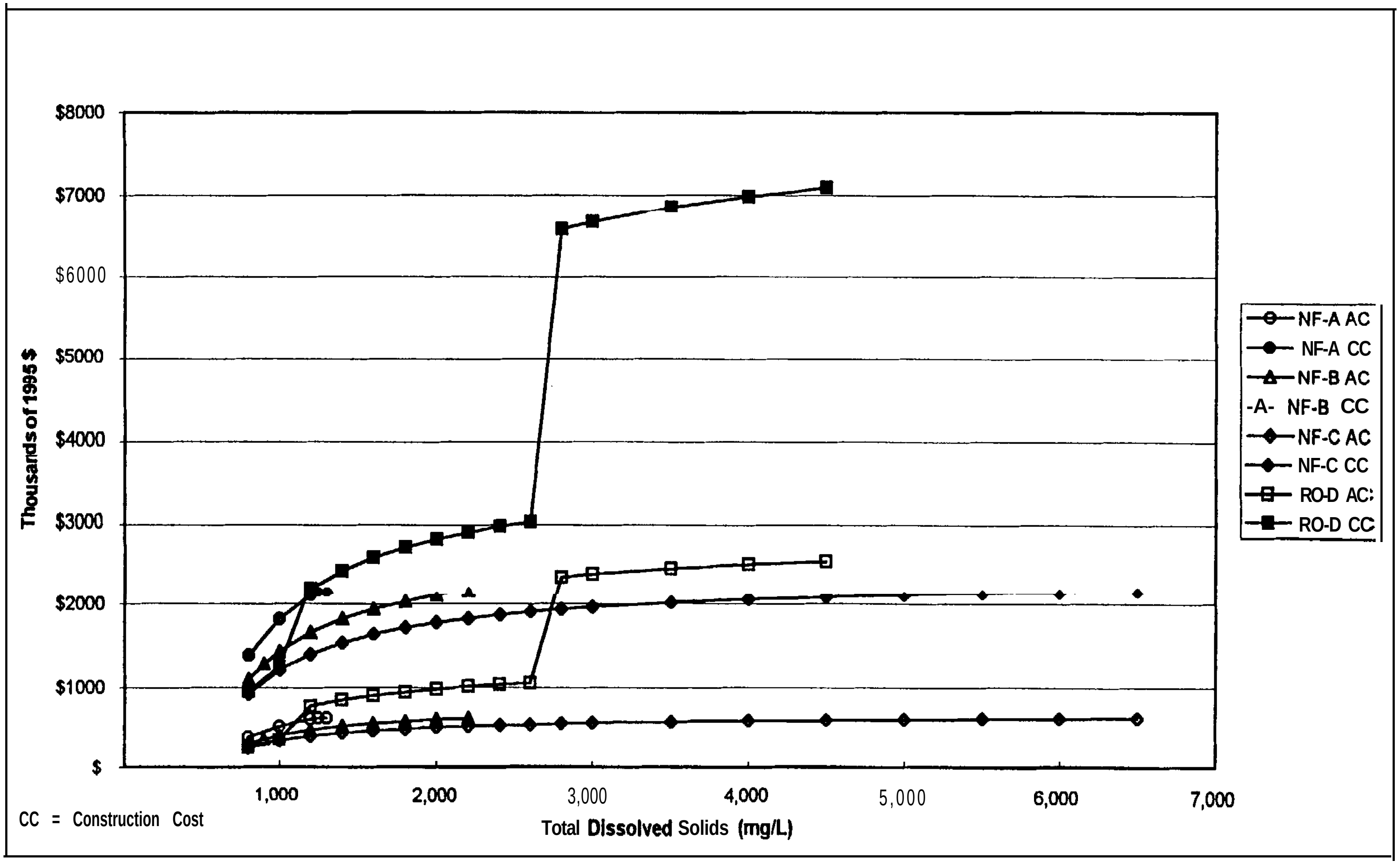


Figure **4.7.—Costs** for plants to produce 7,569 m^3/day including blend water. Multi-valent ion concentration is 28 percent. Interest rate is 7 percent for 15 years. Plant is assumed to operate 85 percent of the year. Brackish water RO costs jump at 1,100 and 2,700 mg/L because the high rejection rate produces a large osmotic pressure differential which must be overcome with a higher operating pressure.

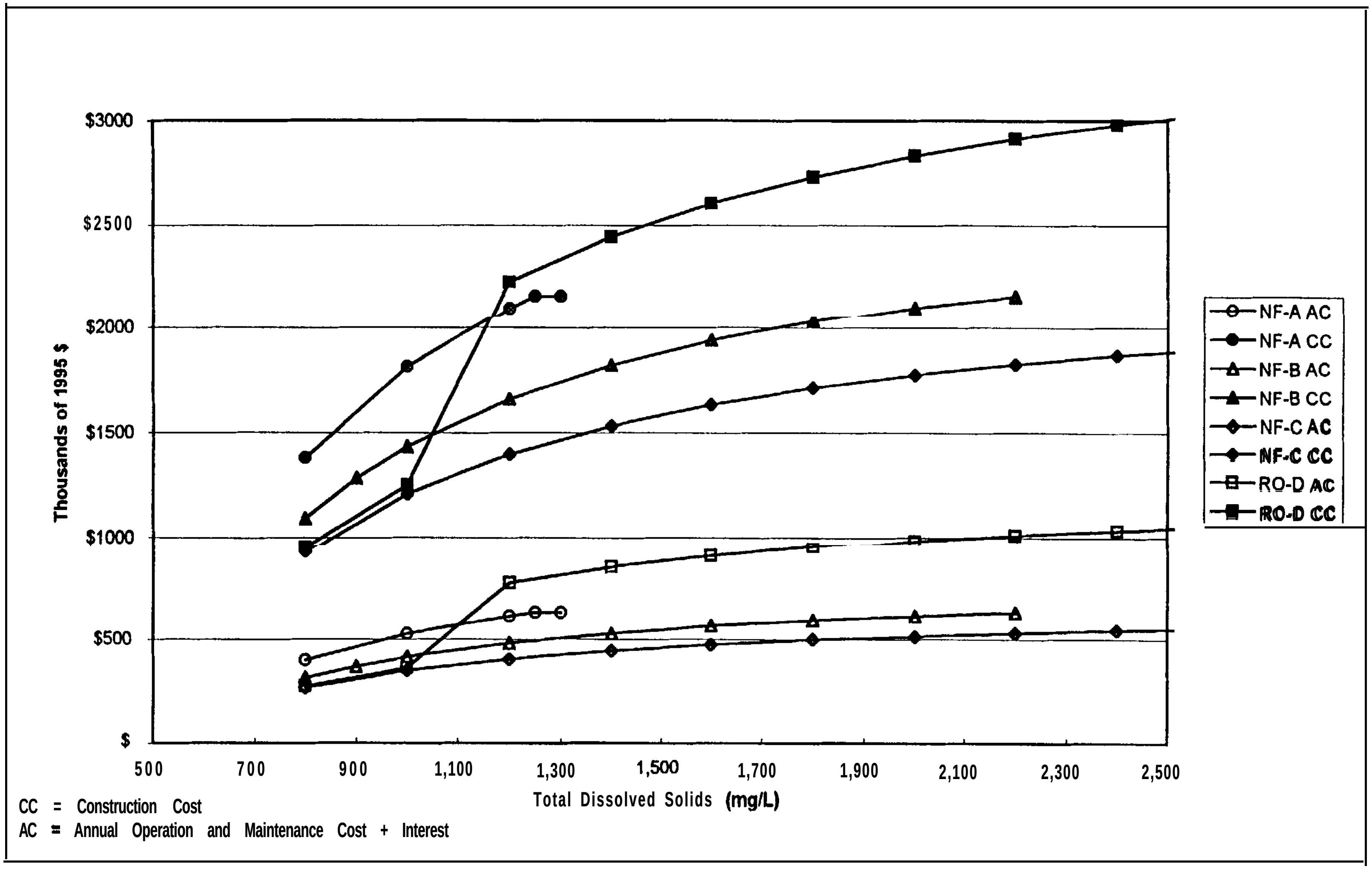


Figure 4.8.—Costs for plants to produce 7,569 m^3/day including blend water. Multi-valent ion concentration is 28 percent. Interest rate is 7 percent for 15 years. Plant is assumed to operate 85 percent of the year. Brackish water RO costs jump at 1,100 mg/L because the high rejection rate produces a large osmotic pressure differential which must be overcome with a higher operating pressure.

In both figures 4.7 and 4.8, costs are calculated to produce 7,569 m^3 per day (2 million gallons per day) assuming that TDS is the only problem with the source water and that the product can be blended back with raw water to achieve 500 mg/L TDS. In figure 4.7, the curves level out at the point where the TDS is too high to allow blending. Notice that the RO curves take two sudden jumps at about 1 ,100 and 2,600 mg/L. This is because of the additional equipment costs involved with operating at higher pressures, such as higher quality pumps and high pressure piping. The NF membrane pressures do not exceed 1,700 kPa (about 250 psi). At low concentrations, shown in greater detail in figure 4.8, the NF-C still has a slight advantage over RO because of the lower operating pressure.

One might wonder why the lower rejection NF membranes would be used if they are more expensive. Concentrate disposal cost is not accounted for in this cost comparison, but as discussed above, NF-A concentrate is more likely to be acceptable for surface discharge or application than RO concentrate. Also, when there is the potential for scaling, RO feed water would need more pretreatment than if a selective membrane, such as the NF-A, were to be used. To bring the total cost of desalting treatment down, scenarios such as these need to be analyzed thoroughly. There is a limit to the economies available if each process is evaluated separately.

But when the complete treatment process is evaluated as a whole, from pretreatment to concentrate disposal, there may be large economies that arise from having the flexibility to choose a design that fits the disposal options available.

4.8 Research Needs in NF and RO

- Development of more robust systems. Improvements in water transport and salt rejection are meaningless if they are accompanied by increased fouling potential. The newer thin film composite membranes do have superior transport qualities but they are extremely sensitive to fouling and difficult to clean (Chapman Wilbert, 1997). A new design is needed that promotes self-maintenance of the membranes, perhaps an improved spacer, and/or membrane material that has a lower fouling potential.

- A new method of dealing with biological fouling is needed. If organisms cannot be kept from the feed stream, there will be biological fouling that will be difficult, if not impossible, to remove without destroying the membrane. There are two possible approaches to this problem: either find a way to keep the biofilms permeable so that they cause a minimal increase in resistance to water transport, or find a way to induce the biofilms inhabitants to leave on their own. Progress is being made in both of these areas. Hopefully, it won’t be too long before research bears fruit in application.

- Concentrate utilization is urgently needed. There must be beneficial uses for concentrated salt solutions. When people spend $30.00 or more for 10 oz. of Salt Lake mineral supplements it is hard to believe that RO concentrate is useless. It may take selectively removing some components and adding others. We need to get some nutritionists working on this problem.

Chapter 5
ELECTRODIALYSIS

5.1 Definition

Electrodialysis, ED for short, can be described, in the most general terms, as a process that moves ions (charged molecular species) from one solution to another, employing an electrical potential as the driving force and using some sort of semipermeable membrane as a separator. This process can be used for concentration, dilution, or segregation of ionic species. The classical reference on electrodialysis of water, for breadth of approach and depth of research, is the book by Wilson (Wilson, 1960).

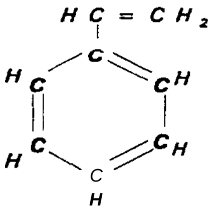
5.2 Background

A primitive form of electrodialysis has been used for salt removal from solutions since the beginning of the twentieth century, primarily in the laboratory. With the invention of the synthetic ion-exchange membrane in 1948, it became a practical means for water desalination. Electrodialysis was publicly demonstrated in 1952 (Lawrence, 1952). The introduction of current reversal electrodialysis in the early 1970s made operation with no chemical addition possible.

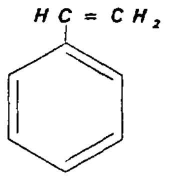
5.3 Membranes

The synthetic ion-exchange membrane contains fixed charges usually of only one sort (positive or negative) attached to a polymer backbone. Because of the fixed charges, the material is usually highly hydrated. The membrane has pores of molecular dimensions, undoubtedly irregular in shape and size, bounded by fixed charges of one sort filled with water containing the mobile ions having a charge opposite to that of the fixed charges. In the membrane, ions of opposite charge to the fixed charges are called “counterions” and ions of the same charge as the fixed charges are called “coions.” There is a low concentration of coions in an ion-exchange membrane.

One material commonly used in making ion-exchange materials is styrene, C_8H_8, which has the following structural formula:

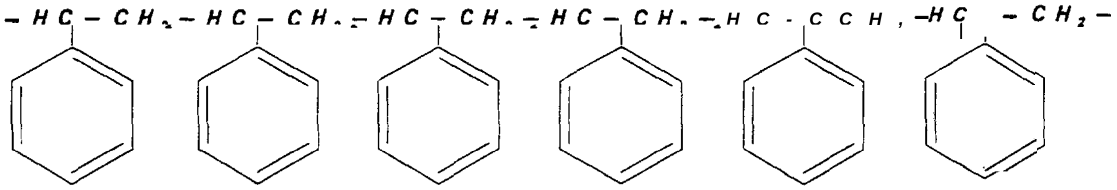


Styrene: or more



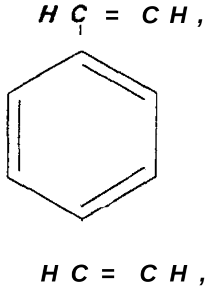
simply:

Styrene is a commonly available organic compound, currently available at a bulk price of less than $0.40 per pound. It is a pungent, volatile aromatic liquid at room temperature. Under fairly gentle conditions, i.e., at temperatures slightly above room temperature and at room pressure, and with a suitable catalyst, this can be converted to a linear polymer of high molecular weight. The linear polymer, a thermoplastic, has the structure like a chain of beads. The chain is drawn as a straight line for convenience. Subject to steric limitations, fairly free rotation around the single carbon-carbon bonds is expected.



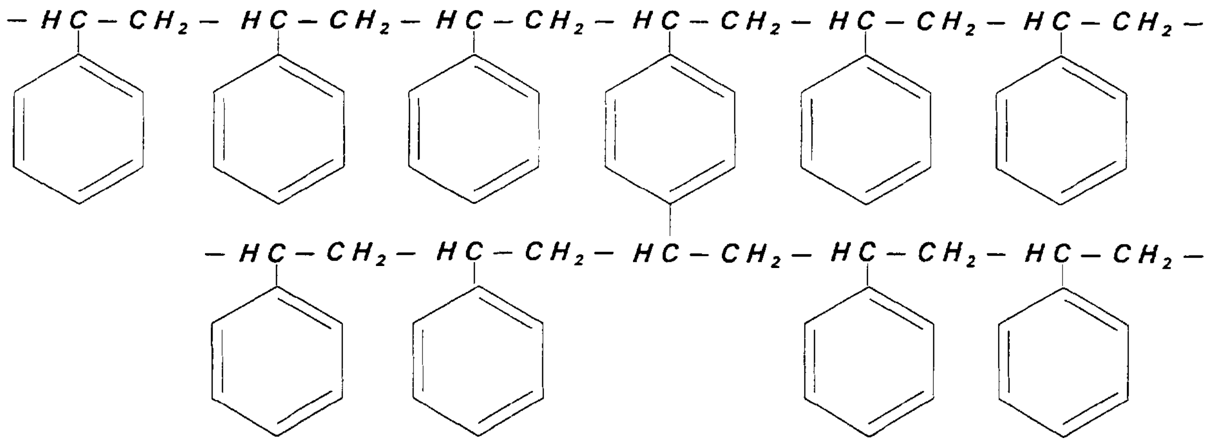
Divinyl benzene, $C_{10}H_{10}$, a fairly common organic compound, is similar to styrene but it contains a second vinyl group attached to the ring. It has three isomers, the **para** isomer is shown.

Divinyl



benzene:

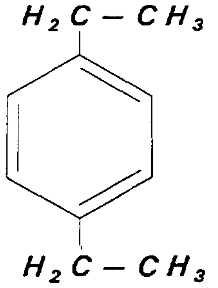
This is used as a cross-linking agent for styrene-based polymers. The sketch below shows two polystyrene chains with a single crosslink.



Crosslinking makes the polymer thermosetting, causing hardness in the polymer and decreased solubility, depending on the mol ratio of divinyl benzene to styrene.

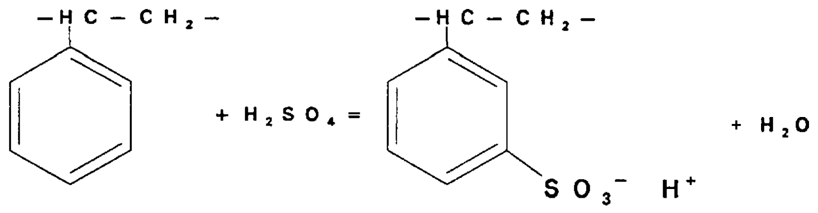
A third compound is normally added to the chemical mixture before polymerization to improve this polymer for use as an ion-exchange membrane, a material soluble in the styrene/divinyl benzene mixture, but incapable of polymerization. Such a material is diethyl benzene, $C_{10}H_{14}$.

Diethyl benzene:



This will cause the resulting polymeric network to be more open and will increase the water content of the finished membrane.

A sulfonate group can be added to a benzene ring of the polymerized crosslinked polystyrene by exposing it to strong sulfuric acid.



Typically, there will be slightly less than one sulfonate group per benzene ring. This reaction supplies a fixed negative charge to the polymer and converts a piece of plastic into an ion-exchange material. When equilibrated in water, the material, though electrically neutral, has mobile positive charges and fixed negative charges. Note that the hydrogen ion is not associated with a particular sulfonate group but is reasonably free to wander about in the aqueous phase of the membrane. Ions in the pores of the membrane can, subject to quantifiable rules of mass transport, exchange with ions in solutions surrounding the membrane. Of particular importance to ED, motion of ions is influenced by electrical potentials applied across the membrane.

In a manner analogous to the creation of this membrane, the polymerized polystyrene can be treated with a tertiary amine to produce a membrane with a fixed positive charge.

Strength and flexibility are gained by casting this polymer mixture on a fabric backing. By such means, one obtains an ion exchange material in sheet form suitable for use in electrodialysis. The resulting product can be called a "cation-passing," "cation-exchange," "cation-transfer," or, simply, "cation" membrane. All these terminologies are used in the literature and in the field. Curiously, since the membranes are named after the mobile species, a cation membrane is actually a polyanion.

High conductivity ion-exchange membranes usually have a high water content. They are translucent and generally tan in color, somewhere between ivory and brown. When exposed to dry air, high water content membranes lose water and can lose physical integrity. Under this condition, the membrane turns opaque due to small cracks in the polymer structure.

A variety of other approaches can be used to generate ion-exchange membranes. In addition to the procedure described above, "homogeneous" membranes can also be produced by sorbing monomer, like styrene, into existing polymer films, such as polyethylene. "Heterogeneous" membranes can be made by using a binder to adhere finely divided ion-exchange particles to each other and, if desired, to a fabric. Kesting (197 1, chapter 7) describes a number of approaches to manufacture ion-exchange membranes.

5.4 Membrane Properties

A number of properties of ion-exchange membranes can be measured. Those of interest to the electrodialysis process are resistance, permselectivity, and flexural or burst strength.

Electrical resistance in a membrane is analogous to electrical resistance in an electronic conductor. It is the ratio of potential drop to current passed. The difference is that in the membrane, current is being carried by ionic species rather than by electrons.

Permselectivity is the ratio of the current carried by the mobile, or counterions, to the total current. While it might appear that all of the current should be carried by the counterions, there is usually a small fraction which is carried by ions of the same charge as the fixed charges in the membrane. For a given membrane, permselectivity may be significantly affected by the concentration and ionic composition of the solutions in which the membrane is immersed. Temperature and pressure, in the range used in water desalination, have little effect on it.

Flexural strength is a concern only when one looks at use of the membrane in ED equipment, described below. Typically, a membrane is optimized in terms of polymer/crosslinker/non-polymerizable ratios to give good transport properties. Then it is left to the equipment engineer to design a system which can contain and support the resulting membrane. Because the pressure applied across a membrane in an ED system is very modest, membranes are rarely taken to the point of bursting. However, the burst strength test is a useful means of quality control on membrane manufacture and a convenient measure of membrane degradation in use. Lior (1986, chapter 2.7) provides details of performing these tests.

5.5 Electrodialysis Process

The basis of the process is straightforward.[1] When dissolved in water, salts, acids, and bases dissociate into ions. An electrical field can be produced across a solution by immersing a pair of electrodes into an ionic solution in a container or cell and applying a direct current potential difference to the electrodes. This produces a general motion of ions. As indicated below, negatively charged anions, like chloride ions (Cl-) or hydroxide ions (OH-), migrate toward the anode, and positively charged cations, like sodium ions or magnesium ions, migrate toward the cathode. In a solution of sodium chloride, about half the current is carried by cations, migrating to the left in the diagram, and half by anions migrating in the opposite direction. This discussion focuses on the motion of the ions. What occurs at the electrodes will be covered later.

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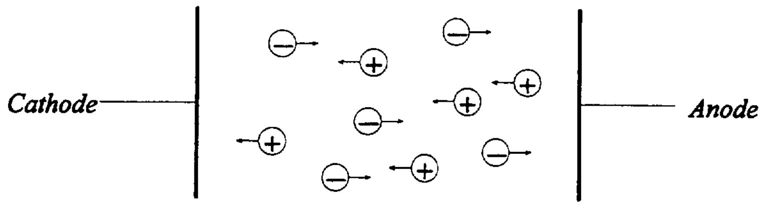
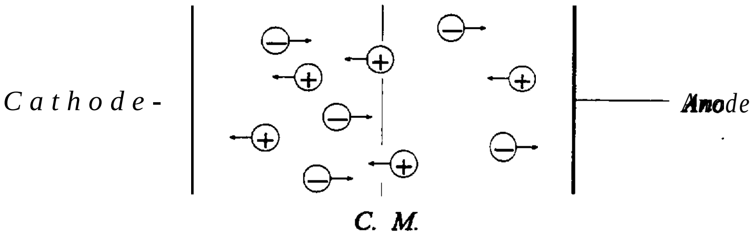
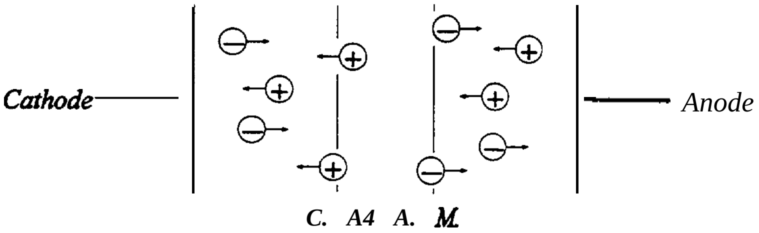
[1] To avoid a frequent repetition of "usually," "often," or "generally," the text below is written to express the common phenomena. The author recognizes that there are frequent exceptions.



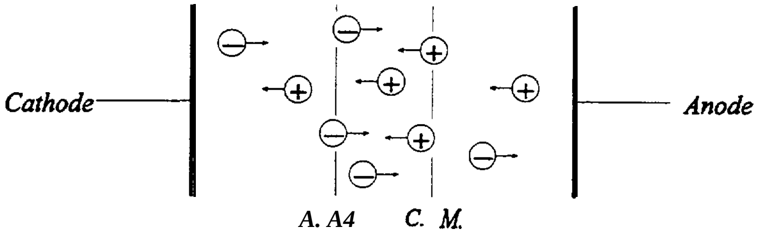
In such a solution, an ion-exchange membrane acts as a one-way street, permitting migration of ions of only one charge. If a cation membrane is inserted between the electrodes, one will observe a depletion of cations (and thus of salt) on the side facing the anode and an increase of cations on the side facing the cathode.



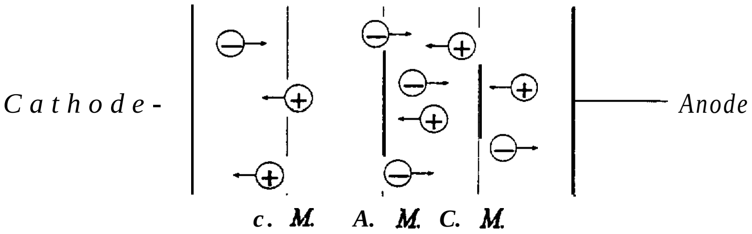
If an anion membrane is inserted in the cell between the cation membrane and the anode, one will observe a general depletion of salt in the compartment between the membranes.



If the membrane positions are reversed with respect to the electrodes, passage of current will cause an increase in concentration of salt within the compartment.



In an alternating series of membranes, three membranes are shown below, we find a diluting compartment next to a concentrating compartment. The grouping consisting of a cation membrane, a diluting compartment, an anion membrane and a concentrating compartment is the fundamental building block, termed a "cell pair," of which an electrodialysis stack is made.



Conceptually, it is only a small step from one cell pair, as shown above, to the large number of cell pairs between one pair of electrodes which comprise an electrodialysis stack.

At the electrodes, electrochemical reactions occur. These reactions change the current **from** a flow of electrons to a flow of ions. The most common reactions are the following:

At the cathode: $2H^+ + 2e^- \rightarrow H_2\ (g)$

At the anode: $2OH^- \rightarrow H_2O + \frac{1}{2}O_2\ (g) + 2e^-$

or $2Cl^- \rightarrow Cl,\ (g) + 2e^-$

The effect of the electrode reactions is minimized by using a large number of cell pairs between one pair of electrodes. However, a certain amount of gas is evolved that must be disposed of.

5.6 Electrodialysis Apparatus

Commercial electrodialysis apparatus is operated on a continuous flow basis. The description above is for a batch cell, but similar principles apply to continuous apparatus. Mason and Kirkham (1959) provide a very thorough analytical treatment of electrodialysis as a process. Another good general reference on electrodialysis is the article by Solt (Meares, 1976). See also (Li, 1972).

A variety of commercial systems are available. The physical structure of the equipment is quite varied. These are based on membranes having an area of 0.5 to 1.5 square meters. The compartments between the membranes are formed by gaskets or spacers between 0.5 and 1 mm thick. These gaskets support and separate the membranes and control and direct the flow of water through the compartments. As many as 600 cell pairs may be placed between a pair of electrodes. When the entire assemblage is placed between a pair of compression blocks, the result is the basic building block of an electrodialysis system, usually termed a "stack." For no obvious reason, the membranes in American electrodialysis stacks are oriented horizontally and those in stacks made in Japan are oriented vertically. The residence time of water in the stack is a small number of seconds.

Some structural details may be of interest. Water is fed into and removed from the compartments by holes cut in the membranes and the frames of the spacer which are aligned to form manifolds, In the areas which connect the open area of the compartment to the appropriate manifold, the membranes need to be supported to prevent leakage from the manifold to the alternate compartment. Two general forms of spacers are in use. The "sheet flow" spacer is an open frame containing a net like material which allows passage of water parallel to the membrane surfaces and passage of current normal to the membrane surfaces. Fluid flow is essentially parallel to the long dimension of the stack. The other sort of spacer is the "tortuous path" spacer in which the flow path bends back and forth on itself yielding a path length which is about four to eight times the length of the longer dimension of the membrane. This type of spacer contains "turbulence promoters" which are half the thickness of the spacer in height and designed to promote mixing at the membrane surface. Because they are chemically more stable, cation membranes are used next to both the cathode and the anode. A more robust, i.e., thicker, membrane is used next to the electrodes.

While it is not the purpose of this paper to provide sufficient detail to allow the reader to design an electrodialysis system, some quantitative relationships may be useful. A large number of variables must be set in design of an ED plant, which is best done by the manufacturer of the equipment. The following estimating equations are given to show the relationship of required area and power for a given desalination. The cell pair area, A_P, is given by

$$A, \cong \frac{C_{di} \mathscr{F} R_{po} f_o F_d}{V_p e} \qquad \text{Eq. 5.1}$$

and the power, P, in watts, is given by

$$P = \frac{V_p C_{di} \mathscr{F} f_o F_d}{e} \qquad \text{Eq. 5.2}$$

Variables in these equations are explained below. The factor $\mathscr{F}$ is constant.

$\mathscr{F}$ = 96,500 A·s/eq

Certain variables depend on the application; i.e., what is the salinity of the feed, how much this is to be reduced, and how much water is to be treated.

C_{di} = inlet feed concentration, eq/m^3

$$f_o = \text{"cut"} = \frac{\text{feed concentration} - \text{product concentration}}{\text{feed concentration}}, \text{ dimensionless}$$

F_d = feed rate, m^3/s

Other variables are determined by the equipment used:

e = current efficiency, dimensionless

R_{po} = resistance per cell pair, $\Omega\cdot m^2$

V_p = cell pair voltage, V

For estimating purposes, one can take e = 90%, R_{po} = 0.01 $\Omega\cdot m^2$, and V_P = 2 volts. The approximation for area in equation 1 becomes inaccurate for cuts greater than 50 percent. If a cut of 75 percent is required. for example, it is best to assume that there will be two stacks each operating at 50 percent and total the area required.

5.7 Electrodialysis Operation

Passage of current through the stack causes ion removal from the solution being desalted. The ion removal occurs right at the membrane solution interface. Almost 100 percent of the current in the membrane is carried by the counterions, while only about 50 percent of the current in the solution is carried by ions of the same charge. This leads to lowering of concentration at the interface. This deficiency of ions at the interface is made up by convective mass transfer and diffusion from the bulk solution. The rate of convective transfer determines the maximum current flow through the stack. Thus, considerable design effort has gone into the problem of improving convective mass transfer to the membrane surface, and fluid velocity through the compartment is an important design variable. Marinčić and Leitz (1972) reported detailed studies on the effect of turbulence promoters on mass transfer to a flat surface like that of the ED membrane.

As the solution passes between the membranes bounding the diluting compartments, salt is gradually stripped from the solution. This salt is transferred into the concentrating compartments. As one moves down the path from the feed end, the cell pair resistance gradually increases, and since the driving potential is constant, the current density slowly decreases. In practice, the system is usually designed so that the salt concentration in the diluate is decreased to no more than one half in a single stage, i.e., between a single pair of electrodes. A sufficient number of stages is used to get the desired total degree of desalination.

If the same spacers are used for the diluting and concentrating compartments, then the same flow velocity is normally used to keep the pressure drop across the membranes to a minimum. If the concentrate stream were discharged directly after a single pass, one of the best features of electrodialysis would be lost. Very high recoveries can be obtained by recirculating most of the concentrate stream, i.e., by operating in the feed-and-bleed mode.

A comparatively recent development has been electrodialysis reversal, often abbreviated to EDR. In this process, the polarity of the applied voltage is changed periodically, every 15 minutes or so. Reversing polarity converts the concentrating compartments to diluting compartments and vice versa. This means that the plumbing connections need to be reversed and for a few seconds after reversal, the product is out of specifications. The electrodes need to be able to function as both anodes and cathodes. What these modest inconveniences provide, however, is a self-cleaning stack. Any solid matter which may have been pushed against the membrane surface by the current is now carried in the opposite direction away from the membranes.

5.8 Electrodialysis Applications

The principal application of electrodialysis in the United States is for the desalination of water. Other applications include whey deashing (desalting) and recovery and concentration of metals from spent plating solutions. In Japan, the principal use has been for concentration of salts in sea water to produce table salt. An extensive list of references showing the breadth of applications is given in the chapter on electrodialysis in Weissberger and Rossiter (1971).

Wangnick (1994, 1996) provides data on the worldwide application of electrodialysis for water desalting. The total capacity of land-based ED desalting plants with a product capacity of 100 cubic meters/day (26,400 U.S. gallons/day) in 1996 is reported to be 1.16 million cubic meters per day (307 million U.S. gallons per day). The 1994 report lists a total of 828 plants with a product capacity of 100 cubic meters/day, of which 13 are reported as out of use or removed. The 1996 report lists a total of 389 plants with a capacity of 500 cubic meters per day, including hybrid process plants of which 7 are reported as out of use or removed. Of the 389 plants, 327 are listed as reversal plants. The great majority of these plants have been supplied by Ionics, Incorporated. Recently, a number of plants have proceeded from Asahi Glass Co., Ltd. Other suppliers over the last three decades have included Asahi Chemical Co., Ltd.; Christ AG; Deutsche Babcock Anlagen AG; Ebara-Infilco Co., Ltd./Tokuyama Soda Co., Ltd.; FISIA SpA; Japan Organo Co., Ltd.; Mitsubishi Heavy Industries, Ltd.; Nippon Rensui Co., Ltd.; Permutit Company, Inc.; Portals Water Treatment, Ltd.; Shinko-Pantec Co., Ltd.; and Société Recherche Technique et Industrielie. Whether some of these companies are still producing or selling electrodialysis equipment is not known.

5.9 Electrodialysis Economics

Electrodialysis costs are difficult to generalize. They are specific to feedwater and sensitive to the economic climate in which the plant is built. Probably the best generalization is that below a feed salinity of 3,000 mg/L ED can be competitive with and may be economically or operationally advantageous over RO.

A very detailed study on operation and costs of electrodialysis plants in the United States is that done by DSS Engineers (Latour, 1980), Volume III. This group visited each ED plant and interviewed the operators on site. While slightly dated, this study goes into considerable detail. For comparison of processes, this group reported on 13 RO plants in Volume II of the same study.

More recently, results of a survey based on responses to a one-page questionnaire received from 17 owners of electrodialysis reversal plants are presented by Leitner (1997). This study also covers RO and membrane softening plants.

An interesting comparison of plant performance and cost for ED, RO, and NF under a single set of operating conditions will be provided by the Port Hueneme demonstration plant. This plant, described by Thompson et *al.* (1997), contains a train of each of these processes designed to reduce water with total dissolved solids of 1,015 ppm to 370 ppm. In the case of RO and NF, the product salinity will be obtained by blending with bypass water. Operation is expected to begin in 1998.

Chapter 6
PRETREATMENT PROCESSES

The importance of pretreatment prior to membrane separations cannot be stressed enough. Membrane performance and life expectancy depend heavily on the quality of the feed water. Ideally, feed water to RO and NF systems should be clean, containing nothing but dissolved salts. In reality, many of these systems are run with a lower quality feed water, which causes problems with fouling, decreased capacity, and increased cleaning and replacement costs. An extensive comparison of pretreatment technologies is beyond the scope of this work, but pretreatment is so important that a summary of common problems and technologies for remediation is offered.

6.1 Particulates

Cartridge microfilters are designed to remove at least 90 percent of particles larger than their nominal rated size. Most are depth filters, which trap particles within tortuous passages as the water runs through. They are dead-end systems that process all the water passing through after an initial wetting period. When the back pressure reaches a specified level (65 to 240 kPa), the filters are replaced. Particulates are disposed of with the filter as solid waste. Cartridge life expectancy depends on the particulate loading and the flow rate. Costs for cartridge filtration depend on the micrometer rating, size, replacement rate, and type of cartridge housing.

6.2 Colloidal Fouling

Colloids are metal oxides, soaps, detergents, proteins, organic matter, silicates, and clay usually found in surface water. These substances generally have a negative surface charge. Water molecules and positively charged ions form a double charge layer around colloidal particles. Positive molecular poles and cations are attracted to the surface of the particle. The negative poles of water molecules around the particle cause a repulsive force between particles. Conditions of high pressure and ionic concentrations at the membrane surface cause these repulsive forces to be overcome and attractive van der Waals forces then cause the particles to coalesce. The result is the formation of a colloidal slime on the membrane that is difficult to remove. For this reason, the SDI (silt density index), a measurement of colloid concentration, should be no higher than 5.0 for spiral wound membranes and no greater than 3.0 for HFF membranes. See van Olphen (1977), Iler (1979), or Hiemenz (1986) for more extensive discussions of colloid and surface chemistry.

6.2.1 Avoidance

Since colloids occur naturally in surface water, one way to avoid the problem of colloidal fouling is to locate intake pipes in wells adjacent to the surface water source. In this way, the soil and rock can be used as a natural filter. This method is used at some seawater RO plants. By using the beach sand as a filter, both SDI and turbidity of the intake water are reduced.

6.2.2 Flocculation, Settling, and Filtration

Colloids can be removed by co-precipitation with Al(III), Fe(III), or Si(IV) hydroxides. The negatively charged colloids are surrounded by the metal cations and thus form a nucleus for their precipitation. The same thing happens when lime is used for softening. When polymer is added for coagulation the long chains can act as bridges linking colloidal particles and aiding in floc formation (O'Melia, 1972).

Effectiveness of colloid removal is dependent on pH, the concentration of colloids, and the charge of the metal cation used in co-precipitation. Higher charged cations, such as Fe(III) and Si(IV), are more effective than sodium and Magnesium (with +1 and +2 charges). The optimum pH depends on the precipitant. The optimum pH for $Al(OH)_3$ precipitation is about 6 while that for $Fe(OH)_3$ is 8.

6.2.3 Diatomaceous Earth Filtration

DE (diatomaceous earth) is the fossilized remains of siliceous shells of marine origin. It is composed of grains with diameters from 5 to 100 μm. Colloids, emulsified oils, and other hydrocarbons are adsorbed onto the rough, porous surface of these grains (Degrémont, 1991). Diatomaceous earth may be treated with a substance such as alum to enhance surface retention of small particles, just as microbial slimes enhance retention in sand filters (Bell and Himes, 1982). However, DE filters are prone to plugging if the feed water bas an SDI of 6 or higher (Permasep Engineering Manual #506, 1982).

6.2.4 Ultrafiltration

Ultrafilters with MWCOs between 200 and 20,000 Daltons can be used to remove silica and colloids. These filters are operated at low pressure and high recirculation rates so that fouling is not as much of a problem as it would be with RO membranes. Some are durable enough to be cleaned under harsh conditions that RO membranes cannot tolerate.

The absence of sludge and the compactness of the process are some of the benefits of using UF as opposed to the traditional precipitation processes. UF requires the same type of equipment as RO. In fact, depending on the feed water composition and the type of membrane used, ultrafilters could be incorporated into the RO system ahead of the RO membrane. Booster pumps may be needed to get the pressure up to the level required for RO.

6.3 Iron and Manganese

Iron is found in groundwater and deep surface water low in dissolved oxygen. It may be in either the soluble ferrous state (Fe”); the ferric state (Fe^{+3}); in inorganic complexes with silicates, phosphates, polyphosphates, sulfates, cyanides, and others; or organic complexes with humic, fulvic, and/or tannic acids (Degremont, 1991). Underground water usually has iron as ferrous bicarbonate. When dissolved ferrous bicarbonate is exposed to air, the iron may become oxidized to the ferric state in which it forms an insoluble hydroxide that is difficult to remove from membrane surfaces. Manganese is usually found with iron and behaves similarly.

6.3.1 Oxidation/Filtration

Iron can be precipitated by increasing the oxidation potential and/or the pH. When iron concentration is between 5 to 10 mg/L and no other problems exist with manganese, color, turbidity, or organic acids, the iron can be precipitated by oxidation with air. The rate of reaction is directly proportional to the pH and dissolved oxygen content. After oxidation, the treated water requires filtration with media capable of removing 0.5- to 1 -mm particles.

If iron concentration is higher or other contaminants are present in the water, a coagulant such as aluminum sulfate or ferric chloride is added to enhance removal of organics. A settling tank between the oxidation and filtration steps is used to remove larger precipitates.

6.3.2 Iron Removal With Lime Softening

When softening is required, iron can be removed in the same step. At a pH of about 8, practically all iron is precipitated as $FeCO_3$.

6.3.3 Biological Removal

Under the right conditions, certain bacteria will convert Fe^{3+} to $Fe(OH)_3$ within or on the sheaths surrounding their bodies. The optimum conditions are:

Dissolved oxygen	0.2 to 0.5 mg/L
PH	6.3
Redox potential	100 mV

Any H_2S should be removed by aeration prior to the fermentation tank (Degremont, 1991).

6.3.4 Oxidation Of Manganese With MnO_2

The reaction rate of oxidation of Mn^{2+} with air is too slow to be practical; manganese dioxide (MnO_2) works much faster. When mixed into a sand filter, MnO_2 reacts with Mn^{2+} by the following reaction:

$$Mn^{2+} + MnO_2 \rightarrow MnO_2 \cdot Mn^{2+} \text{ Fast Rate} \qquad \text{Eq. 6.1}$$
$$MnO_2 \cdot Mn^{2+} + 0, \rightarrow 2MnO_2 \text{ Slow Rate}$$

In this way, manganese is held in the sand filter, giving oxygen in the water a chance to oxidize it. Eventually, the filter bed becomes saturated with manganese dioxide, leading to the formation of Mn_2O_3, and the bed must then be regenerated with potassium permanganate ($KMnO_4$). The regeneration follows the reaction:

$$Mn_2O_3 + 2MnO_4^- + 2H_2O \rightarrow 5MnO_2 + 4H^+ \qquad \text{Eq. 6.2}$$

Theoretically the required ratio of $KMnO_4$ to Mn is 1.9: 1 .O.

Potassium permanganate or ozone can be added directly to the water to oxidize Mn^{2+} in the same manner. The ratio of $KMnO_4$ to Mn^{2+} is 1.9: 1; the ratio for ozone is 0.9: 1 .O. The addition of excess oxidant will produce MnO_4^-, which turns the water pink (Degrémont, 1991).

63.5 Biological Manganese Removal

Manganese can be removed by biological processes similar to iron removal. However, they cannot be removed in the same bioreactor. Bacteria that accumulate manganese do so in an environment with a redox potential of at least 400 mV. At lower potentials, MnO_2 will be dissolved into Mn^{2+} instead of the reverse. Culturing a sufficient bed of manganese-accumulating microbes takes about 3 months (Degrémont, 199 1).

6.4 Scaling

Scaling is caused by the precipitation of slightly soluble salts such as $MgCO_3$, $CaCO_3$, $CaSO_4$, $BaSO_4$, $SrSO_4$, Cal?,, and SiO_2 within the membrane module and the concentrate removal system. Solubility of these salts depends on temperature and pH. Higher temperatures and pH facilitate precipitation of carbonates and sulfates. The concentration of these salts in the feed must be low enough that the concentration in the reject is still below saturation.

The LSI (Langelier Saturation Index) is the standard measure of scaling potential given by the simplified equation:

$$LSI = pH_R - pH_S \qquad \text{Eq. 6.3}$$

The pH_R is the pH of the reject; the pH_S is the saturation pH of the reject stream calculated from total dissolved solids, temperature, calcium concentration, and alkalinity of the feed water and the percent recovery. If the LSI is negative, $CaCO_3$ tends to dissolve; if it is positive, $CaCO_3$ tends to precipitate. The LSI is not a quantitative measure. It does not predict scaling, only the tendency toward scaling or corrosion. Some manufacturers void their guarantee if membranes are used with a positive LSI; others recommend limits. Because of reaction time, and the fact that the actual amount of $CaCO_3$ over the saturation level may be small, running an RO or NF system with a slightly positive LSI may be possible.

Softening techniques to lower the LSI involve shifting the carbonate system by:

- Acidification - converts HCO, to CO,;
- Addition of lime and soda ash - causes precipitation;
- Ion Exchange - replaces cations with sodium; and
- NF - removes dissolved divalent salts.

6.4.1 Acidification

The addition of HCl or H_2SO_4 causes the reduction of bicarbonate ion to water and carbon dioxide. At atmospheric pressure, the CO, bubbles out of solution and is no longer available for carbonate production. Bicarbonate is a potent buffer, the amount of H^+ required to change the pH is logarithmically proportional to the concentration of HCO_3^-.

After acidification has forced the bicarbonate ions out of solution, the metallic cations are balanced by Cl^- if HCI was used, or SO_4^-, if $H_2SO_4^-$ was used. Sulfuric acid is stronger than hydrochloric acid, so the pH can be lowered with a smaller amount, but the fact that SO_4^- can cause scaling problems with barium, lead, and calcium should be taken into consideration.

6.4.2 Lime Softening

Lime, $Ca(OH)_2$, reacts with bicarbonate ions in the following manner:

$$\begin{aligned}&Ca(OH)_2 + Ca(HCO_3)_2 \rightarrow 2CaCO_3\downarrow + 2H_2O \text{ and}\\&Ca(OH)_2 + Mg(HCO_3)_2 \rightarrow CaCO_3\ 1 + MgCO_3 + 2H_2O\end{aligned}$$

Eq. 6.4

Additional lime will precipitate Mg^{2+} to its theoretical solubility under existing conditions:

$$Ca(OH)_2 + MgCO_3 \rightarrow CaCO_3\downarrow + Mg(OH)_2\downarrow$$

Eq. 6.5

Lime softening is an old, reliable process, but it produces massive quantities of sludge that require settling ponds and filter presses for dewatering. To remove 160 mg/L of $Ca(HCO_3)_2$ and 160 mg/L of $Mg(HCO_3)_2$ from one cubic meter of water by the above equations, 15.5 g of lime are required that produce 3 10 g of carbonate solids. The solids are mixed with water though, after settling and pumping off the supematant, the sludge is still about 75 percent water by weight, resulting in a total weight of 1,240 g. Multiply these amounts by a modest 5,000 m^3/d plant, and the sludge is increased to 6,200 kg/d.

6.4.3 Ion Exchange

Another approach for softening is to use cationic resins to replace the calcium ions (or other divalent cations) with sodium ions, which do not form insoluble salts with carbonate ion. Resin beads have multiple sites of ionic attachment. These sites are preferentially taken up by multivalent cations. During the service phase of operation, two sodium ions are displaced by a calcium cation. The number of sites per volume of resin (capacity) is supplied by the manufacturer. The depth of the resin bed and the necessary flow rates can be calculated from the resin capacity, the divalent ion concentration in the feed water, and the volume of water to be treated.

When most of the attachment sites have been taken up, calcium begins to appear in the product water and the resin must be regenerated. During regeneration, a strong sodium chloride solution is passed through the resin bed until the amount of calcium in the product water falls off. Rohm and Haas (1978) recommend 3 to 10 L of 10 percent NaCl solution per liter of resin at a flow rate of 130 mL/min per liter of resin for complete regeneration of their Amberlite® IR-120 Plus (Rohm and Haas, 1978).

Ion exchange is well suited for incorporation into an RO system. Depending on the salinity and pH, the RO concentrate may be used as regenerant solution. Cation exchange resins can be adequately regenerated at lower concentrations than 10 pct with lower flow rates and longer regeneration cycles (Haugseth and Bietelshees, 1974). However, precipitation of $CaCO_3$ may cause problems because the cations removed from the resin are reintroduced to the concentrated bicarbonate ions in the concentrate.

6.4.4 Nanofdtration

NF membranes preferentially reject divalent ions over monovalent ions at a rate of about 95 percent to 75 percent. Chloride ions tend to pass through the membrane while divalent carbonate ions are retained. It is unclear whether bicarbonate ions are rejected at the same rate as chloride. The rate of rejection really depends on the composition of the water. The calcium and magnesium must be balanced by an equivalent negative charge. If divalent anions are scarce then ion size will be the determining factor. Larger monovalent ions will tend to be retained and bicarbonate is almost twice as large as chloride (61:35).

The reject from NF contains approximately 95 percent of the divalent (and higher) cations, 50 to 75 percent of the monovalent cations, and an equivalent percentage of the higher charged, and larger anions in 10 to 15 percent of the water. Because some bicarbonate passes through the membrane, scaling may not be a problem in the reject conveyance system. Only a pilot test can tell exactly what the reject carbonate concentration will be. If supersaturation occurs, the carbonate balance of the reject must be adjusted. $CaCO_3$ crystals can be seeded at this point, encouraging the scale formation on the crystals instead of the plumbing, or acid can be added to shift the carbonate balance toward CO, formation rather than $CaCO_3$.

The benefits of NF are that no sludge is produced, little if any additional chemicals are needed, and it is compact in size. The disadvantages are the difficulty in predicting anion rejection, the need for cleaning and proper storage of the membranes, pretreatment for bacteria, and the disposal of the concentrate stream.

6.5 Biological Matter

Biological matter refers to micro-organisms, living or dead, and pyrogens, which are biological waste products excreted on the outer surface of micro-organisms or bits of micro-organisms. When live microbes are present in the feed water they form colonies on membranes and other surfaces on or in the pressure vessel. Microbial colonies plug the pores of the membrane, decreasing productivity. Ridgeway et al. (1984) found that these colonies did not cause a decrease in salt rejection and the productivity leveled out at a sustainable level. However, when the feed water was heavily dosed with chlorine to kill the micro-organisms and then dosed with an excess of ammonia to convert the free chlorine to chloramine, biological fouling was enhanced. Productivity continued to decline until the end of testing. The accumulation of dead microbes on the membrane surface is similar to colloidal fouling. Colonies are not formed, but the gelatinous slime of decomposing bacteria inhibits water passage. To alleviate this problem, a 0. 1 - to 1 .0-μm filter should be used with disinfection processes that do not involve filtration.

6.5.1 Oxidation

Several oxidants can be used for disinfection of drinking water. Table 6.1 compares relative efficacy, remanent effect, speed, and required dosage for each. Chlorine and chlorine dioxide can be used for both initial disinfection and residual protection from reinfection, but they cannot be present in water used with most thin film composite membranes. Ozone and UV (ultraviolet radiation) are the best candidates for use with TFC membranes. They are effective for initial disinfection, but ozone dissipates rapidly, and *W* is effective only while the water is exposed to it. As mentioned above, a micro filter is needed with any oxidant to remove the dead and living microbes ahead of the membranes.

6.5.2 Media Filtration

Media filtration is a well known, effective treatment process for lowering SDI by removing particulates, colloids, and bacteria. Sand media acts as both a filter and settling chamber, trapping suspended solids within the bed. Filtering action is enhanced by biological growth within the bed. Microbes break down large organics and nitrates and excrete a gelatinous substance which enhances filtration of sub-micron particles (Bellamy, et al., 1985).

A media filtration system may have one, two, or three layers of progressively finer graded material. Mono-media sand filters have only a bed of sand. Dual-media filter systems have a layer of anthracite on top of the sand. A layer of garnet below the sand also may be added to form a multi-media filter. The layers have progressively smaller grain size from top to bottom (1.1 mm anthracite to 0.2 mm garnet), and higher specific gravity (1.5 for anthracite to 4.2 for garnet). The layering of coarse to fine media improves the flow rate of water through the bed and promotes filtration in depth. Single media filters are run as high as 122-163 L $min^{-1}m^{-2}$; dual and multi-media filters can be operated as high as 244-6 10 L $min^{-1}m^{-2}$ (Parekh, 1991).

Table 6.1 .-Qualities of oxidants for drinking water disinfection

	Ozone	Cl_2	ClO_2	Chloramines	u v
Efficacy	+++	++	++	+	++
Remanent effect	0	+	+	++	0
Speed	+++	++	++	+	+++
Dosage	0.4 mg/L 4 min.	0.5 mg/L 30 min. at pH 8	0.2 mg/L 15 min.	Used for remanent effect after 0, or UV	20-25 $mW{\cdot}s/cm^2$
Conditions	No Mn^{2+}: oxidizes to MnO_2	Remove THMPs	Forms ClO_2^- in oxidation of NOM	Used for post treatment with 0, and UV	Turbidity <1 NTU, no iron, NOM, thin stream of water, clean equipment

NOM is natural organic matter, $mW{\cdot}s/cm^2$ is milliwatts second per square centimeter, and THMPs are trihalomethane precursors. Adapted from Degrémont, 1991.

Dual and multi-media beds are back-washed to remove trapped solids at higher flow rates than used in operation. Because of the gradation in specific gravity of the layers, the media settle back in the same sequence (Parekh, 199 1). Single media beds, such as slow sand filters, are not generally back-washed. Instead, when the flow rate declines, the top layer of filtration cake, or schmutzdecke, is scraped off. When cleaning reduces the sand depth to 0.3 to 0.5 m, the bottom sand is removed, and new sand is added to the bottom. The top layer is replaced with the old sand to maintain the flora and fauna of the bed (Bellamy, et al., 1985).

6.5.3 Ultrafiltration

As discussed earlier, UF is useful for reducing bacteria, virus, pyrogens, organics including color and odor causing compounds, and trihalomethanes and their precursors. In sewage treatment, ultrafilters can be used to filter effluent from bioreactors. Effluent is recirculated at a rate sufficient to keep the membrane surface clean. In this way, organics are retained in the reactor until they have been broken down to a MW below the MWCO of the ultrafilter. Bacteria cannot pass the membrane and so are also retained in the bioreactor. Because no added coagulants or granular material are present, the sludge volume remains low, requiring disposal only once every month or two (Stavenger, 1971).

6.6 Deoxidation

The presence of free oxidants is destructive to thin film composite membranes; eventually it breaks down even cellulosic membranes that can tolerate up to 5 mg/L of free chlorine. Unfortunately, oxidation is usually required to keep microbial life in check. If halogens are used, dechlorination is required just before feed water enters the membrane modules. Dechlorination can be achieved by adding some form of SO, or filtering with granular activated carbon (GAC). Reduction with SO, is a fast reaction, which can work in the feed pipes on the way to the RO system. SO, can be sulfur dioxide gas, crystallized sodium sulfite, or sodium bisulfite. The reactions are as follows:

$$\text{SO,} + H_2O \rightarrow H_2SO_3 \qquad \text{Eq. 6.5}$$

With free chlorine: $H_2SO_3 + HClO \rightarrow HCI + H_2SO_4$ Eq. 6.6

With monochloramine: $H_2SO_3 + NH_2Cl + H_2O \rightarrow NH_4Cl + H_2SO_4$ Eq. 6.7

GAC takes longer to deoxidize water than SO,. A reaction chamber is required to keep the water in contact with the carbon for a few minutes. Unfortunately, an activated carbon bed is perfect for culturing bacteria. Organic molecules adsorbed by the carbon provide nourishment. With the water free of oxidants, nothing keeps microbes from flourishing downstream in the RO module.

The presence of fine carbon dust mixed with the grains is another reason not to use GAC for deoxidation, If the GAC is just ahead of the RO system, some carbon fines will deposit in the membrane module where they erode the membranes and clog membrane spacers. If GAC must be used for dechlorination, a sub-micrometer filter should be placed after the reactor, and the quality of the GAC effluent should be monitored closely for bacterial contamination.

6.7 Organic Compounds

Organic compounds are a diverse group of water contaminants. They come from a variety of sources: humic and fulvic acids from the breakdown of vegetative matter and oils, phenols, pesticides, surfactants, tannins, and chlorinated methanes from precipitation runoff from cities and farms, not to mention industrial wastes. Table 6.2 compares processes for removing organic compounds based on polarity, size, and functional groups. Some organics are removed at each stage of pretreatment: aeration removes volatile organic compounds; polar organic compounds are adsorbed during flocculation; and filtration removes organics adsorbed into flocs too small to have settled. Flocculation and filtration are discussed above. Air stripping and granular activated carbon filtration are discussed in this section.

6.7.1 Granular Activated Carbon

Activated carbon has a high adsorbent capacity per volume and can be used on a supporting layer of cellulose, Diatomaceous earth, or in a column by itself to remove organics as listed above. GAC also will remove free chlorine remaining in water after disinfection as discussed above. So, if a remanent effect is desired in the water, chlorine should be added after the GAC.

Table 6.2.—Processes for removing organic compounds

Process	Type of organics removed	Mode of action
Air stripping	Volatile compounds	Evaporation
Coagulation ▪ flocculation ▪ settling	Humic and fulvic acids	Adsorption
Ozonation	Phenols, detergents, polycyclic hydrocarbons, certain pesticides	Oxidation
Activated carbon	High MW, low polarization, sapid compounds (flavor), phenols, surfactants, saturated hydrocarbons, pesticides	Adsorption
Ultrafiltration	Globular shaped molecules over MWCO, some linear molecules significantly over MWCO.	Separation and concentration

Adapted From Wafer Treatment Handbook, Degrémont, 1991.

6.7.2 Air Stripping

Volatile organic compounds can be removed from water by aeration. Air can be incorporated into water in several ways: cascading the stream down a series of steps; bubbling air into the water; or spraying the water up into the air in a fountain. The idea is to bring an excess of oxygen in contact with the water so that volatile organic compounds can escape.

Chapter 7
WASTE DISPOSAL

Typical waste products from membrane systems include the pretreatment sludge, cleaning and storage chemical solutions, the concentrate, and used membranes. Ways to minimize or avoid sludge production are discussed in the chapter on pretreatment. Cleaning and storage chemicals each have their own characteristic disposal requirements and are usually considered industrial wastes. Currently, used membranes are considered solid waste, but possibilities for reconditioning and recycling exist.

Usually the most significant waste disposal issue for membrane systems is the concentrate. The major difficulty is that the concentrate has significantly higher TDS than both a typical wastewater discharge and the receiving water. Source water quality, pretreatment, membrane type, process-added chemicals, and process configuration (recovery) influence the quality and quantity of the concentrate. The method generally accepted for determining the expected strength of the concentrate assumes 100 percent rejection of salts. The concentration factor (CF) is calculated by the following equation:

$$CF=\frac{1}{1-Y} \qquad \text{Eq. 7.1}$$

Where Y equals the recovery portion in decimal.

From this equation, the concentrate from a 75 percent recovery system processing feed water with 3,000 mg/L TDS would have 12,000 mg/L TDS. At 90 percent recovery, the concentrate would have 30,000 mg/L TDS. A membrane system with 10 Mgal/d product water capacity will also produce 3.3 Mgal/d of concentrate at 75 percent, or 1.1 Mgal/d of concentrate at 90 percent. This trade-off between quantity and quality of concentrate can strongly influence, and be influenced by, the disposal options. Sometimes it is advisable to blend the concentrate with some other water or wastewater in order to make disposal more practical or economical. Some examples are blending with treated sewage, an industrial waste stream, or power plant cooling water.

How the concentrate or blend may be disposed of depends on its quality and quantity, geographic availability, and potential impacts on the receiving water, soil, or use. The most common disposal options for membrane system concentrates are:

- Surface water body,
- Wastewater treatment plant,
- Land application,
- Injection well,
- Evaporation pond, and
- Solar pond.

The quantity and quality of the concentrate or blend and the disposal method determine which regulatory requirements apply. Concentrate disposal is regulated under several Federal laws, and state and local governments may impose additional regulations. Any or several of the following Federal laws may be applicable to a specific case of concentrate disposal:

- Clean Water Act (CWA), includes
  - National Pollutant Discharge Elimination System (NPDES) program,
  - National Pretreatment Standards program, and
  - Water Treatment Sludge Disposal regulations (Sec. 503).
- Coastal Zone Management Act (CZMA).
- Comprehensive Environmental Response, Compensation, and Liability Act (CERCLA).
- Hazardous Materials Transportation Act (HMTA).
- Occupational Safety and Health Act (OSHA).
- Resource Conservation and Recovery Act (RCRA), includes
  - Solid Waste Disposal Act (subtitle D), and
  - Hazardous waste regulations (subtitle C).
- Safe Drinking Water Act (SDWA), includes
  - Underground Injection Control (UIC) program, and
  - Wellhead protection program (state-administered).
- Toxic Substances Control Act (TSCA).

States may assume regulatory/permit authority for the NPDES, Pretreatment Standards, and UIC programs if their requirements are at least as stringent as EPA's. Table 7.1 lists state authorities for these programs. In some states, membrane system concentrate may be classified as industrial waste and must therefore meet different disposal requirements than municipal water treatment waste.

A key issue in determining the applicability of some USEPA regulations is whether the concentrate meets the definition of "hazardous" or "toxic." Solid waste is defined broadly and includes waste in liquid, semi-liquid, and solid forms. Membrane system concentrates are generally considered a solid waste. A solid waste is considered hazardous if (1) it is specifically identified on any of the lists published in the RCRA regulations, or (2) it exhibits any one of the four specific characteristics of hazardous wastes: ignitability, reactivity, corrosivity, and toxicity. A "toxic" concentrate would be one which contains substances regulated as toxins under the TSCA as "presenting an unreasonable risk of injury to health or the environment" (15 U.S.C. §2601(a)(3)). In general, membrane system concentrates are not likely to be classified as a hazardous or toxic waste but this possibility should be evaluated, particularly if the raw water quality is unusual or the concentrate disposal options are limited.

The relative cost of permit and regulatory requirements should be considered when selecting a disposal option. In addition to Federal programs, state and local regulations may vary greatly and some agency requirements may conflict. Because regulatory requirements and agency personnel can change with time, care should be taken to ensure that decisions made by agencies are explicitly documented throughout a project.

Table 7.1 .-Regulatory program authority by state as of October 1997
(Does not include Indian lands and some Federal facilities)

USEAP region and state	NPDES program	Pretreatment program	UIC program
Region I			
Connecticut	yes	yes	yes
Maine	no	no	yes
Massachusetts	no	no	yes
New Hampshire	no	no	yes
Rhode Island	yes	yes	yes
Vermont	yes	yes	yes
Region II			
New Jersey	yes	yes	yes
New York	yes	no	no
Virgin Islands	yes	np	no
Puerto Rico	no	no	yes
Region III			
Delaware	yes	no	yes
District of Columbia	no	no	no
Maryland	yes	yes	yes
Pennsylvania	yes	no	no
Virginia	yes	yes	no
West Virginia	yes	yes	yes
Region IV			
Alabama	yes	yes	yes
Florida	yes	yes	(1)
Georgia	yes	yes	yes
Kentucky	yes	yes	no
Mississippi	yes	yes	yes
North Carolina	yes	yes	yes
South Carolina	yes	yes	yes
Tennessee	yes	yes	no
Region V			
Illinois	yes	no	yes
Indiana	yes	no	(1)
Michigan	yes	yes	no
Minnesota	yes	yes	no
Ohio	yes	yes	yes
Wisconsin	yes	yes	yes
Region VI	yes		
Arkansas	yes	yes	yes
Louisiana	yes	yes	yes
New Mexico	yes	no	yes
Oklahoma	yes	yes	yes
Texas	yes	no	yes
Region VII			
Iowa	yes	yes	no
Kansas	yes	no	yes
Missouri	yes	yes	yes
Nebraska	yes	yes	yes
Region VIII			
Colorado	yes	no	no
Montana	yes	no	yes
Nevada	yes	no	yes
North Dakota	yes	no	yes
South Dakota	yes	yes	(1)
Utah	yes	yes	yes
Wyoming	yes	no	yes
Region IX			
Arizona	no	no	no
Hawaii	yes	yes	no
California	yes	yes	(1)
Region X			
Alaska	no	no	(1)
Idaho	no	no	yes
Oregon	yes	yes	yes
Washington	yes	yes	yes

[1] Primacy for UIC is shared with EPA.

7.1 Surface Water Body

The primary concern about discharging membrane system concentrate to surface waters is degradation of the receiving water. Receiving water quality influences whether treatment of the concentrate is required before discharge. In general, concentrates may have low dissolved oxygen, high hydrogen sulfide, and low pH. The concentrate may also have "common ion toxicity" which occurs when the concentrate has an excess or deficiency of one or more common ions relative to the ion composition of the receiving water.

An alternative to discharging membrane system concentrate directly to surface water is to combine it with treated sewage or storm water runoff. If seawater membrane concentrate is mixed with 1,000 mg/L TDS water (very high for treated sewage) at a ratio of 2: 1 (concentrate:sewage), the TDS can be reduced to that of the ambient seawater. Similar dilution of brackish membrane system concentrate would make it compatible with an inland surface water. Concentrates may also be combined with power plant cooling water which provides thermal dilution for the power plant discharge and TDS dilution for the concentrate.

All discharges into the navigable waters of the U.S. (all streams, lakes, rivers, and oceans) are required to be permitted through the NPDES program. The five-year permits generally contain numerical effluent limitations for specific pollutants

If good circulation is present at the concentrate outfall, the high TDS should dissipate rapidly, but discharging large volumes of high TDS water into small lakes, enclosed bays, shellfish beds, or valuable fishing waters might pose some hazards. The influence of local currents may also assist in dispersing the plume concentrations. Impact studies of the effect of discharge on aquatic life may be required before an NPDES permit is issued. Wetlands are regulated under the CWA. Wetlands protection laws are an important consideration in the siting and construction of membrane systems and concentrate disposal facilities.

The concentrate must meet the enforceable MCLs (maximum contaminant levels) promulgated under the SDWA if it is disposed of into a protected drinking water supply source as determined by the state (SDWA, 42 U.S.C.A. § 300g). Table 7.2 lists the SDWA regulated contaminants. The MCLGs (Maximum Contaminant Level Goals) are recommended, but not enforceable, limits. Local regulations must also be complied with. For example, in Florida, discharges into surface water also must meet requirements for Class III waters that are to be kept suitable for recreation and the propagation and maintenance of fish and wildlife populations. These requirements involve regulations limiting hydrogen sulfide, dissolved oxygen levels, sulfide toxicity, low pH, corrosiveness, and toxins (Malaxos and Morin, 1990).

Most coastal membrane systems discharge concentrate to the sea. If the raw water was obtained from some inland source, the TDS of the concentrate will be less than or equal to the TDS of seawater. However, if the raw water is seawater, the TDS of the concentrate will be significantly higher than the seawater TDS. Salinity affects density, osmotic pressure, and buffering properties of seawater and supply of oxygen. Measurements of seawater

Table 7.2.—SDWA regulated contaminants (1997)

Contaminants	MCLG (mg/L)	MCL (mg/L)	Contaminants	MCLG (mg/L)	MCL (mg/L)
Fluoride	4.0	4.0	Ethylbenzene	0.7	0.7
			Styrene	0.1	0.1
Trichloroethylene	zero	0.005	Heptachlor	zero	0.0004
Carbon Tetrachloride	zero	0.005	Heptachlor epoxide	zero	0.0002
1,1,1-Trichloroethane	0.2	0.2	Nitrite (as N)	1	1
1,2-Dichloroethane	zero	0.005	Aldicarb sulfoxide	0.001	0.004
Vinyl Chlorite	zero	0.002	Aldicarb sulfone	0.001	0.002
Benzene	zero	0.005			
p-Dichlorobenzene	0.075	0.075	Lead	zero	I-T+
1,1-Dichloroethylene	0.007	0.007	Copper	0.07	I-r++
Total Coliform	zero	<5%	Dichloromethane	zero	0.005
Turbidity	N/A	I-r	1,2,4-Trichlorobenzene	0.07	0.07
Giardia lamblia	zero	I-r	Hexachlorobenzene	zero	0.001
Viruses	zero	I-r	Antimony	zero	0.006
Standard plate count	zero	I-r	Nickel	0.1	0.1
Legionella	zero	I-r	Thallium	0.002	0.002
			Beryllium	0.004	0.004
Tetrachloroethylene	zero	0.005	Cyanide	0.2	0.2
Chlorobenzene	0.1	0.1	Endrin	0.002	0.002
trans-1,2-Dichloroethylene	0.007	0.007	Dalapon	0.2	0.2
cis-1,2-Dichloroethylene	0.07	0.07	Diquat	0.1	0.1
o-Dichlorobenzene	0.6	0.6	Endothall	0.1	0.1
Barium	2	2	Gfyphosphate	0.7	0.7
Cadmium	0.005	0.005	Di(2-ethylhexyl)adipate	0.4	0.4
Chromium (total)	0.1	0.1	2,3,7,8-TCDD (Dioxin)	zero	3x10"
Mercury (inorganic)	0.002	0.002	1,1,2-Trichloroethane	0.005	0.005
Nitrate (as N)	10	10	Oxamyl (Vydate)	0.2	0.2
Selenium	0.05	0.05	Simazine	0.004	0.004
Asbestos (fiber >10μm)	7 MFL	7 MFL	PAHs Benzo(a)pyrene	zero	0.0002
Lindane	0.0002	0.0002	Diethylhexyl Phthalate	zero	0.006
Methoxychlor	0.04	0.04	Picloram	0.5	0.5
Toxaphene	zero	0.003	Dinoseb	0.007	0.007
2,4-D	0.07	0.07	Hexachlorocyctopentadiene	0.05	0.05
2,4,5-TP (Silvex)	0.05	0.05			
Aldicarb	0.003	0.003	**Radionuclides (Proposed)**		
Chlordane	zero	0.002	Radium 226	zero	20pCi/L
Carbofuran	0.04	0.04	Radium 228	zero	20pCi/L
Alachlor	zero	0.002	Gross beta Particle and Photon		
Epichlorohydrfn	zero	I-r	Emitters	zero	4mRem/y
Totuene	1	1	Uranium	zero	r
PCBs	zero	0.0005	Gross alpha Emitters	zero	30pCi/L
Atrazine	0.003	0.003	Radon	zero	15pCii
Acrylamide	zero	I-r			300pCi/L
Dibromochforopropane (DBCP)	zero	0.0002	**Sulfate (Proposed)**	500	
1,2-Dichloropropane	zero	0.005			
Pentachlorophenol	zero	0.001	**Arsenic (Interim)**	0.002	0.05
Ethylene dibromide (EDB)		0.0000			
Xylenes (total)	10	510	**Disinfection By-Products (Interim)**		
			Total Trihalomethanes	0.01	0.1

+ Action Level = 0.015 mg/L
++ Action Level = 1.3 mg/L
TT Treatment technique requirement
Source: USEPA

characteristics show natural salinity variations of the order of +/- 1,000 mg/L due to tidal actions, fresh water runoff, and seasonal variations. Conservatively, it would appear that a benthic ecosystem could easily tolerate a salinity perturbation of 1,000 mg/L, which is about 3 percent deviation from the ambient.

Marine organisms can tolerate gradual increases in salinity, but sudden changes are fatal. A specific gravity of 1.030 g/cm’ is the upper limit of tolerance for most species outside of the Mediterranean. Assuming a linear relationship between specific gravity and TDS, an increase of 0.008 g/cm^3 (normal specific gravity = 1.022 for seawater) equates to an increase in TDS of only 36 percent (from a normal 35,000 to 47,727 mg/L). Mobile creatures can move away from the concentrate discharge but immobile species will die if the TDS suddenly increases too much. Denser seawater RO concentrate sinks rapidly to the ocean floor and may harm benthic organisms because of the high salinity.

NPDES permits for ocean outfalls for regular wastewater treatment plants (WWTPs) often require a certain depth of discharge so that the rising plume of the lighter-than-seawater effluent has adequate time to mix and dilute before reaching the surface. In contrast, the higher density of seawater membrane system concentrate results in a directly discharged plume which sinks to the ocean floor. So an NPDES permit should require a certain depth of discharge and/or mixing zone. Mixing and dispersion may also be accomplished by discharging the concentrate through a seabed distribution system of pipelines, or one pipeline with many outlets.

The CZMA is administered by the Office of Coastal Zone Management in the National Oceanic and Atmospheric Administration, which is part of the U.S. Department of Commerce. Water development and construction projects which are within designated coastal zones and that require one or more Federal permits are subject to CZMA requirements. This would include construction or modification of an ocean outfall. The coastal zone states are all states adjacent to the Great Lakes and the East, West, and Gulf coasts. Each state defines its own coastal zone; some have included river segments more than 100 miles (160 km) from the coastline.

7.2 Wastewater Treatment Plant (WWTP)

As indirect dischargers, membrane systems which discharge concentrate to WWTPs are exempt from NPDES regulation. Rather, they are subject to USEPA National Pretreatment Standards and to any additional regulations imposed by the state or the WWTP. Municipal sewer authorities may restrict concentrate disposal to sanitary sewers because of concern that the high TDS would adversely affect their biological treatment processes. Discharge quality limitations and conditions may be imposed by the WWTP through a permit, a local ordinance, or both. Many WWTPs set limits to enable them to achieve compliance with their own permit requirements. If the use of an existing permitted wastewater ocean outfall is proposed for additional concentrate disposal, the mixing characteristics of the combined discharge may require a NPDES permit modification. WWTPs also need to control the pollutants that enter their sludge to ensure that they will be able to continue managing and disposing of their sludge in accordance with USEPA regulations (CWA, section 503).

7.3 Land Application

The interrelated actions of soil filtering, plant uptake, plant metabolism, soil biochemical and physicochemical reactions, and soil particle adsorption must all be considered in evaluating the impacts of land application disposal of concentrate.

Permit requirements will usually depend upon the ultimate destination of the concentrate. State and local regulations usually govern, and an NPDES permit may be required if ultimate discharge is to surface water. In general, application of irrigation water must not adversely affect the groundwater being recharged, its users, or the surrounding environment. State and local wellhead protection programs may limit land application disposal that could adversely affect local drinking water wells. Spray irrigation results in concentrate entering the groundwater unless under-drains are provided. Careful planning is required to prevent groundwater contamination. An NPDES permit will be required if under-drains are designed to discharge to surface water.

If the concentrate TDS is low enough, or if it can be blended with another water (e.g., wastewater effluent) which will result in a sufficiently low TDS, the concentrate may be usable for irrigation of certain crops. However, even salt-tolerant crops require irrigation water with a chloride concentration of 1,000 mg/L or less. Table 7.3 lists threshold levels of irrigation water TDS for several crops. These are levels above which yields begin to decrease with increased TDS. The levels are based on the conductivity of saturated soil extract from the root zone; a factor of 640/1.5 was used to convert the conductivity of soil extract in deciSiemens/meter to irrigation water TDS in mg/L.

Table 7.3.—TDS threshold levels in applied irrigation water

Crop	Maximum TDS	Crop	Maximum TDS
Lettuce	555	Alfalfa	853
Cotton lint	3,285	Grapes, table	640
Carrots	427	Cantaloupe	1,422
Wheat	2,560	Dates	1,707
Oranges	725	Sugar beets	2,987
Grapefruit	768	Lemons	768
Onions	512	Beans	427
Corn	726	Cabbage	768
Celery	768	Peppers	640
Potatoes	725	Spinach	853
Strawberries	427	Sweet potato	640
Almonds	640	Berries/plums	640
Peaches	725	Avocados	427

Source: Lohman, Milliken, and Dorn, 1988, p. 23.

Because certain crops bioaccumulate certain minerals, irrigation water should not contain excessive amounts of trace elements or metals. Table 7.4 lists guidelines for maximum trace element levels in irrigation water. Of course, the degree of this risk is also influenced by the type of crop, the soil, and irrigation frequency and timing.

Table 7.4.—Recommended maximum trace element levels in irriaation water

Element	Suggested maximum irrigation water level (mg/L)
Aluminum (Al)	10.0
Arsenic (As)	0.1 - 2.0
Beryllium (Be)	0.1 - 0.05
Boron (B)	0.5 - 2.0
Cadmium (Cd)	0.01 - 0.05
Chromium (Cr^{+6})	0.5 - 1 .o
Cobalt (Co)	0.1 - 5.0
Copper (Cu)	0.2 - 5.0
Fluoride (Fl)	1.8
Iron (Fe)	5.0 - 20
Lead (Pb)	5.0 - 10
Lithium (Li)	2.5
Manganese (Mn)	0.02 - 10
Molybdenum (Mo)	0.01 - 0.05
Nickel (Ni)	0.2 - 2.0
Selenium (Se)	0.02
Silver (Ag)	4.0 - 8.0
Vanadium (V)	0.1 - 1.0
Zinc (Zn)	2.0 - 10.0

Source: Edwards and Bowdoin, 1990.

Soil permeability problems occur when the rate of water infiltration into and through the soil is reduced by the effects of salts in the water. The infiltration rate can be reduced to such an extent that the crop is not properly supplied with water. Generally, three factors determine a water's long-term influence on soil permeability: (1) sodium content relative to calcium and magnesium; (2) bicarbonate and carbonate content; and (3) total salt concentration of the water. Most salts in irrigation water are left behind in the soil as water is taken up by the crop. These salts may accumulate and reduce the availability of water in the root zone. This is called the osmotic effect and can be measured as a force the plant must overcome to obtain the water. Soils behave as cation exchangers. Various cations in the soil can be replaced by each other by proper choice of concentrations.

The potential impacts of the concentrate on the existing irrigation distribution system should also be evaluated. What are the impacts on aquatic organisms, on canal system materials, on field runoff areas, etc.?

Another option for land application is irrigation of constructed saline marshes. Some pilot studies of this concept have been conducted by Reclamation and others.

7.4 Injection Wells

Direct discharges to groundwater are subject to the UIC regulations set by USEPA under the SDWA. The UIC program deals with the construction and operation of underground injection wells. UIC regulations prohibit subsurface disposal through a well unless authorized by rule or by permit. Five classes of well are established, and disposal of certain types of wastes within each class are authorized by rule. The class of well that may be used and the associated requirements depend on the quality of the concentrate. Generally, injection of any fluid is prohibited if it will cause an underground source of drinking water to exceed any SDWA maximum contaminant level or otherwise affect public health. Depending upon the site-specific circumstances, RCRA and TSCA may also have some effect, but generally, state legislation or regulation covers the specifics.

Disposal by injection well is a simple, effective means of disposing of large volumes of concentrate continually under varying weather conditions. However, the development and construction of this option is very complicated. There are very particular geological requirements; the site must be over a confined aquifer which is unsuitable for drinking water, is below all drinking water aquifers in its vicinity, and has a relatively high transmissivity in order to accept the injected concentrate at economical pressures. The operation should be designed not to degrade the transmissivity of the aquifer. A considerable amount of testing occurs during the construction of an injection well. Water samples are collected during drilling to correlate water quality with depth and to identify underground sources of drinking water. Drill cuttings are collected to establish site-specific lithology, and pumping tests may be performed to locate production zones. Geophysical logs are run to help identify aquifer characteristics and interpret other field data. Field testing is used for determining preliminary water quality parameters such as chlorides, conductivity, temperature, and pH. Because concentrate has been classified as industrial waste by EPA, deep injection wells are required to have a liner from the wellhead to the receiving aquifer in addition to the casings which protect successive geologic layers from leakage from lower layers. Figure 7.1 illustrates the geologic strata and well construction requirements.

7.5 Evaporation Ponds

Evaporation ponds are a practical disposal method in locations with relatively high evaporation rates, low precipitation rates, and sufficient inexpensive land area. Ponds are typically lined and protected from flooding. They are designed so that no liquid is discharged to surface water or groundwater; thus, no permits are required. However, local regulations may require leak detection or monitoring or underdrains to assure groundwater protection.

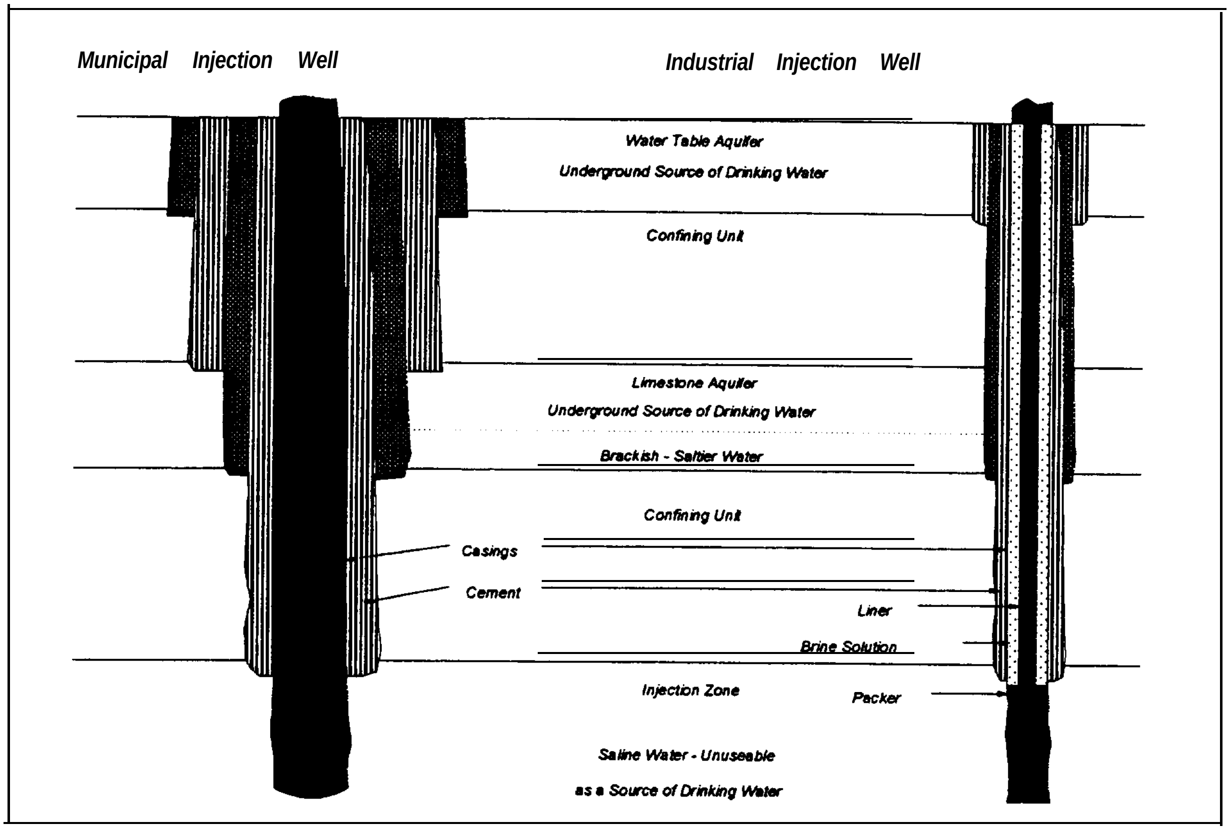


Figure 7.1 .-Typical municipal and industrial injection well construction. From *Florida Underground Injection Control Program Handbook,* FDER, 1983.

Residual solids must be periodically removed and transported to a landfill for disposal. The Solid Waste Disposal Act requires states to establish solid waste management plans governing non-hazardous waste disposal activities. USEPA regulations set minimum criteria for municipal solid waste landfills, and a state program must be at least as stringent. In general, landfill disposal rules apply to the residue remaining after evaporation of membrane system concentrates. If the residue is classified as a hazardous waste or toxic substance, RCRA and HMTA, or TSCA rules apply to its storage, transport, and disposal.

7.6 Solar Ponds

Solar ponds are heat traps that use high TDS water to trap heat from the sun which can then be used to generate electrical power or desalt water. A solar pond is 3 - 7 m deep, constructed in three distinct layers as on figure 7.2. The surface has a thin layer of fresh water. Convective currents created by wind and evaporation move vertically through this layer just as they do in any other body of water, though wind currents are kept to a minimum with wave suppression netting on the surface. Below the surface layer is a zone of increasing salinity, from near fresh at the top to about 20 percent (by weight) sodium chloride or other suitable salts, at the bottom. Normal convection currents cannot circulate water in this zone because of the density gradient. The bottom of the pond is the thermal storage zone. This

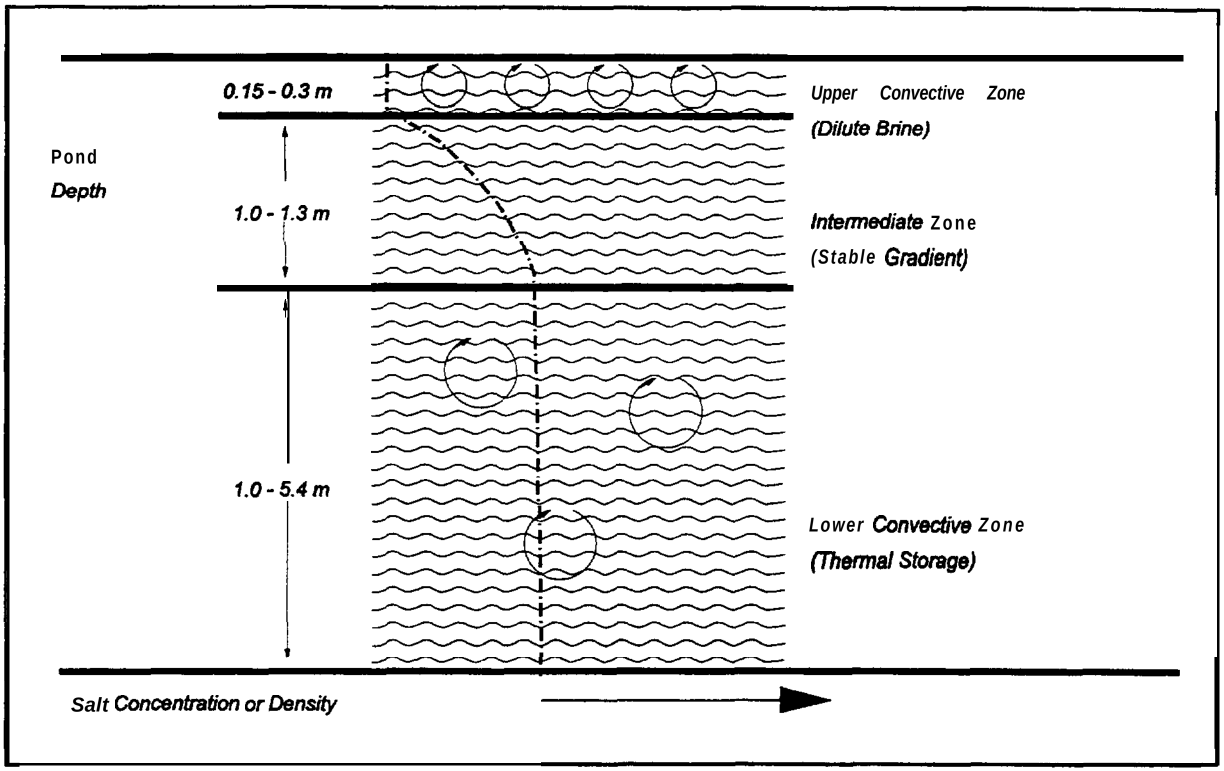


Figure 7.2.—Cross section of a solar pond (From Boegli et al., 1984).

layer has a uniform concentration as high as ten times that of seawater and is the heaviest layer in the pond. This layer has convective currents, but they do not circulate into the salinity-gradient layer above. The bottom of the pond is lined to prevent seepage into the groundwater.

The sun's energy penetrates the upper layers, warming the thermal storage zone just as in any lake, but the heat cannot circulate and dissipate into the atmosphere as it normally would. It becomes trapped in the dense lower layer of the pond. The temperature of the storage zone becomes as high as 70 to 100 ° C (160 to 2 12 °F) within a few months after the salinity gradient is established. Heat in the storage zone may be used as process heat, or may be converted to electricity with thermal effkiency of up to 15-20 percent. Even during the winter when the surface of the pond may be frozen, the storage zone will be hot enough to generate electricity. (Boegli et al., 1983).

The first solar pond in the U.S. to generate electricity was Reclamation's El Paso Solar Pond. The El Paso Solar Pond project began as a University of Texas at El Paso research project sponsored by Reclamation and the Bruce Foods Corporation and was later cost-shared with the Texas Energy and Natural Resources Advisory Council and the El Paso Electric Company. After some experimentation with process heat production, the solar pond was tied to the El Paso Electric Company power grid and began producing electricity on September 19, 1986. Figure 7.3 diagrams the process of electricity generation that is used at the El Paso

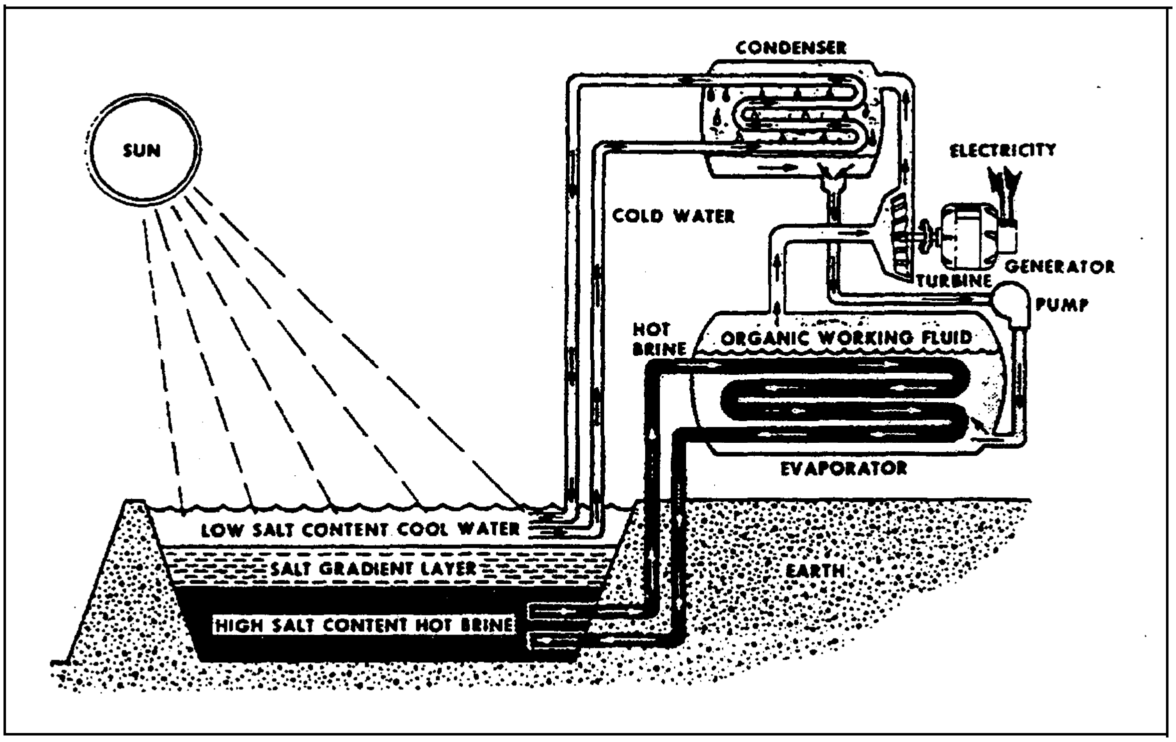


Figure **7.3.—Solar** pond heat transfer for power generation (From Boegli et al., 1984).

Solar Pond. The hot concentrate can be circulated to the food processing plant to be used as preheat for boiler feedwater as needed, to a 100-kW organic Rankine cycle power conversion module, or to a multistage flash desalting system. (Reid and Swift, 1987).

To build the solar pond, the storage layer concentrate must be concentrated to 10 times the salinity of sea water, or about 350,000 mg/L TDS. The normal RO recovery rate for sea water is 30-40 percent, which produces a TDS of only 50,000 to 67,000 mg/L. At another solar pond at the California Department of Water Resources, Los Banos, California, Demonstration Desalting Facility, concentrate was supplied by a three stage RO system designed to obtain 90 percent recovery from a feed water with 8,930 mg/L TDS. Thermal evaporation was used to increase recovery to 96 percent, resulting in a TDS of 223,250 mg/L. Further concentration of the RO concentrate could be carried out with heat energy from the solar pond, but to start with, energy was supplied by the local power utility (Smith, 1990). The Israelis have since developed an enhanced evaporation system that quickly concentrates the concentrate by spraying it down into evaporation ponds. This system is currently being used successfully at a saltworks.

The density gradient is important in maintaining a high temperature in the heat storage region. The density and temperature are monitored daily to ensure that the gradient is intact. If it is upset by physical mixing or operator error, the pond will become convective, transferring heat away from the bottom layer to the atmosphere. Gradient maintenance is accomplished periodically with diffusers that can be raised or lowered to inject concentrate at the right levels.

Solar ponds are a beneficial solution to the concentrate disposal problem. The initial block of power produced can be used to run pumps for the membrane system. Power production depends on the membrane system for a source of concentrate, so as long as replacement concentrate can be made from salt when needed, excess power can be sold to utilities to defray expenses. During the first thirty years of the project, the salt concentrate can be disposed of by constructing additional solar ponds. When the initial capital cost has been amortized, the excess power can be sold to pay for the disposal of excess RO concentrate by further concentration, deep well injection, or other means.

Chapter 8
CLEANING AND STORAGE

Each membrane manufacturer publishes specific instructions for cleaning and storing membranes when necessary. Rather than repeating these instructions, which may change with time, general cleaning and storage procedures are discussed in this chapter. Cleaning and storage are critical operations that can extend or shorten the life of membranes. It is important to the success of these touchy processes that the "why" is well understood.

8.1 Cleaning Procedures

Over time, even membranes with optimum pretreatment systems need to be cleaned. As mentioned earlier, this should be one of the action items triggered by a 10 percent change in AP, NPF, or salt rejection.

8.1.1 Cleaning Regimes

There are two basic types of cleaning regimes as described in table 8.1. One for organics and biological fouling, and one for scaling. Organic and biological films are best broken down with a high pH solution at the maximum temperature and pH limits for the membrane. Sodium hydroxide works well. Some suggest adding enzymes to help break down cellular matter, surfactants to help penetrate and dissolve the film, and a chelating agent to bind calcium ions. Calcium is an important component in extra-cellular polysaccharides (EPS, or extra-cellular polymeric slime) which is produced by well-established bacteria. EPS protects them from disinfectants and cleaning agents. Depleting the EPS of its calcium building blocks helps the cleaning solution penetrate the biofilm.

Table 8.1 .-Two tiered classification of membrane cleaning regimes

	Low temperature, low pH	High temperature, high pH
W/chelate	Soluble inorganic substances (carbonates, sulfates)	
W/detergent		Inorganic colloidal materials (i.e., silica, metallic hydroxides, etc.)
		Soluble organic substances (precursors of trihalomethanes)
W/chelate and/or enzymes		Microorganisms (bacteria, protozoa, fungi, algae)

Membranes with scaling problems are cleaned with a low pH, room temperature solution at low pressure and moderate flow rate for longer periods of time. Citric acid works well because it is also a chelating agent; it binds higher charged cations so that they cannot re-precipitate before they exit the system.

There may be layers of biological and scaling material in the same element. In this case, it will be necessary to use both solutions, one after the other, repeatedly. Since some precipitates are less soluble in hot water, the unheated, low pH solution should always be used first.

8.1.2 Equipment

Figure 8.1 is a diagram for a state of the art cleaning system. There are advantages to having a special skid just for cleaning, but if your system is designed with flexible plumbing connections, it is often possible to improvise. The most important components are:

- Mix tank with some method of mixing (volume should be twice the capacity of the vessel or stage that is to be cleaned)
- Pump with some method of flow control
- Flexible plumbing connectors (to isolate stages or vessels)
- Temperature control
- Cartridge filter
- Temperature/pH sensor (portable or hand-held model will do)

8.1.3 Generic Cleaning Process

No more than one stage should be cleaned at a time. Ideally, only one element would be cleaned at a time, but that would be far too labor intensive. As a compromise, each stage should be cleaned separately so that the cleaning solution needs to go through only one vessel before returning to the mix tank. When two stages are cleaned at once, the foulants from the first stage have to be moved all the way through the second stage. It is difficult enough to get them out of the first stage; the likelihood of getting them all through the second is very low.

The following cleaning procedures were generalized from FilmTec's "Cleaning and Disinfection" chapter of their Technical Manual.

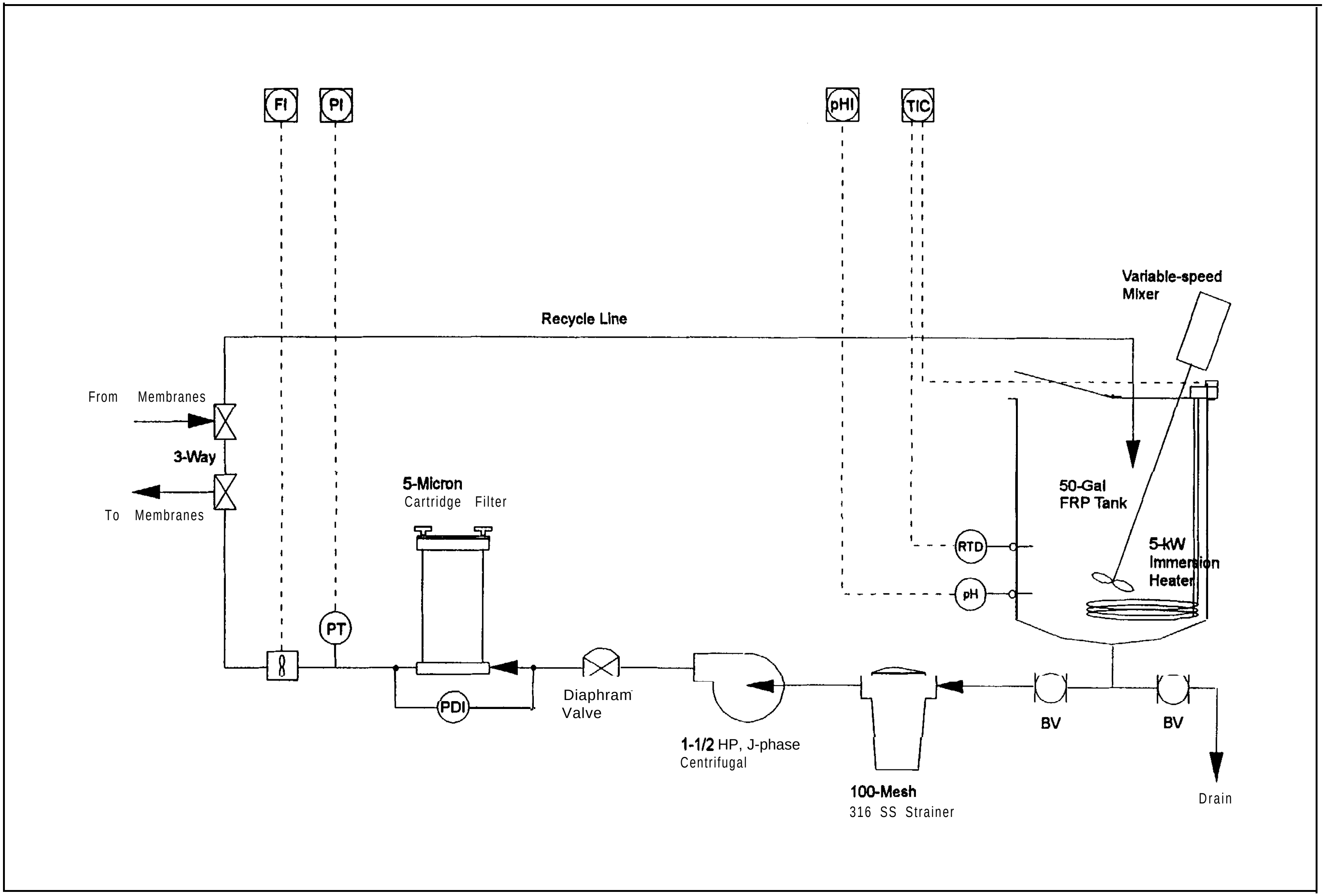


Figure 8.1 .-State-of-the-art cleaning system.

8.1.3.1 Preparation

- Isolate the vessel or stage that is to be cleaned.
- If pump does not have a variable speed drive, be sure that plumbing allows for flow bypass of pressure vessels so that flow can be controlled without increasing pressure to the membranes.
- Check temperature control system. If there is none, use sealed buckets of ice in the mix tank to keep the cleaning solution temperature from exceeding maximum limitations (may sound strange, but temperature is vitally important). See table 8.2 for maximum pH/temperature limits for Filmtec BW-30 membrane. Other membrane temperature limits can be found in chapter 10.

Table 8.2.—Maximum pH/temperature limits for Filmtec BW-30 membrane

pH range	2-10	1-11	1-12
Maximum temperature	50	35	30

- Temperature can be monitored by hanging a temperature probe in the tank. However, be sure not to let the water level drop below the level of the probe.
- Fill the mix tank with an adequate volume of RO product water, at least two times the volume of the piping and pressure vessel that is being cleaned.
- Begin mixing and warming the RO product water with a heater or 100 percent bypass.

8.1.3.2 Cleaning Solution Preparation

Bring the water to the appropriate temperature by mixing and heating or by pumping with 100 percent bypass back to the mix tank. Except perhaps in very cold climates, heating is necessary only during cleaning solution preparation. Once the cleaning cycle has begun, cooling is more important.

Add the cleaning chemicals slowly while mixing. If dry chemicals are added all at once, it will be more difficult to get them dissolved. Some dry detergents and thick surfactants are very difficult to get into solution. Be sure to allow enough mixing time to ensure the chemicals are dissolved. They will not work if they are not completely dissolved! They will likely cause further fouling and/or degradation, or else they will be removed by the cartridge filter and never get to the membranes. Take samples of the unused cleaning solution to compare before and after cleaning chemical analyses.

8.1.3.3 Fill System

Introduce the cleaning solution into the vessel or stage at a low flow rate with the concentrate stream going to drain. To determine when the system is completely filled, monitor the pH of the concentrate and divert the concentrate stream back to the mix tank when it matches that of the cleaning solution.

8.1.3.4 Low Flow Pumping

Pump the mixed, preheated cleaning solution at a low flow rate with solution recirculating to the mix tank for about 15 minutes or until the return solution looks cleaner. The back pressure valve should be completely open so that no permeate is produced. The objective of this cycle is to dislodge larger particles and remove them from the system. Table 8.3 lists sample cleaning flow rates for different sized elements.

8.1.3.5 Moderate Flow Pumping

Increase to moderate flow rate with the solution still recirculating for another 15 minutes with the back pressure valve still open. Check solution appearance and make note of any cloudiness or suspended solids accumulation.

8.1.3.6 *Soak*

Reduce flow to lowest possible level for the soak period. Soaking helps to dissolve stubborn films and precipitates. The membranes should soak for anywhere from one to fifteen hours, depending on the degree of fouling. It is important to control the temperature during long soaking periods. The cleaning solution should be kept at the optimum temperature. In most cases, this means keeping the temperature from rising too high; however, if the ambient temperature is low and the optimum temperature is high, heating may be necessary.

8.1.3.7 High Flow Pumping

After soaking, slowly increase the flow rate to the maximum allowable for your system. Filmtec’s specifications are given in table 8.3. The pressure drop must be monitored during the high flow cycle. Maximum pressure drops for each element type are listed in chapter 10. Flow rate should be increased slowly to flush out any large material loosened by soaking. Then the high flow may remove stubborn material and particles stuck in the spacer material.

Table 8.3.—Recommended feed flow rate per pressure vessel during cleaning

Cleaning cycle	Feed flow rate, Umin (gal/min)		
Module diameter	64 mm (2.5")	102 mm (4")	203 mm (8")
Soaking	2 (0.5)	3 (1)	15 (4)
Low flow	10 (2.6)	19 (5)	75 (20)
Moderate flow	15 (4)	30 (8)	100 (26)
High flow	20 (5)	38 (10)	150 (40)

8.1.3.8 Flush and Sanitize

Take samples of the spent cleaning solution for chemical analysis. Drain tank and piping. Rinse the tank, and refill it with RO permeate. Flush out the system at a moderate flow rate with the concentrate stream diverted to drain. After flushing, refill the tank and add a sanitizer recommended by the membrane manufacturer. Sodium bisulfite or chlorine can be used with CA membranes. Formaldehyde or hydrogen peroxide can be used with thin film composites. Use caution with these chemicals, though; high concentrations may cause damage.

After the cleaning process, some cleaning solution will still be on the permeate side of the membrane; so, after putting the clean stage back on line, divert product to drain until quality is acceptable.

8.1.4 Which Cleaning?

It is often possible to determine what has happened to a membrane system by noting changes in the NPF, AP, and rejection for each stage over time. Table 8.4 describes the symptoms of major types of fouling and which cleaning regime is called for. Generally, a decrease in NPF in the first stage means particulate fouling. Particulates could be mineral, vegetable, or animal in nature. Both spiral wound and HFF membranes serve as excellent cartridge filters • most particulates are trapped in the first couple of elements. The cleaning strategy indicated for front end fouling of this sort is a high pH/high temperature regime.

Precipitative fouling, or scaling, occurs at the tail end of the system where the feed stream is at its highest concentration. Effects are a decrease in rejection and NPF and an increase in AP. A low temperature/low pH cleaning regime is needed for scaling. Low temperature means normal operating temperature. Since calcium carbonate is less soluble at higher temperatures, it is best to use a low pH/low temperature cleaning solution first if more than one type of cleaning will be used.

Table 8.4.—Foulant characterization based on chemical composition

	Soluble inorganic substances	Soluble organic substances	Colloid materials (water insoluble inorganic compounds: silica, iron hydroxides, etc.)	Biological materials (bacteria, algae, fungi, etc.)
Causes	Over saturation Presence of crystallization centers	Humic and fulvic acids natural to surface waters Lack of adequate pretreatment Over utilization	Over utilization Inadequate sedimentation period	Inadequate pretreatment Inadequate flow through module • dead spaces Hydrophobic attraction between cell and membrane surfaces Production of extracellular polymeric substances Cell fimbriae may help attach bacteria to molecular matrix of the membrane
Effects	Decrease In salt rejection in end stages Increase in pressure drop in end stages Decrease Normalized Permeate flow (NPF) Scale formation on membrane surface or In bulk w/subsequent deposition Formation of 'salt bridge' facilitating protein adsorption High concentration at membrane surface can cause denaturation of proteins which then are more of a fouling problem	Formation of H bonds on contact w/ membrane Partial diffusion through membrane • dependent on degree of branching	Gel formation on membrane surface Decrease in NPF Decrease in salt rejection Symptoms most likely to appear in last stage	Decrease in NPF Initial increase in salt rejection Increase in pressure drop “Symptoms most likely to appear in first stage Accumulation of byproducts of metabolism Eventual detedoration of the membrane resulting in a decrease in rejection Decrease in flow at membrane surface can exacerbate concentration polarization phenomena
Prevention	Softening Acidification Use of chelating agents	Ultrafiltration Coagulation/sedimentation	Softening	Prefiltration Use of surfactants during normal operation has been shown to prevent bacterial attachment Reduce recovery rate
Remediation	Low pH w/ chelate Normal operating temperature “Soak cycle Physical methods: ultrasound, magnetic, hydrodynamic	“Same as Colloids	High pH High temperature High flow rate “Detergent	“Same as Colloids Use of enzymes has been shown to help loosen biofilm

Adapted from Saran, A.A. 1990.

• * DHP: Paul, 1993.

8.1.5 Passive Cleaning

Passive cleaning is a milder form of cleaning that does not use chemicals or even a separate pump. If, say at 5 percent change in AP, NPF, or rejection, passive cleaning is performed, it may be possible to extend the time between chemical cleanings. The following are some procedures that can have a beneficial effect without voiding the manufacturers warranty.

- Turn off the system for an hour. Osmotic pressure will draw product water through the membrane to the concentrate side. This can help lift foulants off the surface. If live bacteria are present, stopping flow could encourage a growth spurt, so this method should be used only when disinfectants are present.
- Reduce back-pressure for a short time. Permeation rate should drop, and flow rates through the system should increase. This changes flow patterns through the system and can disrupt films that may be in process of taking root. It also washes away the high concentration layer right at the membrane surface and, in doing so, could slow the scaling process.
- Perform a high flow, low pressure flush with product water. RO product water is very aggressive in dissolving precipitates. The change in flow pattern will also have the effect described above.

These are experimental procedures at this point, but there is nothing in them that manufacturers would object to. For more information about fouling and membrane cleaning philosophy, see Chapman Wilbert, 1997.

8.2 Membrane Storage

Membrane systems work best when they operate continuously. It is unavoidable to have occasional shutdowns, though. When a membrane system is to be down for a few days, flush the process water from the system and replace it with RO product water pumped into the system at low pressure.

If the concentrated process water and pure product water are left in the membrane vessels, the difference in concentration across the membranes will cause osmotic flow from the product side to the concentrate side. This can be beneficial in lifting foulants from the membrane surface in small quantities, but too much back flow can destroy the glue lines.

For thin film composite membranes, the storage water should be oxidant-free. Otherwise, the membranes will degrade over time. CA membranes should be stored in chlorinated water, though. While the danger of biological growth is the same for both types of membrane, damage from oxidation would be worse than biofouling for thin film composites.

If the membranes are to be stored for longer than about a week, a biostat, such as glutaraldehyde or sodium metabisulfite, should be added as a preservative. The solution must be changed periodically, and the elements should be kept in a temperature-controlled environment away from direct sunlight.

Alternative methods for storing membranes for long periods have been investigated at the Yuma Desalting Plant. While some have proven successful, the final report of findings is still under review. For more information, check the Water Treatment and Engineering Group's intemet library home page at (http://www.usbr.gov/water/water.html).

Chapter 9
PROCESS DESIGN

9.1 Introduction

The purpose of this section is threefold: first, to illustrate a simple solvent flux modeling approach for estimating the overall performance of a pressure-driven, liquid filtration, membrane module; second, to demonstrate an element-by-element analytical procedure for the preliminary design of membrane desalting systems; and third, to present several vendor-supplied computer programs for estimating membrane system performance. More accurate and complex models and algorithms are available in the literature, and the reader is referred to those for detailed design projects (Hwang and Kammermeyer, 1975).

9.2 Solvent Flux Models

Simple models are useful for exploring the effects of changes in membrane characteristics and module operating conditions. A “simple approach” means the solution can be obtained without complex numerical computation techniques. Ideally, a spreadsheet program could be used to perform “what if’ case studies. The reader is cautioned at the outset that, because of the assumptions required in a simple model, significant differences will probably exist between the actual performance and the estimate.

The model presented here assumes perfectly mixed feed and permeate compartments, with consideration of concentration polarization. This model is the simplest to use and provides a conservative estimate of module performance. In the general case, solution of the design problem requires an iterative (trial and error) method, but in special circumstances, an algebraic solution can be obtained.

This module model can be used for RO, NF, UF, and MF with appropriate modifications. The overall material balances and boundary layer aspects are essentially the same for all these applications. The definition of the flux, membrane resistances, and the concentrations can be modified to suit the specific type of filtration. Therefore, this discussion will begin with the specific flux models for the different types of filtration. The overall material balance equations will be developed using the RO flux model and boundary layer estimation techniques. Finally, a sample design problem will be presented.

Figure 9.1 is a diagram of the dynamics of a pressure-driven filtration process. A gel layer is shown to illustrate how solute build up at the membrane surface affects the flow of solvent through the membrane.

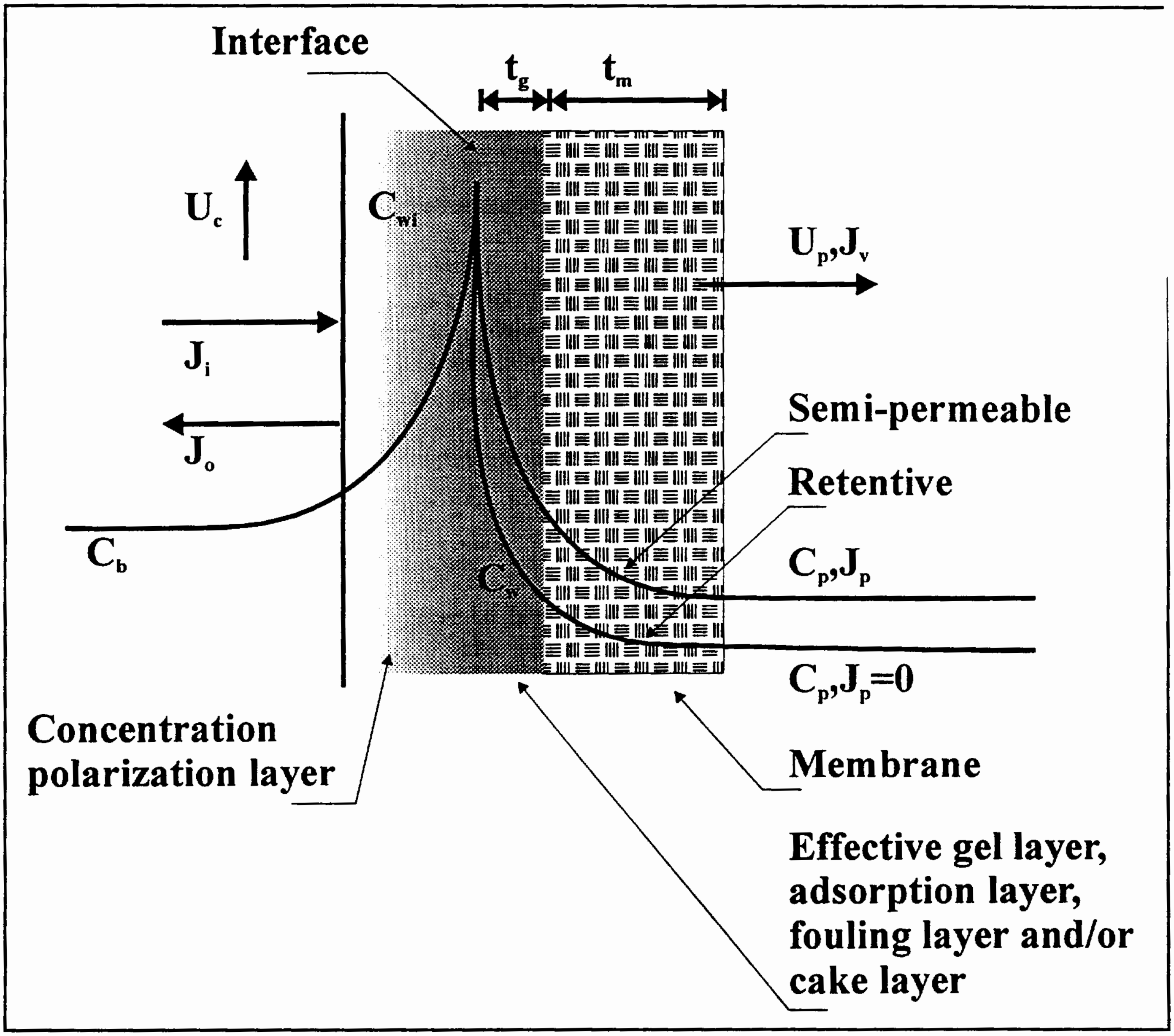


Figure 9.1.—Schematic of pressure-driven membrane processes.

C_b	Concentration of solute in solvent	J_i	Permeation of solvent into the polarization layer
C_{wi}	Maximum concentration of solute within polarization zone	J_o	Permeation of solvent out of the polarization layer back into main feed stream
C_w	Minimum concentration of solute within the gel layer	J_v	Permeation of solvent through the membrane
C_p	Concentration of solute in product stream	J_p	Permeation of solute through the membrane
U_c	Cross flow velocity of feed stream	t_g	Thickness of gel layer
U_p	Velocity of product stream	t_m	Thickness of membrane

The following discussion of flux models will be limited to the phenomenological description of the solvent flux. Models also exist for the solute flux, which can be used to provide greater detail in process design, but are beyond the scope of this particular presentation.

9.2.1 Reverse Osmosis

Permeation of water through the membrane, J_v, in RO is given by the following equation:

$$J_v = \frac{P_v}{t_m}(\Delta P - \Delta\pi) \qquad \text{Eq. 9.1}$$

Where:

P_v = specific water permeability,
t_m = membrane thickness,
ΔP = applied, transmembrane mechanical pressure (at a specific point),
$\Delta\pi$ = actual osmotic pressure gradient between C_w and C_p (at a specific point).

This case assumes that no gel layer with thickness t_g is present; therefore $C_w = C_{wi}$. The intrinsic rejection of salt, $R°$, is less than 1. The term P_v/t_m is often referred to as the coefficient of water transport, which can be determined from experimental data by dividing the observed water flux per unit area of membrane by the net applied pressure (ΔP - $\Delta\pi$).

9.2.2 Nanofiltration, Ultrafiltration, and Microfiltration

The most general form for water permeation in NF, UF, and MF is:

$$J_v = \frac{\Delta P - \Delta\pi}{\left[\frac{t_m}{\hat{P}_v} + \frac{t_g}{\hat{P}_g}\right]\mu} \qquad \text{Eq. 9.2}$$

Where:

$\hat{P}_v$ = specific water conductance of clean membrane,
t_m = membrane thickness,
$\hat{P}_g$ = specific conductance of the 'g' layer,
t_g = thickness of effective adsorption, gel, fouling and/or cake layers,
ΔP = applied, transmembrane, mechanical pressure (at a specific point),

$\Delta\pi$ = actual osmotic pressure gradient based on C_w and C_p (at a specific point),
μ = shear viscosity of the fluid passing through the membrane.

Often further lumping of parameters is done, so that the form:

$$J_v = \frac{\Delta P - \Delta\pi}{\mu(R_m + R_g)} \qquad \text{Eq. 9.3}$$

is used.
Where:
R_m = clean membrane resistance,
R_g = additional resistance from gels, cakes and adsorption (fouling).

For NF and UF, both $\Delta\pi$ and R_g can be significant. For MF, $\Delta\pi$ is likely to be negligible unless significant filtration of smaller molecules occurs because of the 'g' layer. This case often occurs in biotechnology (fermentation) applications.

For all these filtration processes, $\Delta\pi$ and R_g will vary along the length of the module and can vary with time.

9.2.3 Estimation of Permeability Parameters

Solvent conductance through the membrane, $\hat{P}_v$, and solvent conductance through the gel layer, $\hat{P}_g$ can be estimated from some simple geometric viewpoints.

- For a group of pores containing laminar flow:

$$\hat{P}_v \text{ or } \hat{P}_g = \frac{\pi}{8} \sum_{i=0}^{i} n_i d_i^4 \qquad \text{Eq. 9.4}$$

Where:
n_i = number of pores with diameter "i" per unit area,
d_i = diameter of pore "i".

- For packed particles (Carmen-Kozeny rigid particles):

$$\hat{P}_v \text{ or } \hat{P}_g = \frac{\epsilon^3}{5(1-\epsilon)^2}\left(\frac{V_p}{S_p}\right)^2 \qquad \text{Eq. 9.5}$$

Where:

ϵ = void fraction,
V_p = volume of particles,
S_p = surface area of particles.

9.2.4 Estimation of Osmotic Pressure

Osmotic pressure is a thermodynamic property and can be obtained by measurement of a variety of properties. The osmotic pressure of solutions can be related to the freezing points or vapor pressures of those solutions. Data on freezing points and vapor pressure is available in data handbooks. For dilute, ideal solutions the van't Hoff equation provides a means of calculating the osmotic pressure, by:

$$\pi = CRT \qquad \text{Eq. 9.6}$$

Where:

C = ionic concentration,
R = gas constant,
T = temperature in °K.

Even though the van't Hoff equation is frequently not valid for practical problems, it is often useful to fit experimental or calculated data to a linear form, such as:

$$\pi = aC \qquad \text{Eq. 9.7}$$

The relationship $\pi = aC^n$, where a and n are constants ($n > 1$ and often ~2), is more general and can also be used for engineering design (Wijmans, et al., 1984). The general flux equation contains the term $\Delta\pi$, which is the difference in osmotic pressure on the two sides of the membrane based on the concentrations C_w and C_p. Therefore, using the linear form for π from above, and the definition of the intrinsic rejection:

$$R^\circ = \frac{C_w - C_p}{C_w} \qquad \text{Eq. 9.8}$$

results in:

$$\Delta\pi = a(C_w - C_p) = aR^\circ C_w \qquad \text{Eq. 9.9}$$

To calculate the osmotic pressure from vapor pressure data, the following relationship can be used (Reid, 1966):

$$\pi \cdot v_{solvent} = RT \ln \left[\frac{P^{v}_{solvent}}{P^{v}_{solution}} \right] \qquad \text{Eq. 9.10}$$

Where:

$v_{solvent}$ = partial molar volume of the solvent (m^3/mole),
R = gas constant (8.314 Pa m^3 $mole^{-1}$ $°K^{-1}$),
T = solution temperature, °K,
$P^{v}_{solvent}$ = vapor pressure of pure solvent,
$P^{v}_{solution}$ = vapor pressure of solution with concentration C.

Often, freezing point data of solutions are more widely available. To calculate the osmotic pressure from freezing point data, the following relationship can be used (Reid, 1966):

$$\pi = \frac{(T_f^v - T_f)\,\Delta H_f\,T}{v_{solvent} T_f\,T_f^v} \left(\frac{R_{press}}{R_{energy}} \right) \qquad \text{Eq. 9.11}$$

Where:

T_f^v = freezing point of pure solvent,
T_f = freezing point of solution with concentration c,
ΔH_f = latent heat of freezing of pure solvent,
T = actual temperature for estimating π, °K,
R_{press} = gas constant in pressure units,
R_{energy} = gas constant in energy units,
$v_{solvent}$ = specific volume in cm^3/gmole.

Useful values:

R_{press}/R_{energy} = 3.14×10^{-6} cm^3 cmHg/kcal
v_{water} = 18.095 cm^3/gmole

9.2.5 Module Mass Balance

For the perfectly mixed feed and permeate model, with concentration polarization, the method developed by Rao and Sirkar (1978) and illustrated by Wankat (1990) is used. Figure 9.2 illustrates the values of importance.

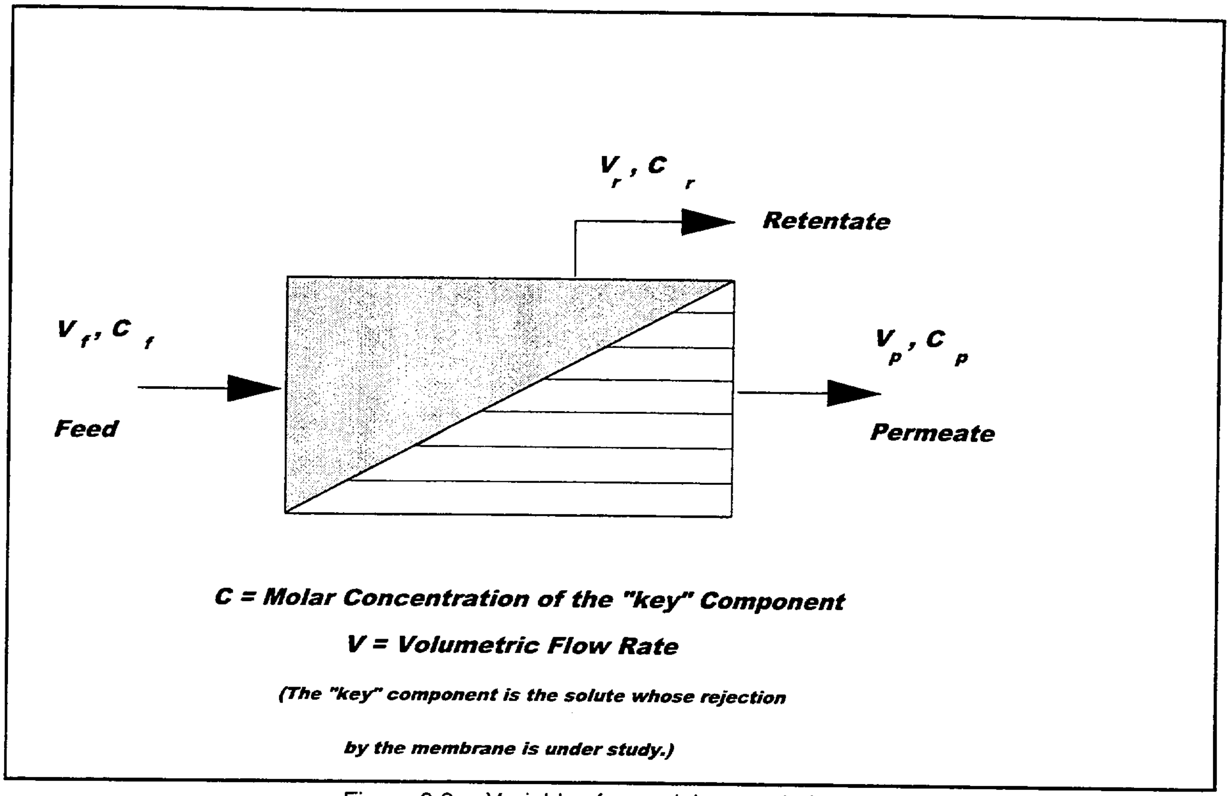


Figure 9.2.—Variables for module mass balance.

External mass balance:

$$V_f = V_r + V_p \qquad \text{Eq. 9.12}$$

define Θ = stage cut or recovery factor = V_p/V_f

External key component balance:

$$V_f C_f = V_r C_r + V_p C_p \qquad \text{Eq. 9.13}$$

The overall volumetric flux through the membrane is J_v, therefore $V_p = J_v * A$, where A is the membrane area.

Let C_w = boundary layer concentration at the membrane interface caused by concentration polarization. Furthermore, assume that no gel formation occurs.

Because the feed side is perfectly mixed, $C_r = C_b$, where C_b = bulk concentration.

Then:

$$C_b = \frac{C_f - \theta C_p}{1-\theta} \qquad \text{Eq. 9.14}$$

The definition of the intrinsic rejection of a membrane, $R°$, $R° = 1 - C_p/C_w$. This is different from the apparent rejection, $R_a = 1 - C_p/C_f$ The intrinsic rejection is a characteristic of the membrane. The apparent rejection is determined by the operating conditions.

From the material balance $C_r = C_b$ is defined by:

$$C_r = C_b = \frac{C_f}{(1-\theta) + \dfrac{\theta(1-R°)\exp(J_v/k)}{R° + (1-R°)\exp(J_v/k)}} \qquad \text{Eq. 9.15}$$

Then, from the simple boundary layer model for concentration polarization and assuming that $R°$ is constant, the following relationship for C_p is obtained:

$$C_p = \frac{C_r(1-R°)\exp(J_v/k)}{R° + (1-R°)\exp(J_v/k)} \qquad \text{Eq. 9.16}$$

With k = the boundary layer mass transfer coefficient. The variable 'k' is obtained via correlations that usually assume that $J_v \ll U_c$, where U_c is the average cross flow velocity.

Sources of k:

- Supplied by the module manufacturer
- Estimated from literature and the assumed conditions in the module

Estimation of k based on correlations:

Definitions:

$$\text{Reynolds No.} \quad Re = \frac{d\bar{U}_c\rho}{\mu} \qquad \text{Eq. 9.17}$$

$$\text{Schmidt No.} \quad S_c = \frac{\mu}{\rho D} \qquad \text{Eq. 9.18}$$

d = representative channel or tube dimension for flow (i.e., diameter)
$\bar{U}_c$ = average cross flow velocity
ρ = density
μ = shear viscosity

D = solute diffusivity
ω = stirrer speed (radians/s)

For turbulent flow in tubes:

$$k = \overline{U_c}\,\frac{0.0791}{2}\,Re^{-1/4}Sc^{-2/3} \qquad \text{Eq. 9.19}$$

Re > 20,000 (in general)
Re > 2000 (in UF)

or

$$k = 0.023\,\frac{D}{d}\,Re^{0.83}Sc^{1/3} \qquad \text{Eq. 9.20}$$

For turbulent flow in batch stirred vessels:

$$k = 0.0443\,\frac{D}{d}\left(\frac{\rho}{\mu D}\right)^{1/3}\left(\frac{\omega\mu d^2}{\rho}\right)^{3/4} \qquad \text{Eq. 9.21}$$

d = vessel diameter

For laminar flow in round tubes with diameter d:

$$k = 1.295\left(\frac{2\cdot\overline{U_c}D^2}{d\,L}\right)^{1/3} \qquad \text{Eq. 9.22}$$

L = distance along tube length

For laminar flow between parallel plates spaced at 2h:

$$k = 1.177\left(\frac{\overline{U_c}D^2}{hL}\right)^{1/3} \qquad \text{Eq. 9.23}$$

Remember that in laminar flow, k varies along the length of module. The correlations represent what exists at a given point in the fluid's path. Often, one uses an average value for the overall material balance.

In summary:

For the perfectly mixed feed and permeate module, the following relationships exist:

[1] The retentate concentration, C_r:

$$C_r = \frac{C_f}{(1-\Theta)+\dfrac{\Theta(1-R^\circ)\exp(J_v/k)}{R^\circ+(1-R^\circ)\exp(J_v/k)}} \qquad \text{Eq. 9.24}$$

[2] The wall concentration, C_w:

$$C_w = C_r\left[\frac{\exp(J_v/k)}{R^\circ+(1-R^\circ)\exp(J_v/k)}\right] \qquad \text{Eq. 9.25}$$

[3] The permeate concentration, C_p:

$$C_p = \frac{C_r(1-R^\circ)\exp(J_v/k)}{R^\circ+(1-R^\circ)\exp(J_v/k)} \qquad \text{Eq. 9.26}$$

[4] Estimate J_v using the appropriate flux model, i.e., for RO:

$$J_v = \left(\frac{P_v}{tm}\right)(\Delta P - aR^\circ C_w) \qquad \text{Eq. 9.27}$$

[5] Estimate k using appropriate boundary layer correlation and physical dimensions of the system.

In the case of other filtration processes, the same general form can be followed, with the major exception being the definition of J_v (Refer to section on flux models).

This group of relationships is non-linear because of the exponential term and must be solved iteratively (using successive substitutions). This approach can be readily accomplished using a spreadsheet model or in a general programming language model (i.e., Fortran, C, etc). For an interative solution of RO with concentration polarization, we define J_v as follows:

If $J_v/k \ll 1$, then the problem can be solved explicitly, because the approximation that $\exp(J_v/k) \sim 1 + J_v/k$, leads to an algebraic solution as follows:

$$a'J_v^2 + b'J_v + c' = 0 \qquad \text{Eq. 9.28}$$

Where:

$$a' = \frac{1-R^\circ}{k} \qquad \text{Eq. 9.29}$$

$$b' = 1-\Theta R^\circ - \left[\frac{P_v\,(1-R^\circ)\Delta P}{t_m\,k}\right] + \left[\frac{P_v C_f\,aR^\circ}{t_m\,k}\right] \qquad \text{Eq. 9.30}$$

$$c' = \left[\frac{P_v}{t_m}\,C_f\,aR^\circ\right] - \left[\Delta P(1-\Theta R^\circ)\,\frac{P_v}{t_m}\right] \qquad \text{Eq. 9.31}$$

The "a" in the equations for b' and c' is the constant from the linear model for osmotic pressure.

The general form for the roots of a quadratic is:

$$J_v = \frac{-b' \pm \sqrt{b'^2 - 4a'c'}}{2a'} \qquad \text{Eq. 9.32}$$

Once J_v is known, then the apparent rejection and the recovery rate, C_p and C_r, can be calculated.

9.2.6 Sample Design Problem

Using the perfectly mixed model of RO with concentration polarization and brackish water with TDS = 1500 mg/L, expressed as NaCl, the following steps outline the method for estimating percent recovery and salt rejection.

$$C_f = 0.0375 \text{ mol/L} \qquad \text{Eq. 9.33}$$

1) Develop linear relationship for osmotic pressure using example literature data shown in table 9.1.

2) Intrinsic rejection of membrane defined to be $R^\circ = 0.987$.

Table 9.1.—Data for determining the relationship between osmotic pressure and concentration

π (kPa)	Conc (mol/L)
100	0.0375
300	0.125
700	0.3
π = 2286 * C	a = 2.286 (kPa m^3/mole)

3) Available transmembrane pressure, $\Delta P = 1550$ kPa (225 lb/in^2).

4) Spiral wound module with:

   L = 1.016 m (40 in),
   h = 2.5×10^{-3} m,
   A = 28 m^2 (300 ft^2) per module.

5) Pure water permeability was determined to be 28 m^3/day (1.16×10^{-5} $m^3\ m^{-2}\ sec^{-1}$) at 1550 kPa. Therefore, estimate $P_v/t_m = 1.16 \times 10^{-5}/1550 = 7.47 \times 10^{-9}\ m^3\ m^{-2}\ sec^{-1}\ kPa^{-1}$.

6) Estimate mass transfer coefficient *k*:

- Need estimate of U_c
- Pump characteristic curve will give 130 L/min @ 1550 kPa this rating is within specifications of the membrane module)
- Need to approximate cross sectional area for flow in the spiral wound module: $A_x = (A * h)/2L = (28*0.0025)/(2*1.016) = 0.034\ m^2$
- Average $U_c = \overline{U}_c = V_f/A_x$
- V_f= 130 L/min = $2.17 \times 10^{-3}\ m^3/s$
- $\overline{U}_c$:

$$\frac{2.16x10^{-3}}{3.44x10^{-2}} = 0.0629 m/s\ (6.3 cm/s) \qquad \text{Eq. 9.33}$$

- Approximation for Re:

$$Re \cong \frac{h\overline{U}_c\rho}{\mu} = \frac{(2.5x10^{-3})(6.29x10^{-2})(10^6)}{1} \qquad \text{Eq. 9.34}$$

Re ≅ 157, probably laminar flow

- Use k correlation for laminar flow in a flat channel
- $D_{NaCl} = 1.2 \times 10^{-9}$ m²/s
- Calculate k

$$k \cong 1.177 \left(\frac{6.29x10^{-2}\ (1.2x10^{-9})^2}{2.5x10^{-3}\ (1.016/2)} \right)^{1/3} \qquad \text{Eq. 9.35}$$

Note: use $L/2$ so that k is some average value along the channel

- $k \cong 4.88 \times 10^{-6}$ m/s.

7) Check pure water value of J_v (1.16 x 10^{-5} m/s) versus k (4.88 x 10^{-6} m/s) versus U_c (6.29 x 10^{-2} m/s). Because J_v is significantly less than U_c and greater than k, the correlations used thus far are reasonable. But J_v/k is not <<1. Therefore, the approximation that $\exp(J_v/k) \cong 1 + J_v/k$ should not be used in this case.

8) Set up an iterative solution using the feed flow divided by the area as a first estimate of J_v. An initial C_w, C_p, C_r, and recovery are calculated with the first estimate of J_v. Subsequent iterations use the J_v calculated according to equation 9.27 to refine the estimates. See table 9.2 for an example of the spreadsheet output. Figures 9.3 and 9.4 show the changes in Jv, recovery rate, the intrinsic rejection, and the apparent rejection.

9) This model shows:

- J_v converges to 8.09 x 10^{-6} m/s. This value represents about 70 percent of the pure water flux.

- The recovery is 10.34 percent.

- The apparent rejection of salt, defined as $R_a = (1 - C_p/C_f) * 100$ is 92.96 pct.

9.3 Membrane System Design

In the previous paragraphs of this section, a solvent flux modeling approach was presented for estimating the performance of a single membrane desalting module (element). A comparatively simple method will now be presented for the preliminary design of membrane "systems" employing several elements arrayed in a manner to achieve a desired permeate recovery and total production. Also discussed will be the availability of several vendor-supplied computer programs for estimating system performance.

Figure 9.5 shows a diagram of a 2:1 array with N elements installed in each of three pressure vessels. The two pressure vessels in the 1st stage receive pretreated feed water. Based on

Table 9.2.—Sample spreadsheet for estimating spiral wound RO module design performance

Pure water permeability (m^3/sec)	3.24E-04
Feed flow (m^3/s)	2.17E-03
Transmembrane pressure (kPa)	1550
Area (m^2)	28
Length (m)	1.016
Channel height (m)	2.50E-03
Cf (mol/L)	0.0375
a (Pa m^3/mol)	2286
Calculated parameters determined by configuration and operating conditions.	
Jv (m/s) 1st pass	**1.157E-05**
Pv/tm ($m^3 m^{-2} s^{-1} kPa^{-1}$)	7.47E-09
Average Uc (m/s)	0.0629
Renolds number	157
k (m/s) for laminar flow in flat channel	4.88E-06

Final performance parameters	
Jv (m/s)	8.09E-06
Fraction of pure water permeability	69.86%
Recovery is	10.34%
Apparent rejection =(1-Cp/Cf)*100	92.96%

Solving the design equations	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Jv/k	2.37	0.95	2.02	1.35	1.85	1.51	1.75	1.59	1.71	1.62	1.68	1.64
Recovery	0.1496	0.0602	0.1275	0.0853	0.1164	0.0954	0.1107	0.1001	0.1077	0.1023	0.1061	0.1034
Intrinsic Rejection, R'	0.987	0.987	0.987	0.987	0.987	0.987	0.987	0.987	0.987	0.987	0.987	0.987
Apparent Rejection R_a	0.858	0.965	0.898	0.947	0.914	0.938	0.921	0.933	0.925	0.931	0.927	0.930
Cw (mol/L)	0.4104	0.1013	0.2951	0.1522	0.2489	0.1787	0.2275	0.1925	0.2170	0.1995	0.2119	0.2031
Cp (mol/L)	0.0053	0.0013	0.0038	0.0020	0.0032	0.0023	0.0030	0.0025	0.0028	0.0026	0.0028	0.0026
Cr (mol/L)	0.0432	0.0398	0.0424	0.0408	0.0420	0.0412	0.0418	0.0414	0.0417	0.0415	0.0416	0.0415
Jv Theoretical (m/s)	4.66E-06	9.87E-06	6.60E-06	9.01E-06	7.38E-06	8.56E-06	7.74E-06	8.33E-06	7.92E-06	8.21E-06	8.00E-06	8.15E-06
EXP (J_v/k)	10.711	2.597	7.550	3.867	6.335	4.536	5.780	4.885	5.512	5.064	5.380	5.155

Solving the design equations	**13**	**14**	**15**	**16**	**17**	**18**	**19**
Jv/k	1.67	1.65	1.66	1.65	1.66	1.66	1.66
Recovery	0.1054	0.1040	0.1050	0.1043	0.1048	0.1044	0.1047
Intrinsic Rejection, R'	0.987	0.987	0.987	0.987	0.987	0.987	0.987
Apparent Rejection R_a	0.927	0.929	0.928	0.929	0.928	0.929	0.928
Cw (mol/L)	0.2093	0.2049	0.2080	0.2058	0.2074	0.2062	0.2070
Cp (mol/L)	0.0027	0.0027	0.0027	0.0027	0.0027	0.0027	0.0027
Cr (mol/L)	0.0416	0.0415	0.0416	0.0416	0.0416	0.0416	0.0416
Jv Theoretical (m/s)	8.05E-06	8.12E-06	8.07E-06	8.11E-06	8.08E-06	8.10E-06	8.09E-06
EXP (J_v/k)	5.314	5.201	5.281	5.224	5.264	5.236	5.256

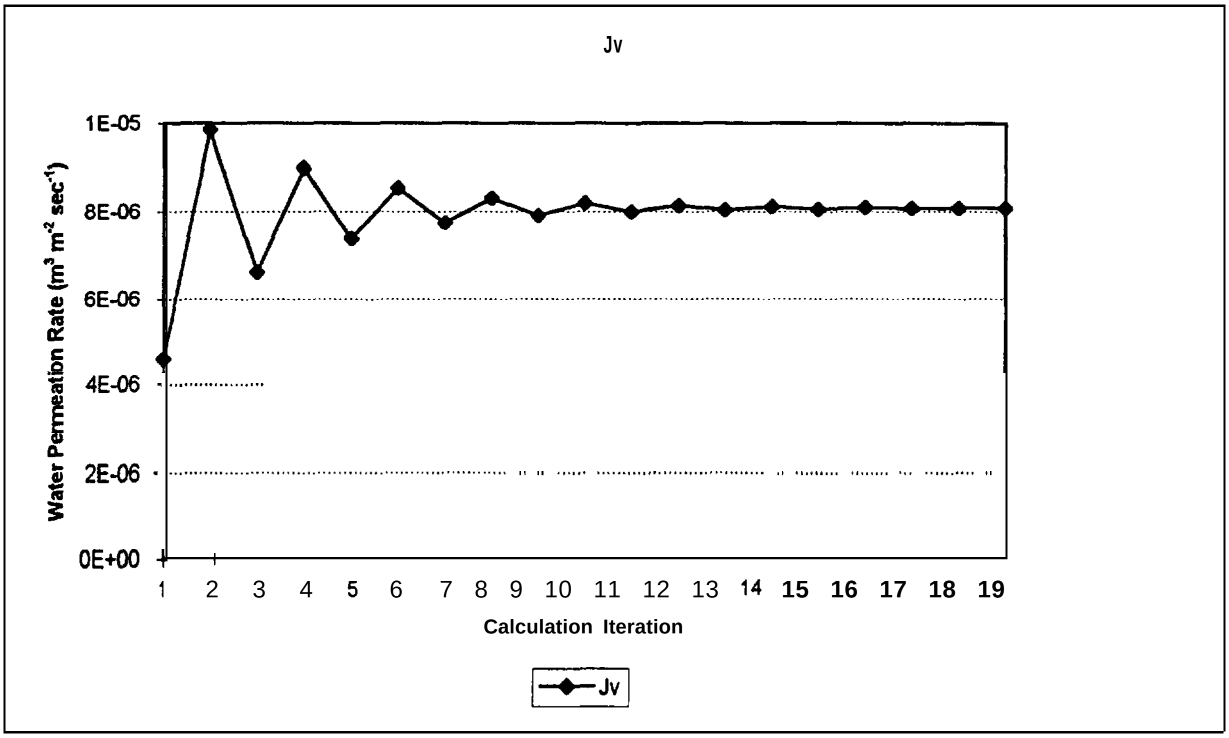


Figure 9.3.—**Iterative** solution for water permeation rate, Jv, every iteration has more accurate estimates of recovery rate, product, and wall concentration. The solution is found when the calculated values level out at the most accurate estimate.

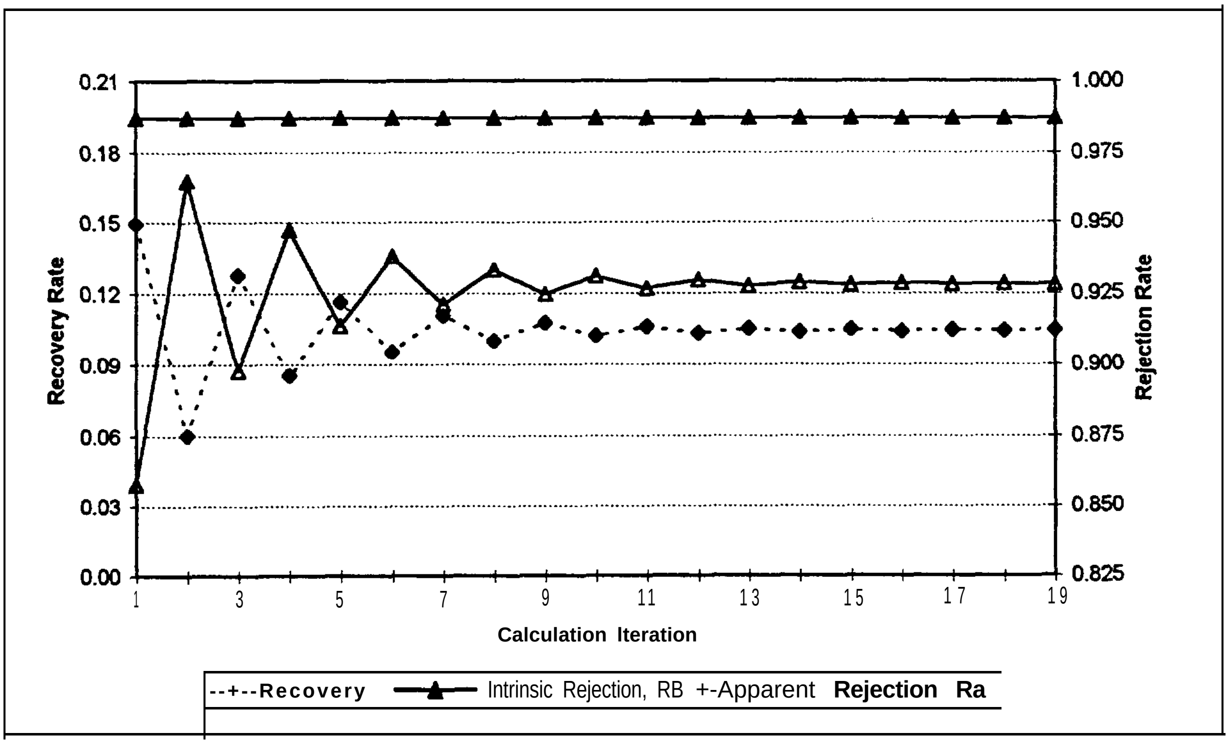


Figure **9.4.—Oscillation** of recovery rate and the apparent rejection rate as the solution becomes more accurate.

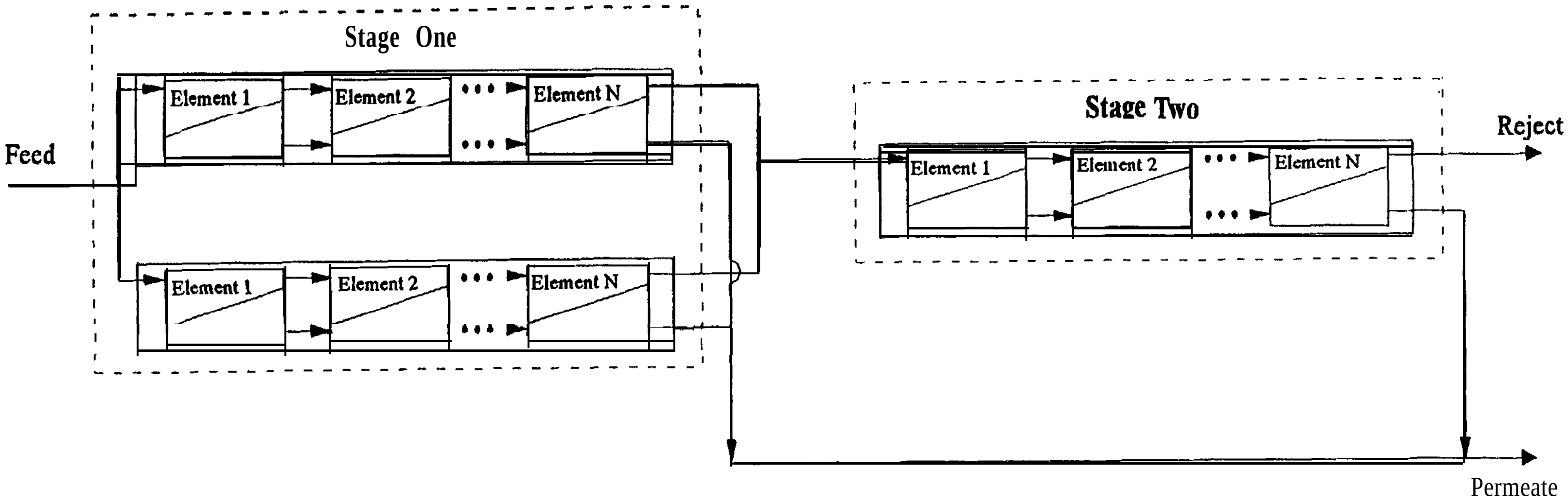


Figure 9.5.—Diagram of a two-stage membrane system.

system operating conditions, i.e., feed pressure, temperature, etc., and the type and number of elements loaded in the vessels, a percentage of the feed will diffuse through the membranes and will be collected as permeate. The feed flow to each succeeding element in a vessel will be lower, by the amount of permeate developed, and the concentration will be higher. Concentrate (reject) flows from the two 1” stage vessels are then combined and proceed to a 2nd stage where additional desalting takes place. Permeates from both stages are collected and represent the total system production. The array and number of elements per vessel are generally dictated by feed water quality, element hydraulic limitations, and economic considerations. A 2:1 array, as shown in figure 9.5, can be used to recover about 70 to 75 percent of the system feed, while a 4:2:1 array can recover up to 90 percent. Several other configurations are also commonly used.

9.3.1 System Equations

A generalized process diagram of four lead elements of a membrane desalting system is presented on figure 9.6. This type of diagram is used to develop material balance and other equations around individual elements to accommodate system design calculations. Although only four elements are shown on this figure, the computations apply to the entire system. Definitions of the symbols used in the diagram are as follows:

F_{fi} - Feed water flow
F_{pi} - Permeate flow
F_{ri} - Reject flow
C_{fi} - Feed water concentration
C_{pi} - Permeate concentration
C_{ri} - Reject concentration
P_{fi} - Feed pressure
P_{pi} - Permeate back pressure
P_{ri} - Reject pressure
ΔP - Differential pressure

Where i represents the element number.

Applicable equations describing the first element on figure 9.6 would include:

Mass Balance - $F_{f1}C_{f1} = F_{p1}C_{p1} + F_{f2}C_{f2} = F_{p1}C_{p1} + F_{r1}C_{r1}$ Eq. 9.36

Differential Pressure - $\Delta P = P_{f1} - P_{f2}$ Eq. 9.37

Average Feed Pressure - $P_{avg} = (P_{f1} + P_{f2})/2$ Eq. 9.38

Average Salt Concentration - $C_{avg} = (C_{f1} + C_{f2})/2$ Eq. 9.39

Net Driving Pressure - $NDP = P_{f1} - P_o - P_{p1}$ Eq. 9.40

Where P_o is the element osmotic pressure.

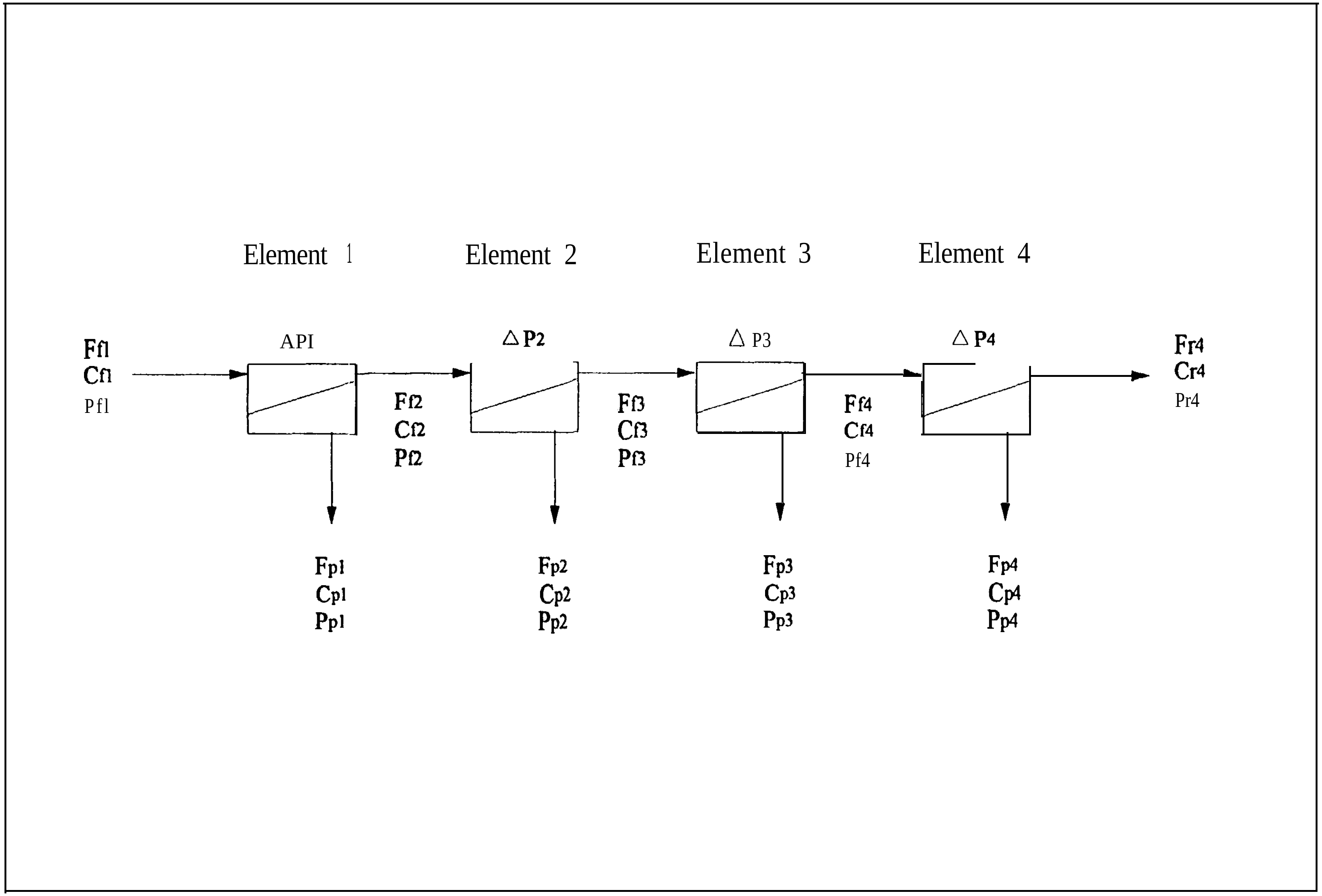


Figure **9.6.—Generalized** process diagram of the four lead elements of 1st stage.

9.3.2 Element-By-Element Analysis

A fairly good estimate of system performance can be determined using an element-by-element analysis developed by David H. Paul (Paul, 1993). As the name implies, this procedure steps through the proposed membrane desalting system one element at a time, calculating individual productivities based on net driving pressure (*NDP*). A spreadsheet program such as Excel or Lotus l-2-3 is typically used to perform the calculations.

The following sample problem, used to illustrate the element-by-element design procedure, is based on treatment calculations for a high-nitrate (64 mg/L NO,-) and slightly brackish [599 mg/L TDS (total dissolved solids)] groundwater in southern California. Treatment objectives were to remove sufficient NO_3^- and TDS to allow blending with filtered well water to optimize net recovery while still meeting state and Federal drinking water standards and the Santa Ana River Basin water quality objective of 400 mg/L TDS. The well in question has an average production rate of 1500 gal/min at a wellhead temperature of 22.5 °C. Based on pretreatment calculations and total product requirements, it was decided to design a low-pressure RO system to desalt 1125 gal/min at a recovery rate of 75 percent and then blend the resulting permeate with 375 gal/min for filtered well water, for a total net recovery (permeate plus filtered well water) of slightly in excess of 81 percent.

The required system permeate flow would then be determined as follows:

$$F_p = (75/1\ 00)\ I\ I25 = 844\ gal/min \qquad \text{Eq. 9.41}$$

The first step in the design process is to select a membrane element and element configuration required to produce the total amount of permeate desired. This determination will be affected by the achievable recovery of the pretreated water to avoid scaling by sparingly soluble salts. FilmTec's BW30-8040 thin-film composite membrane element (now designated as BW30-330) was chosen for evaluation. The characteristics of this element are shown at the top of table 9.3. Based on a design element productivity of 5.2 gal/min (well/softened water supplies), an initial estimate of the total number of elements required would be:

$$Number\ of\ elements = \frac{844\ gallmin}{5.2\ \ gallmin} = 162 \qquad \text{Eq. 9.42}$$

Table 9.3.—Sample spreadsheet for an element-by-element analysis

* * *

Project Identification:
Wellhead Treatment Study
Van Buren Well No. 8

System Design Parameters:

RO Feed Flow, gal/min •	1125
Blend Flow. gal/min •	375
Desired Recovery, % •	75
Req'd Permeate, gal/min •	844
System Amy •	24:12
Feed Temperature, oC •	22.5

Membrane Element Description/Characteristics:

Manufacturer •	FilmTec
Model/Type •	BW30-330, Thin-Film Composite
Max Operating Pressure, psi •	600
• Mas Feed Flow. gal/min •	62
*Max Permeate Flow. gal/min •	5.2
*Max Element Recovery, % •	19
Average Cl- Rejection, % •	99

*Well Water/Softened Water Supply

* * *

FIRST STAGE	#Elem	Elem #2	Elem #3	Elem #4	Elem #5	Elem #6
Feed Flow, gal/min	46.88	42.29	37.81	33.62	29.49	25.48
Feed Concentration, mg/L	599.0	663.4	739.9	832.7	948.1	1096
Feed Pressure, prig	187.0	180.6	175.3	171.0	167.4	164.6
Osmotic Pressure, psi (est.)	5.99	6.63	7.40	8.33	9.48	10.96
Net Driving Pressure. psi	181.0	174.0	167.9	162.6	157.9	153.7
Permeate Flow. gal/min	4.59	4.41	4.26	4.13	4.01	3.90
Permeate Conc., mg/L	6.0	6.6	7.4	8.3	9.5	11.0
Element Recovery, %	9.79	10.4	11.2	12.3	13.6	15.3
Pressure Drop, prid	6.36	5.31	4.37	3.53	2.80	2.15
Reject Flow, gal/min	42.29	37.87	33.62	29.49	25.48	21.59
Reject Pressure, psi8	180.6	175.3	171.0	167.4	161.6	162.5
Reject Concentration, mg/L	bb3.4	739.9	832.7	948.1	1096	1292

	Vessel	1d Stage
Permeate Flow. gal/min	25.29	607.1
Rcjcct Flow. gal/min	21.59	518.0
Permeate Conc., mg/L	8.04	8.04
Reject Concentration, mg/L	1292	1292
Recovery, %	53.96	53.96
Pressure Drop, psid	24.52	24.52

SECOND STAGE	Elem #1	Elem #2	Elem #3	Elem a4	Elem #5	Elem #6
Feed Flow, gal/min	43.17	39.50	36.01	32.67	29.47	2b.41
Feed Concentration, mg/L	1292	1410	1546	1702	1885	2101
Feed Pressure, psi8	157.5	151.9	147.1	143.0	139.6	136.7
Osmotic Pressure, psi (est.)	12.92	14.10	15.46	17.02	18.85	21.01
Net Driving Pressure, psi	144.6	137.8	131.6	126.0	120.7	115.7
Permeate Flaw. gal/min	3.67	3.50	3.34	3.20	3.06	2.93
Permeate Conc., mg/L	12.9	14.1	15.5	17.0	18.8	21.0
Element Recovery, %	8.49	8.85	9.27	9.78	10.39	11.11
Pressure Drop, psid	5.59	4.80	4.08	3.44	2.88	2.37
Reject Flow. gal/min	39.50	36.01	32.67	29.47	26.41	23.48
Reject Pressure, psig	151.9	147.1	143.0	139.6	136.7	134.3
Reject Concentration, mg/L	1410	1546	1702	1885	2101	2361

	Vessel	2nd Stage	System
Permeate Flow. gal/min	19.7	236.3	843.4
Reject Flow, gal/min	23.5	281 .7	281.7
Permeate Conc., mg/L	16.4	16.4	10.4
Reject Concentration, mg/L	2361	2361	2361
Recovery, %	45.6	45.6	75.0
Pressure Drop, psid	23.2	23.2	52.7

Blended Product Flow. gal/min	1218
Blended Product Conc., mg/L	192
Net Recovery, %	81.2

Assuming 6 elements per pressure vessel:

$$Number\ of\ pressure\ vessels = \frac{162\ elements}{6\ elements/vessel} = 27 \qquad \text{Eq. 9.43}$$

Using this number of pressure vessels in a 2-stage system (based on 75 percent recovery), the resulting array would be 18:9. The feed flow per first stage pressure vessel would be:

$$Feed\ flow\ per\ vessel = \frac{1125\ gal/min}{18\ vessels} = 62.5\ galfmin \qquad \text{Eq. 9.44}$$

This feed flow rate is higher than the maximum recommended feed flow of 62 gal/min for the element (refer to top of table 9.3), and could result in a higher fouling potential. Therefore, it was decided that a more conservative 24: 12 array would be used which would result in a 1st stage feed flow per vessel of 46.9 gal/min.

Referring to table 9.3, element 1 of the 1” stage, entries for feed flow and concentration would be 46.9 gal/min and 599 mg/L, respectively. Feed pressure will be varied, as will be explained later, to obtain a system recovery of 75 percent. An estimate of the osmotic pressure gradient can be calculated as:

$$P_o = \frac{599\ lb/in^2\ feed\ pressure}{100} = 5\ 99\ lb/in^2 \qquad \text{Eq. 9.45}$$

The net driving pressure (*NDP*) for each element can be calculated by subtracting the osmotic pressure gradient and permeate back-pressure (assumed to be zero in this case) from the element feed pressure. For the first stage lead element, this would be:

$$NDP = P_f - P_o - P_p = 187 - 5.99 - 0 = 181\ lb/in^2 \qquad \text{Eq. 9.46}$$

This *NDP* is then used to determine element productivity (permeate flow):

$$F_p = Design\ Productivity\ \frac{NDP_{actual}}{NDP_{design}} \qquad \text{Eq. 9.47}$$

$$= 5\ 2\ \frac{181\ lb/in^2}{205\ lb/in^2} = 4.59\ gallmin \qquad \text{Eq. 9.48}$$

Where NDP_{design} is equal to the design pressure of 225 lb/in^2 minus an estimated osmotic pressure of 20 lb/in^2 (design concentration of 2,000 ppm divided by 100). This design information, based on manufacturer's test data, is available in FilmTec's technical bulletin for the BW30-330 element.

Permeate concentration is calculated using the published average salt passage of 1.0 percent (average salt rejection of 99 percent) for the membrane element:

$$C_p = Feed\ TDS\ x\ \%\ Salt\ Flux = 599\ x\ 0.01 = 6.0\ mg/L \qquad \text{Eq. 9.49}$$

Percent recovery for the element is:

$$Element\ Recovery = \frac{F_p}{F_f}\ 100 = \frac{4.59}{46.88}\ 100 = 9.79\ \% \qquad \text{Eq. 9.50}$$

For this example, the interstage pressure drop was assumed to be 5 lb/in*, and element pressure drops were computed based on the following formula provided by FilmTec:

$$\Delta P_{Elem} = 0.01\ [(F_f + F_r)/2)]^{1.7} \qquad \text{Eq. 9.51}$$

$$= 0.01\ [(46.88 + 42.29)/2]^{1.17} = 6.36\ lb/in^2$$

Where:

F_f = element feed flow, gal/min,
F_r = element reject flow, gal/min.

The element reject flow, pressure, and concentration are determined as follows:

$$F_r = F_f - F_p = 46.88 - 4.59 = 42.29\ gal/min \qquad \text{Eq. 9.52}$$

$$P_r = P_f - \Delta P_{elem} = 187 - 6.36 = 180.6\ lb/in^2 \qquad \text{Eq. 9.53}$$

$$C_r = \frac{(F_f \times C_f) - (F_p \times C_p)}{F_c} \qquad \text{Eq. 9.54}$$

$$= \frac{(46.88 \times 599) - (4.59 \times 6.0)}{42.29} = 664\ mg/L$$

This completes the calculations for the lead element of the 1” stage. As shown in table 9.3, the first element reject flow, pressure and concentration are now used as input (feed) parameters for the second element. Once the equations for the first element are input to the spreadsheet, they can be copied and pasted to the remaining columns to complete the analysis.

Vessel, stage, and system summaries can also be included in the spreadsheet, as shown in table 9.3. Notice, at the bottom of the table, that the overall membrane system recovery (prior to blending) is 75 percent, which is equivalent to the original requirement. This was achieved iteratively by adjusting the value of the permeate flow for element 1 of the 1st stage.

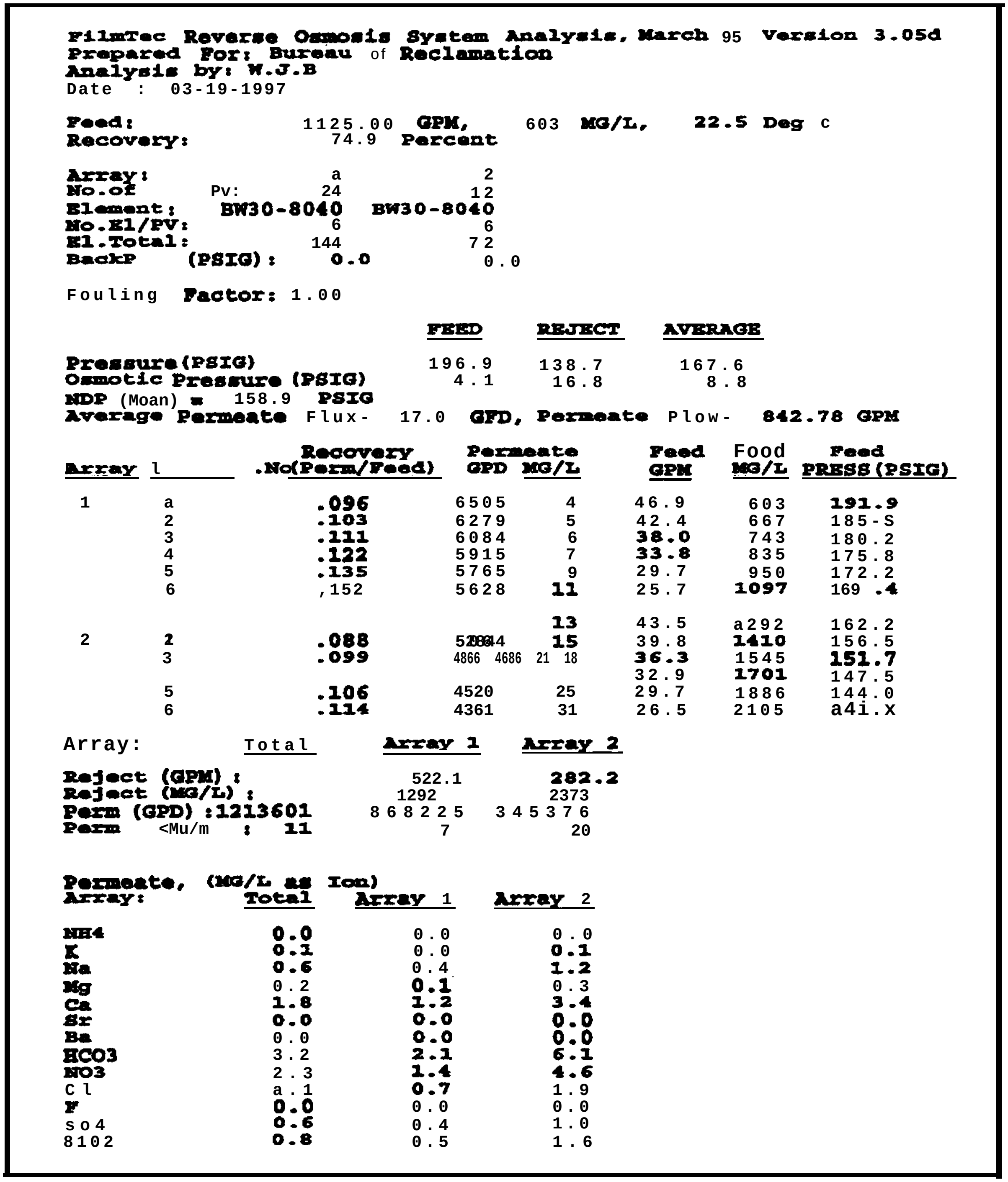
9.3.3 Manufacturers’ Computer Programs

A number of manufacturers provide software for predicting the performance of their membranes for site-specific design applications. Among these are Dow FilmTec (ROSA), Fluid Systems (ROPRO), Hydranautics (RODESIGN), Desal (Solutions), and DuPont Permasep (CPP). All operate in the Windows environment except for the Desal program which is DOS based. These programs are not for final design specifications, and each of the manufacturers includes a liability disclaimer stating that final designs should be reviewed by an appropriate applications engineer. These programs provide estimates of water production and quality from given input parameters and are useful for examining alternative system configurations.

An example of a Dow FilmTec computer analysis using Reverse Osmosis System Analysis (ROSA) software is shown in table 9.4. Input parameters include the feed water analysis, design temperature, system configuration, percent recovery, fouling factor (indicates degree of membrane fouling; set at 1 .OO for new membranes), and desired feed <or> permeate flow rate. Interstage booster pumping and concentrate recirculation options are also available. As shown on table 9.4, the results of the analysis include system flows and pressures; average permeate flux; ion concentrations for the feed, reject and permeate streams [for the system and individual stages (referred to as arrays in the program)];estimated permeate pH; and scaling calculations.

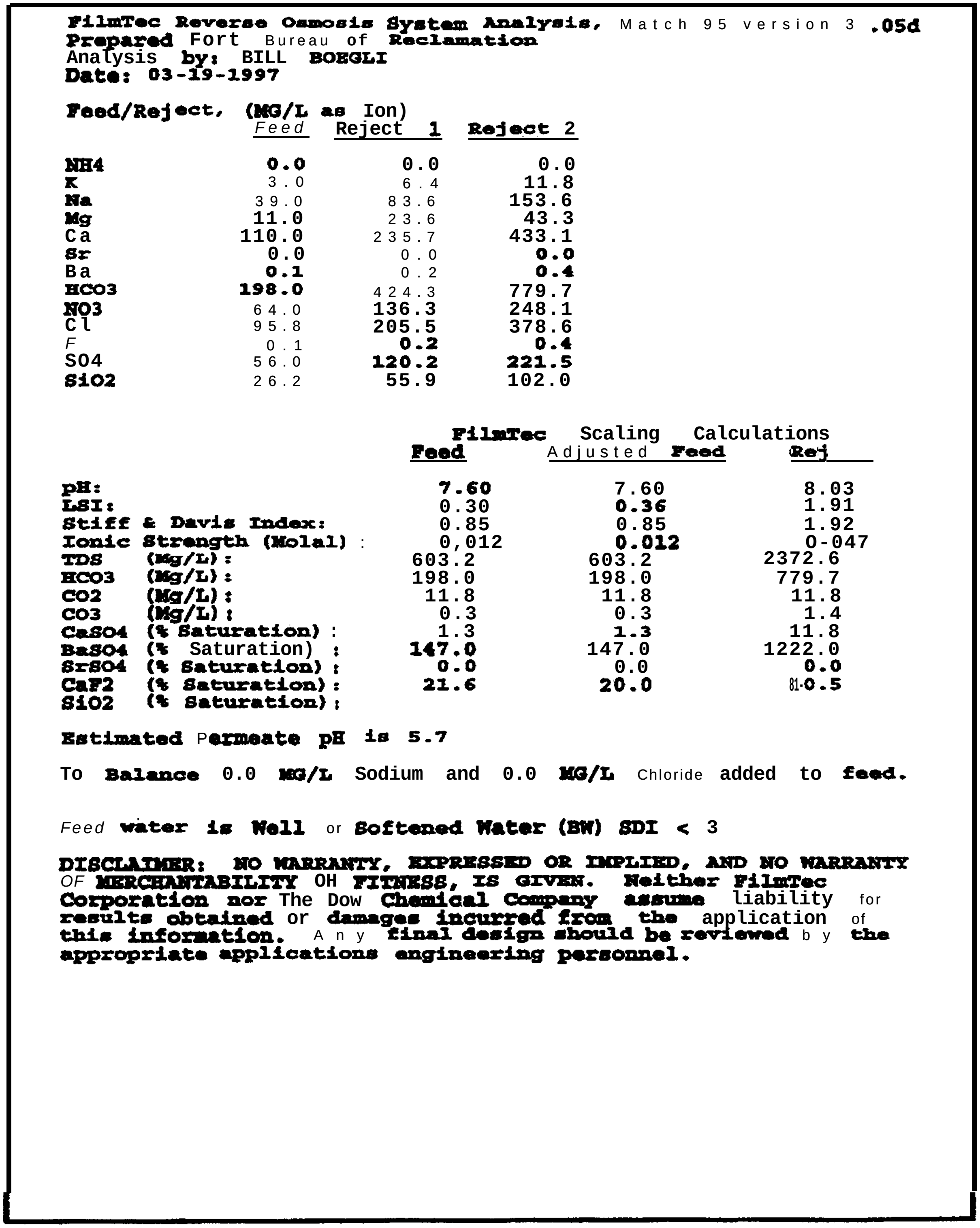
This computer analysis was run with the same input parameters used for the earlier element-by-element analysis (refer to figure 9.4). The results of the computer and manual (spreadsheet) methods compare very favorably as shown in table 9.5.

Table 9.4.—Sample printout of a FilmTec RO system analysis using ROSA Version 3.05d software



```
FilmTec Reverse Osmosis System Analysis, March 95  Version 3.05d
Prepared For: Bureau of Reclamation
Analysis by: W.J.B
Date  :  03-19-1997

Feed:                 1125.00  GPM,     603  MG/L,    22.5  Deg  c
Recovery:               74.9  Percent

Array:                   a           2
No.of         Pv:        24          12
Element:        BW30-8040   BW30-8040
No.El/PV:                 6           6
El.Total:               144          72
BackP     (PSIG):       0.0          0.0

Fouling  Factor: 1.00

                                  FEED      REJECT     AVERAGE

Pressure(PSIG)                   196.9      138.7      167.6
Osmotic Pressure (PSIG)            4.1       16.8        8.8
NDP (Moan) =   158.9  PSIG
Average Permeate  Flux-   17.0  GFD, Permeate  Plow-   842.78 GPM

                  Recovery        Permeate        Feed    Food    Feed
Array l       .No(Perm/Feed)      GPD  MG/L       GPM     MG/L  PRESS(PSIG)

 1      a          .096           6505    4       46.9     603    191.9
        2          .103           6279    5       42.4     667    185-S
        3          .111           6084    6       38.0     743    180.2
        4          .122           5915    7       33.8     835    175.8
        5          .135           5765    9       29.7     950    172.2
        6          ,152           5628   11       25.7    1097    169 .4

                                         13       43.5    a292    162.2
 2      2          .088           520844 15       39.8    1410    156.5
        3          .099      4866  4686 21  18    36.3    1545    151.7
                                                  32.9    1701    147.5
        5          .106           4520   25       29.7    1886    144.0
        6          .114           4361   31       26.5    2105    a4i.x

Array:           Total       Array 1      Array 2

Reject (GPM) :                522.1        282.2
Reject (MG/L) :              1292          2373
Perm (GPD) :1213601        868225       345376
Perm   <Mu/m   :  11             7            20


Permeate,  (MG/L as Ion)
Array:           Total       Array 1      Array 2

NH4               0.0          0.0          0.0
K                 0.1          0.0          0.1
Na                0.6          0.4          1.2
Mg                0.2          0.1          0.3
Ca                1.8          1.2          3.4
Sr                0.0          0.0          0.0
Ba                0.0          0.0          0.0
HCO3              3.2          2.1          6.1
NO3               2.3          1.4          4.6
Cl                a.1          0.7          1.9
F                 0.0          0.0          0.0
so4               0.6          0.4          1.0
8102              0.8          0.5          1.6
```

Table 9.4.—Sample printout of a FilmTec RO system analysis using ROSA Version 3.05d software (continued)



```
FilmTec Reverse Osmosis System Analysis, Match 95 version 3 .05d
Prepared Fort Bureau of Reclamation
Analysis by: BILL BOEGLI
Date: 03-19-1997

Feed/Reject, (MG/L as Ion)
                    Feed     Reject 1    Reject 2

NH4                  0.0        0.0         0.0
K                    3.0        6.4        11.8
Na                  39.0       83.6       153.6
Mg                  11.0       23.6        43.3
Ca                 110.0      235.7       433.1
Sr                   0.0        0.0         0.0
Ba                   0.1        0.2         0.4
HCO3               198.0      424.3       779.7
NO3                 64.0      136.3       248.1
Cl                  95.8      205.5       378.6
F                    0.1        0.2         0.4
S04                 56.0      120.2       221.5
SiO2                26.2       55.9       102.0


                               FilmTec   Scaling   Calculations
                            Feed       Adjusted Feed      Rej

pH:                          7.60          7.60           8.03
LSI:                         0.30          0.36           1.91
Stiff & Davis Index:         0.85          0.85           1.92
Ionic Strength (Molal) :     0,012         0.012          0-047
TDS     (Mg/L):            603.2         603.2         2372.6
HCO3    (Mg/L):            198.0         198.0          779.7
CO2     (Mg/L):             11.8          11.8           11.8
CO3     (Mg/L):              0.3           0.3            1.4
CaSO4   (% Saturation) :     1.3           1.3           11.8
BaSO4   (%  Saturation) :  147.0         147.0         1222.0
SrSO4   (% Saturation) :     0.0           0.0            0.0
CaF2    (% Saturation):     21.6          20.0          81·0.5
SiO2    (% Saturation):

Estimated Permeate pH is 5.7

To Balance 0.0 MG/L Sodium and 0.0 MG/L Chloride added to feed.

Feed water is Well or Softened Water (BW) SDI < 3

DISCLAIMER: NO WARRANTY, EXPRESSED OR IMPLIED, AND NO WARRANTY
OF MERCHANTABILITY OH FITNESS, IS GIVEN. Neither FilmTec
Corporation nor The Dow Chemical Company assume liability for
results obtained or damages incurred from the application of
this information. Any final design should be reviewed by the
appropriate applications engineering personnel.
```

Table 9.5.—Comparison of results from FilmTec's ROSA program and the element-by-element spreadsheet analysis

	1st Stage		2nd Stage		System	
Parameter	Spread sheet	ROSA	Spread Sheet	ROSA	Spread sheet	ROSA
Permeate flow, gal/min	607	603	236	240	843	843
Reject flow, gal/min	518	522	282	282	282	282
Permeate concentration, mg/L	8.0	7	16.4	20	10.4	11
Reject concentration, mg/L	1,292	1,292	2,361	2,373	2,361	2,373
Recovery, %	540	53.6	45.6	45.9	75.0	74.9
Pressure drop, psig (stage AP only; excludes interstage connections)	24.5	24.7	23.2	23.5	52.7	53.2

Chapter 10
PRODUCT REVIEW

The last 5 years have seen the number of membrane products skyrocket. Even since this publication was begun, there have been many new products released. Fortunately, the membrane manufacturers have also made great strides onto the information superhighway! It was possible to update the product lists as the rest of this publication was being reviewed. Table 10.1 lists the membrane manufacturer's addresses, phone numbers, web sites, and contacts. Many of them have complete product information on the Internet.

10.1 Product Listings

The tables of membrane specifications are categorized as seawater, brackish water, nanofiltration, ultrafiltration, and microfiltration. The products were categorized as the manufacturer had them. Seawater membranes are those with high NaCl rejection rates designed to withstand 7 to 8 MPa (approximately 1000 lb/in^2). Brackish water membranes have rejections from 99.7 percent to 85 percent with maximum pressure ratings of 2 to 4 MPa (275 to 600 lb/in^2). Nanofiltration membranes have rejections between 20 percent and 85 percent, with a wide range of pressure ratings from 1.7 to 7 MPa (250 to 1000 lb/in^2). Ultrafiltration membranes are those characterized by a Molecular Mass Cutoff, and microfiltration membranes are those with micron pore sizes.

For seawater, brackish water RO, and NF membranes, there are three categories of information supplied by the manufacturer and one category of calculated parameters.

- **Physical Dimensions.** The physical dimensions given in tables 10.2-10.6 are the membrane composition, and configurations, total module length, and diameter. MF membrane information includes pores size, bubble point, and foam point if available.

- **Performance Under Test Conditions.** Production rate and rejection are the primary performance parameters for RO and NF membrane. UF and MF membrane performance specifications give only the productivity. Percent variability is included for RO membrane, but was left off the NF tables since few manufacturers provided the information. The test conditions are very important here as the performance is directly related to the conditions. Whenever available, the standard test conditions for each company are provided. A complete description of test conditions should include feed flow rate, test solution composition, temperature, pH, operating pressure, and recovery rate. Usually, the feed flow rate is not provided. It must be calculated from the recovery rate and production rate. A few companies persist in stating their products are tested with "San Diego Tap Water" or "Typical Brackish Water." While this may be convenient for people in San Diego, it is not very meaningful to the rest of the world.

UF and MF manufacturers are consistent in testing their products with “clean water,” which is fine since performance of these types of membrane is not significantly affected by dissolved solids concentration.

- **Operating Limits.** RO and NF operating limits are given for feed flow, concentrate flow and/or recovery, operating pressure, pressure drop across the module, SDI, turbidity, temperature, pH during operation and cleaning cycles and exposure to oxidants. Optimum pH is provided by one company. UF and MF membrane limitations are given only for pressure, temperature, and pH.

- **Calculated Parameters.** The calculated parameters are the water (A) and salt (B) transport coefficients under the specified test conditions calculated as described in chapter 4 (RO and NF), equations 4.4 and 4.5, and also the productivity per unit module volume. The productivity per unit volume is simply the claimed production rate under standard test conditions divided by the module volume calculated from the given dimensions. It must be emphasized that the values of these parameters are highly dependent on the operating conditions. For that reason, A and B are not listed for the NF membranes that were not tested with a di-valent solution. The B value for these membranes would be inordinately high and incomparable to the B values for those calculated from test data using divalent test solutions such as $MgSO_4$. They are included in figure 10.1 for comparison.

  There are no calculated parameters for UF and MF membranes. One could calculate a water transport coefficient and productivity per unit volume, but it would not be meaningful for anything but clean water. The best way to evaluate UF and MF membrane is to compare them side by side on the same test solution using the best operating conditions for each membrane.

Data for these tables has been gathered from manufacturers specification sheets in hard copy or from web sites. It has been reviewed by the company’s representative for accuracy. As many different manufacturers as possible were included. If someone has been left out, we apologize. The paucity of data from those outside the United States is due to the difficulty in obtaining data from such companies and not entirely due to our preference for supporting domestic industry.

In using these tables, remember that performance is always different with “real” water. All of the membranes listed were tested with salt solutions made from clean water, RO product water in most cases, and food grade NaCl or $MgSO_4$ (if not laboratory grade). Complex solutions of salts, organic compounds, colloids, suspended solids, and wildlife will have different rejection and productivity rates.

Table 10.1 Membrane Manufacturers

Company/Address	Phone Numbers	Web Site - http://www.	Products	Config
A/G Technology Corp. 101 Hampton Ave. Needham, MA 02194-2628	Phone: 617 449-5774 FAX: 617 449-5786	*	UF, MF	HF
Advanced Membrane Technology, Inc 10350 Barnes Canyon Rd SanDiego, CA 92121	Phone: 619 4574488 FAX: 619 457-4422	advancedmembrane.com	RO, NF, UF, MF	SW, TU
Amicon Div. W.R. Grace & Co. 72 Cherry Hill Dr. Beverly, MA 01915	Phone: 508 777-3622 FAX: 5118 921-4038	amicon.com	NF, UF, MF	Disc
Anglian Water/Fluid Systems 10054 Old Grove Rd. San Diego, CA 92131	800 525-4369 Phone: 619 695-3840 FAX: 619 695-2176	fluid-systemscorn	RO, NF, UF	S W
Applied Membrane Inc. 110 Bosstick Blvd. San Marcos, CA 92069	Phone: 619 727-3711 FAX: 619 727-4427	appliedmembranes.com	RO, NF, UF	SW, TU
Aqua Source, North America 2924 Emerywood Parkway Richmond, VA 23294	Phone: 804 756-7620	*	UF	HF
CeraMem Separations 12 Clematis Avenue Waltham, MA 02154	Phone: 617 899-0467 FAX: 617 899-1227	ceramem.com	UF	Cmpd TU
Desal/Osmonics 760 Shadowridge Dr. Vii CA 92083-7986	800 423-3725 Phone: 619 598-3334 FAX: 619 5983335	osmonics.com	RO, NF, UF, MF	S W
Dow Chemical/The Filmtec Corp. P.O. Box 1206 Midland, MI 48641-1206	800 447-4369 Phone: 517 832-1556 FAX: 517 832-1465	filmtec.com or dow.com	RO, NF	SW
Du Pont **Permasep*** Products P.O. Box 6101 Newark, DE 19714-6101	Phone: 302 45 l-968 1	dupont.com	RO, NF	HFF
FPI - Separations & Systems 8314 Tiogawoods Dr. Sacrametno, CA 95826	Phone: 619 689-2328 FAX: 619 689-1035	*	UF	SW
Graver **Separations, Inc.** 200 Lake Dr., Pencader Corp. Center Glasgow, DE 19702-3319	I Phone: 302 731-1700 FAX: 302 731-1707	graver.com	MF	Tu
Hydranautics 401 Jones Rd Oceanside, CA 92054	Phone: 619 901-2500 FAX: 619 901-2578	membranes.com	RO, NF, UF	SW
Ionics, Inc. 65 Grove Street, POB 9 13 1 Watertown, MA 02272-9131	I Phone: 617 926-2500	ionics.com	RO, ED, NF, UF, MF	Systems

Table 10.1 Membrane Manufacturers (continued)

Company/Address	Phone Numbers	Web Site - **http://www.**	Products	**Config**
Klockner Filter Products, Inc. **8314 Tiogawoods Dr** **Sacramento, CA 96828**	**Phone: 916 689-2328**	*	**UF**	**SW**
Millipore Corporation **80 Ashby Road** Bedford, MA 01730-2271	Phone: 617 275-9200 FAX: 617 275-5550	**millipore.com**	**UF**	HF
Osmonics, Inc **5951 Clearwater Dr.** **Minnetonka, MN 55343-8995**	**800 848-1750** **Phone: 612 933-2277** **FAX: 612 933-0141**	**osmonics.com**	**RO, NF, UF,** **MF**	**SW**
PCI Membrane Systems Ltd. **Laventoke** Mill Whitchurch, Hants RG28 **7NR,** England	Phone: 44 0 1256 896966 FAX: 44 0 1256 893835	*	**RO, NF,** UF	TU
Setec, Inc. **P.O. Box 2816** **Livermore, CA 94550**	**Phone: 510 294-795** **FAX: 510 449-6147**	*	**MF**	
Trisep Corporation 93 S. La **Patera** Lane Goleta, CA 93 177	Phone: 805 964-8003 FAX: 805 964-1235	**trisep.com**	RONF, UF, **MF**	SW
U.S. Filter **12442 East Putnam St.** **Whittier, CA 90602**	**Phone: 310 698-9414** **FAX: 310 698-1960**	**usfilter.com**	**UF, MF**	**Cmpd TU**
Water Equipment Technologies 832 Pike Rd W. **Palm** Beach, FL 3341 l-3855	800 6844844 Phone: 407 684-6300 FAX: 407 697-3342	•		

Table 10.2. Seawater Membranes, 203 mm (8") diameter or equivalent.

		Physical Characteristics					Performance at Test Conditions			Test Conditions				
Manufacturer	Model No.	Comp.	Configuration	Diameter, mm	Length, mm	Effective Membrane Area, m^2	Permeate Flow Rate, m^3/d	Permeate Flow Variation, %	Nominal Salt Rejection, %	Feed TDS, mg/L	Feed Pressure, MPa	Recovery, %	Temp. °C	pH
Adv. Mem. Tech.	ATFSW-B-8040-H	thin film	SW	200	1000	34.4	23	15	99.2	32,000	5.5	10	25	8
Du Pont	6835T	B-10 aramid	HFF	279	1200	999	26.5	15	99.35	35,000	6.89	35	25	*
Du Pont	6880T B-10 Twin	B-10 aramid	HFF	279	2051	1810	60.5	*	99.55	35,000	6.89	35	25	*
DuPont	6882TM B-10 Twin	B-10 aramid	HFF	279	2108	2283	68.9	*	99.35	32,000	6.89	35	25	*
Du Pont	B-10 6845TR	Aramid	HFF	216	990	700	26.5	15	99.3	35,000	6.9	35	25	*
Du Pont	SW-H-8540	B-10T asymmetric aramid	HFF	216	1032	1219	30.3	15	99.6	35,000	6.89	35	25	*
Du Pont	SW-M-8540	B-10TM asymmetric aramid	HFF	216	1032	1219	39.4	15	99.45	32,000	6.89	35	25	*
FilmTec	SW30-8040	PA/PS/PE thin film composite	SW	201	1016	27.7	23	15	99.1	35,000	5.5	10	25	8
FilmTec	SW30HR-380	PA/PS/PE thin film composite	SW	201	1016	35.3	23	-15 to +25	99.6	32,000	5.5	8	25	8
FilmTec	SW30HR-8040	PA/PS/PE thin film composite	SW	201	1016	27.1	15	-5 to +15	99.4	35,000	5.5	8	25	8
Fluid Systems	2822SS	TFC	SW	203	1016	27.9	19	15	99.6	32,890	5.52	7	25	7.5
Fluid Systems	2822SS Premium	TFC	SW	203	1016	27.9	19	15	99.8	32,800	5.52	7	25	1.5
Fluid Systems	2832SS Mag	TFC	SW	203	1524	43.2	29.1	15	99.6	32,800	5.52	1.1	25	7.5
Hydranautics	8040-HSY-SWC1	Comp. PA	SW	201.9	1016	29.26	18.9	5	99.5	32,000	5.5	10	25	6.0 - 7.0
Hydranautics	8040-HSY-SWC2	Comp. PA	SW	201.9	1016	29.26	23.5	5	99.2	32,000	5.5	10	25	6.0 - 7.6
Osmonics/Desal	Desal-11 AD8040F	Thin Film Membrane	SW	200.2	1016	31.59	18.9	15	99.4	32,000	5.516	15	25	7.5
Osmonics/Desal	Desal-3 High Pres. SC8040F	Thin Film Membrane	SW	200.2	1016	31.59	13.61	15	99.0	32,000	5.516	15	25	6.5

Table 10.2. Seawater Membranes, 203 mm (8") diameter or equivalent (continued)

		Operating Limits										Calculated Parameters		
Manufacturer	**Model No.**	Max Feed Flow, **L/min,** Max% Recovery or Min Brine Flow (Jfmin)	Max **Oper Pres., MPa**	Max Press Drop, **kPa**	Max Feed **SDI**	Max/Highly Rec. Turbidity, NTU	**Max.** Oper Temp. °C	**pH Range;** continuous operation	**pH** Range; **cleaning**	Chlorine Tolerance, **ppm or** (ppm days)	Optimum **pH** for Rejection and or Permeatio	Water Transport Coef. A* 10^{12}, $m^3m^{-2}Pa^{-1}sec^{-1}$	Salt Transport Coef. **B*10^6, m/sec**	Mod. **Eff.** $m^3day^{-1}m^{-}$
Adv. Mem. Tech.	**ATFSW-B-8040-H**	230	6.9	60	5	1	45	2 - 11	1 - 12	0	*	2.88	0.059	732
Du Pont	**6835T**	(15.9)	8.274	*	*	*	40	4 - 9	*	*	*	0.10	0.002	361
Du Pont	**6880T B-10 Twin**	(31.8)	8.274	*	*	*	40	4 - 9	*	*	*	0.12	0.001	482
Du Pont	**6882TM B-10 Twin**	(31.8)		*	*	*	30	4 - 9	*	*	*	0.10	0.002	535
Du Pont	**B-10 6845TR**	105.2	8.274	419		*	40	4 - 9	.3 - 11.9	0	*	0.14	0.002	730
Du Pont	**SW-H-8540**	132 (11)	8.3	69	5	*	30	4-9	2.5 - 12	<0.1	*	0.09	0.001	801
Du Pont	**SW-M-8540**	132 (11)	7	69	5	*	30	4 - 9	2.5 - 12	<0.1	*	0.11	0.002	1042
FilmTec	**SW30-8040**	230	6.9	*	5	1	45	2 - 11	1 - 12	< 0.1	*	3.96	0.083	713
FilmTec	**SW30HR-380**	280	7.5	*	5	1	45	2 - 11	2 - 12	< 0.1	*	2.79	0.029	713
FilmTec	**SW30HR-8040**	230	6.9	*	5	1	45	2 - 11	1 - 12	< 0.1	*	2.56	0.036	465
Fluid **Systems**	**2822SS**	17%	8.275	69	5	1 / 0.2	45	4 - 11	2.5 - 11	0	8	2.95	0.031	578
Fluid Systems	2822SS **Premium**	17%	8.275	104	5	1 / 0.2	45	4 - 11	2.5 - 11	0	8	2.95	0.015	578
Fluid Systems	**2832SS Mag**	19%	8.275	104	5	1 / 0.2	45	4 - 11	2.5 - 11	0	8	2.99	0.029	590
Hydranautics	**8040-HSY-SWC1**	283	8.27	69	4	1	45	3 - 10	3 - 10	<0.1	*	2.79	0.036	581
Hydranautics	**8040-HSY-SWC2**	283	8.27	69	4	1	45	3 - 10	3 - 10	<0.1	*	3.46	0.071	722
Osmonics/Desal	**Desal-11 AD8040F**	15%	6.895	69	<5	<1	50	4 - 11	2 - 11.5	(41.5)	6.5 - 7	2.65	0.038	591
Osmonics/Desal	**Desal-3 High Pres SC8040F**	15%	6.895	69	<5	<1	50	2 - 11	1 - 11.5	(20.8)	5.5 - 7	1.90	0.046	426

Table 10.3. Brackish Water **Membranes, 203 mm diameter or equivalent.**

Manufacturer	**Model No.**	**Physical Characteristics** Composition	Config.	Dia. (mm)	Len. (mm)	Eff. Area, (m^2)	**Performance at Test Conditions** Permeate Flow Rate, m^3/d	Permeate Flow Variation, %	Nominal Salt Rejection, %	**Test Conditions** Feed TDS, mg/L	Feed Pressure, MPa	Rec. (%)	Temp. °C	pH
Adv. Mem. Tech.	ATFRO-B-8040-H	Thin Film	SW	200	1,000	34.4	54	15	99.5	2,000	1.6	15	25	8
Adv. Mem. Tech.	ATFRO-8040	Thin Film	SW	200	1,016	34	30	15	98	2,000	1.55	10	25	8
Adv. Mem. Tech.	M-C4040A	CTA/CAB	SW	99	1,016	*	2.84	15	97.5	2,000	1.55	10	25	5.5
Adv. Mem. Tech.	M-T4040AHF	Thin Film	SW	99	1,016	*	7.57	15	96	2,000	1.55	10	25	8
Du Pont	B-9 0840R	Aramid	HFF	245	1,220	1280	60	10	92	1,500	2.76	75	25	*
Du Pont	BW-L-8540	Asymetric aramid	HFF	216	1.032	1280	41.6	10	97.3	1,500	1.55	75	25	*
FilmTec	BW30-330	PA/PS/PE	SW	201	1,016	28	28	15	99.5	2,000	1.55	15	‘25	8
FilmTec	BW30-365	PA/PS/PE	SW	201	1,016	34	36	7	99.5	2,000	1.55	15	25	8
FilmTec	BW30-400	PA/PS/PE	SW	201	1,016	37	40	7	99.5	2,000	1.55	15	25	8
FilmTec	BW30LE-400, High P	PA/PS/PE	SW	201	1,016	37	54	15	99.0					8
FilmTec	BW30LE-400, Low P	PA/PS/PE	SW	201	1,016	37	34	15	98.0	2,000	1.05	15	25	8
FilmTec	BW30LE-440	PA/PS/PE	SW	200	1,000	41	44	15	99.0	2,000	1.03	15	25	8
Fluid Systems	8221HR	CA Blend	SW	203	1,016	31.6	23.9	15	98.0	2,000	2.9	10	25	5.7
Fluid Systems	8221SD	CA Blend	SW	203	1,016	31.6	30.3	I5	95.5	2,000	2.9	10	25	5.7
Fluid Systems	8231HR Mag	CA Blend	SW	203	1,524	49.1	39.7	15	98.0	2,000	2.9	16	25	5.7
Fluid Systems	8231SD Mag	CA Blend	SW	203	1,524	49.1	49.2	I5	95.5	2,poo	2.9	16	25	5.7
Fluid Systems	8821ULP	TFC	SW	203	1,016	30.7	20.1	15	98.0	500	0.524	10	is	7.5
Fluid Systems	8821ULP	TFC	SW	203	1,016	30.7	27.3	15	98.5	500	0.69	10	25	7.5
Fluid Systems	8821ULP	TFC	SW	203	1,016	30.7	41.6	is	99.0	500	1.04	10	25	7.5
Fluid Systems	8821ULP-400	TFC	SW	203	1,016	37.2	23.8	15	98.0	500	0.524	10	25	7.5
Fluid Systems	8821ULP-400	TFC	SW	203	1,016	37.2	32.2	15	98.5	500	0.69	10	25	7.5
Fluid Systems	8821ULP-400	TFC	SW	203	1,016	37.2	49.2	15	99.0	500	1.05	10	25	7.5
Fluid Systems	8822HR	TFC	SW	203	1,016	30.7	32.2	15	99.5	2,000	1.55	10	25	7.5
Fluid Systems	8822HR Premium	TFC	SW	203	1,016	30.7	32.2	15	99.7	2,000	1.55	10	25	7.5
Fluid Systems	8822HR-400	TFC	SW	203	1,016	37.2	39.7	15	99.5	2,000	1.55	10	25	7.5
Fluid Systems	8822HR-400 Premium	TFC	SW	203	1,016	37.2	39.7	15	99.7	2,000	1.55	10	25	7.5
Fluid Systems	8822XR	TFC	SW	203	1,016	30.7	25.7	15	99.6	2,000	2.24	10	25	7.5
Fluid Systems	8822XR-400	TFC	SW	203	1,016	37.2	31.4	15	99.6	2,000	2.24	10	25	7.5
Fluid Systems	8832HR Mag	TFC	SW	203	1,524	47.4	49.2	15	99.5	2,000	1.55	16	25	7.5
Hydranautics	8040-LHY-CPA2	Comp. PA	SW	201.9	1.016	37.2	41.6	15	99.0	1,500	1.55	15	25	6.5 - 7.

Table 10.3. Brackish Water Membranes, 203 mm diameter or equivalent (continued)

		Operating Limit3										Calculated Parameters		
Manufacturer	Model No.	Max Feed Flow, L/min, Max % Recovery or Min Brine Flow (L/min)	Max Oper Pres., MPa	Max Press Drop, kPa	Max Feed SDI	Max/Highly Rec. Turbidity, NTU	Max. Oper Temp. °C	pH Range; continuous operation	pH Range; cleaning	Chlorine Tolerance, ppm or (ppm days)	Optimum pH for Rejection and or Permeatior	Water Transport Coef. A* 10^{12}, $m^3m^{-2}Pa^{-1}sec^{-1}$	Salt Transport Coef. B* 10^6, m/sec	Efficienc $m^3day^{-1}m$
Adv. Mem. Tech.	ATFRO-B-8040-H	16	4.8	103	5	1	50	4 - 9	1 - 12	0	*	12.8	0.084	1719
Adv. Mem. Tech.	ATFRO-8040	303	4.2	*	5	1	46	2 - 11	1 - 12	<0.1	*	7.4	0.197	940
Adv. Mem. Tech.	M-C4040A	38	2.1	*	4	1	40	4-6		0.3 - 0.5	*	*	*	363
Adv. Mem. Tech.	M-T4040AHF	38	2.1	*	4	I	45	2 - 11		0	*	*	*	968
Du Pont	B-9 0840R	(65.9)	2.76	137	<3	*	40	4 - 11	2.3 - 11.9#	< 0.1	*	0.220	0.019	1043
Du Pont	BW-L-8540	132 (11)	2.76	69	5	*	40	4 - 11	2.5 - 12	< 0.1	*	0.302	0.004	1100
FilmTec	BW30-330	265	4.1	*	5	1	45	2 - 11	1 - 12	< 0.1	*	8.5	0.053	869
FilmTec	BW30-365	265	4.1	*	5	1	45	2 - 11	1 - 12	< 0.1	*	9.0	0.057	1117
FilmTec	BW30-400	265	4.1	*	5	1	45	2 - 11	1 - 12	< 0.1	*	9.1	0.058	1241
FilmTec	BW30LE-400, High P	265	4.1	*	5	1	45	2 - 11	1 - 12	< 0.1	*	12.3	0.157	1675
FilmTec	BW30LE-400, Low P	265	4.1	*	5	1	45	2 - 11	1 - 12	< 0.1	*	11.9	0.199	1055
FilmTec	BW30LE-440	322	4.1	*	5	1	45	2 - 11	1 - 12	< 0.1	*	14.6	0.115	1401
Fluid Systems	8221HR	16%	4.14	69	5	1 / 0.2	40	4 - 6	2.5 - 7	1	5.7	3 . 2	0.169	727
Fluid Systems	8221SD	16%	4.14	69	5	110.2	40	4-6	2.5 - 7	1	5.7	4.1	0.495	921
Fluid Systems	8231HR Mag	20%	4.14	1 04	5	1 / 0.2	40	4 - 6	2.5 - 7	1	5.7	3.4	0.174	805
Fluid Systems	8231SD Mag	20%	4.14	104	5	1 / 0.2	40	4-6	2.5 - 8	1	5.7	4.3	0.499	997
Fluid Systems	8821ULP	17%	2.44	69	5	1 IO.2	45	4 - 11	2.5 - 1.1	0	8	15.8	0.147	611
Fluid Systems	88211JLP	17%	2.44	69	5	110.2	45	4-11	2.5 - 11	0	8	15.9	0.148	830
Fluid Systems	8821ULP	17%	2.44	69	5	1 / 0.2	45	4 - 11	2.5 - 1 I	0	8	15.7	0.150	1265
Fluid Systems	8821ULP-400	17%	2.44	69	5	110.2	45	4-11	2.5-11	0	8	15.4	0.143	724
Fluid Systems	8821ULP-400	17%	2.44	69	5	1 / 0.2	45	4 - 1 1	2.5 - 11	0	8	15.5	0.145	979
Fluid Systems	8821ULP-400	17%	2.44	69	5	110.2	45	4 - 11	2.5 - 11	0	8	15.2	0.146	1496
Fluid Systems	88 22HR	17%	4.14	69	5	1 / 0.2	445	4 - 11	2.5 - 11	0	8	8.8	0.058	979
Fluid Systems	8822HR Premium	17%	4.13	69	5	110.2	45	4 - 11	2.5 - 11	0	8	8.8	0.035	979
Fluid Systems	8822HR-400	17%	4.14	69	5	1 / 0.2	45	4 - 11	2.5 - 11	0	a	9.0	0.059	1207
Fluid Systems	8822HR-400 Premium	17%	4.13	69	5	110.2	45	4 - 11	2.5 - 11	0	8	9.0	0.035	1207
Fluid Systems	8822XR	17%	4.13	69	5	1 / 0.2	45	4 - 11	2.5 - 11	0	8	4.7	0.037	782
Fluid Systems	8822XR-400	17%	4.13	69	5	1 / 0.2	45	4 - 11	2.5 - 11	0	8	4.7	0.037	955
Fluid Systems	8832HR Mag	20%	4.14	104	5	1 / 0.2	45	4-11	2.5 - 11	0	8	8.8	0.055	997
Hydranautics	8040-LHY-CPA2	283	4.2	69	4	1	45	3 - 10	3 - 10	<0.1	*	9 2	0 120	1279

Table 10.3. Brackish Water Membranes, 203 mm diameter or equivalent.

		Physical Characteristics					Performance at Test Conditions			Test Conditions				
Manufacturer	Model No.	Composition	Config.	Dia. (mm)	Len. (mm)	Eff. Area, (m^2)	Permeate Flow Rate, m^3/d	Permeate Flow Variation, %	Nominal Salt Rejection, %	Feed TDS, mg/L	Feed Pressure, MPa	Rec. (%)	Temp. °C	pH
Hydranautics	8040-LHY-CPA3	Comp. PA	SW	201.9	1,016	37.2	41.6	15	99.5	1,500	1.55	15	25	6.5 - 7.0
Hydranautics	8040-LHY-CPA4	Comp. PA	SW	201.9	1,016	37.2	22.7	15	99.5	1,500	1.55	15	25	6.5 - 7.0
Hydranautics	8040-LSY-CPA2	Comp. PA	SW	201.9	1,016	33.9	34	15	99.0	1,500	i.55	15	25	6.5 - 7.0.
Hydranautics	8040-MSY-CAB1	CA Blend	SW	201.9	1,016	31.6	32.2	15	95.0	2,000	2.89			5.0 - 6.0
Hydranautics	8040-MSY-CAB2	CA Blend	SW	201.9	1,016	31.6	25.7”	15	98.0	2,000	2.89	10	25	5.0 - 6.0
Hydranautics	8040-MSY-CAB3	CA Blend	SW	201.9	1,016	31.6	17.0	15	99.0	2,000	2.89	10	25	5.0 - 6.0
Hydranautics	8040 ESPA1	Comp. PA	SW	201.9	1,016	37.2	45.4	15	99.0	1,500	1.05	15	25	6.5 - 7.0
Hydranautics	8040 ESPA2	•	•	201.9	1.016	*	34.1	15	PP.5	1,500	1.05	15	25	6.5 - 7.0
Hydranautics	8040 ESPA3	*	SW	201.9	1,016	*	56.8	15	98.0	1,500	1.05	is	25	6.5 - 7.0
Hydranautics	8040 LFC1	•	•	201.9	1,016	•	37.8	15	99.0	1,500	1.55	15	25	6.5 - 7.0
Hydranautics	8040 LFC2	*	SW	201.9	1,016	*	41.6	15	95.0	1,500	1.55	15	25	6.5 - 7.0
Osmonics	811-HR	CDTA	SW	211	1,016	34.7	27.4	15	97.5	2,000	2.9	15	25	6 - j
Osmonics	811-PR	CDTA	SW	211	1,016	34.7	33.4	15	87.0	2,000	2.21	15	25	6 - 7
Osmonics	811-SR	CDTA	SW	211	1,016	34.7	34.6	15	95.0	2,000	2.9	15	25	6-7
Osmonics	813-HR	CDTA	SW	211	1,016	32.5	30.5		97.5	2,000	2.9	10	25	6 - 7
Osmonics	813-PR	CDTA	SW	211	1,016	32.5	37.1	15	87.0	2,000	2.21	10	25	6-7
Osmonics	813-SR	CDTA	SW	211	1,016	32.5	38.7	15	95.0	2,000	2.9	10	25	6 - 7
Osmonics	815-HR	CDTA	SW	201	1,016	29.4	24.3	15	97.5	2,000	2.9	15	25	6 - 7
Osmonics	815-PR	CDTA	SW	201	1,016	29.4	29.6	15	87.0	1,000	2.21	15	25	6 - 7
Osmonics	815-SR	CDTA	SW	201	1,016	29.4	30.7	15	95.0					6 - 7
Osmonics	817-HR	CDTA	SW	201	1,016	31	27.2	15	97.5	2,000	2.9	10	25	6 - 7
Osmonics	817-PR	CDTA	SW	201	1,016	31	33.2	15	87.0					6 - 7
Osmonics	817-SR	CDTA	SW	201	1,016	31	34.5	15	95.0	2,000	2.21	10	25	6 - 7
Osmonics	811-HF (PA)	Polyamide	SW	211	1,016	38.6	38.2	15	99.0	2,000	2.21	15	25	7 - 8
Osmonics	811-HR (PA)	Polyamide	SW	211	1,016	34.8	31.8	15	99.0	2,000	2.21	15	25	7 - 8
Osmonics	813-HR (PA)	Polyamide	SW	211	1,016	37.62	34.7	15	99.0	2,000	2.21	10	25	7-8
Osmonics	815-HF (PA)	Polyamide	SW	201	1,016	34.4	‘j4.1	15	99.0	2,000	2.21	15	25	7 - 8
Osmonics	815-HR (PA)	Polyamide	SW	201	1,016	31.6	28.4	15	99.0	2,000	2.21	15	25	7-8
Osmonics	817-HR (PA)	Polyamide	SW	201	1,016	33.4	30.9	15	99.q:	2,000	2.21	10	25	7 - 8
Osmonics	823-HR (PA)	Polyamide	SW	201	1,016	37.16	39.7	15	99.0	2.000	1.55	*	25	*
Osmonics/Desal	Desal-11 AG8040F	Thin Film Membrane	SW	200.2	1,016	32.52	39.69	15	99.0	2,000	1.551	15	25	7.5
Osmonics/Desal	Desal-11 Durasan ™ AG8040C	Thin Film Membrane	SW	201	1,016	32.52	39.69	15	99.4	2,000	1.551	15	25	7.5
Osmonics/Desal	Desal-3 Low Pres. SG8040F	Thin Film Membrane	SW	200.2	1,016	32.52	29.11	15	99.0	2.000	1.951	15	25	6.5

Table 10.3. Brackish Water Membranes, 203 mm diameter or equivalent (continued)

Manufacturer	Model No.	Operating Limits: Max Feed Flow, L/min, Max% Recovery or Min Brine Flow (L/min)	Max Oper Pres., MPa	Max Press Drop, kPa	Max Feed SDI	Max/Highly Rec. Turbidity, NTU	Max. Oper Temp. °C	pH Range; continuous operation	pH Range; cleaning	Chlorine Tolerance, ppm or (ppm days)	Optimum pH for Rejection and or Permeatior	Calculated Parameters: Water Transport Coef. A* 10^{12}, $m^3m^{-2}Pa^{-1}sec^{-1}$	Salt Transport Coef. B* 10^6, m/sec	Efficiency $m^3day^{-1}m^{-}$
Hydranautics	8040-LHY-CPA3	283	4.2	69	4	1	45	3 - 10	3 - 10	<0.1	*	9.2	0.060	1279
Hydranautics	8040-LHY-CPA4	283	4.2	69	4	1	45	3 - 10	3 - 10	<0.1	*	5.0	0.033	698
Hydranautics	8040-LSY-CPA2	283	4.2	69	4	1	45	3 - 10	3 - 10	<0.1	*	8.2	0.108	1045
Hydranautics	8040-MSY-CAB1	283	4.14	103	4	1	40	4 - 6	4 - 6	1	*	4.3	0.588	990
Hydranautics	8040-MSY-CAB2	283	4.14	103	4	1	40	4 - 6	4 - 6	1	*	3.5	0.182	790
Hydranautics	8040-MSY-CAB3	283	4.14	103	4	1	40	4 - 6	4 - 6	1	*	2.3	0.060	523
Hydranautics	8040 ESPA1	283	4.16	69	4	1	45	3 - 10	3 - 10	<0.1	*	15.4	0.131	1396
Hydranautics	8040 ESPA2	283	4.16	69	4	1	45	3 - 10	3 - 10	<0.1	*	*	*	1048
Hydranautics	8040 ESPA3	283	4.16	69	4	1	45	3 - 10	3 - 10	<0.1	*	*	*	1746
Hydranautics	8040 LFC1	283	4.14	69	4	1	45	3 - 10	*	<0.1	*	*	*	*
Hydranautics	8040 LFC2	283	4.14	69	4	1	45	3 - 10	*	<0.1	*	*	*	*
Osmonics	811-HR	15%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	3.4	0.215	773
Osmonics	811-PR	15%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	5.4	1.530	942
Osmonics	811-SR	15%	6.9	*	<5	*	50	3 - 7	*	0.2 - 2	*	4.2	0.558	976
Osmonics	813-HR	10%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	4.0	0.264	859
Osmonics	813-PR	10%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	6.4	1.870	1044
Osmonics	813-SR	10%	6.9	*	<5	*	50	3 - 7	*	0.2 - 2	*	5.0	0.687	1089
Osmonics	815-HR	15%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	3.5	0.225	756
Osmonics	815-PR	15%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	5.7	1.600	918
Osmonics	815-SR	15%	6.9	*	<5	*	50	3 - 7	*	0.2 - 2	*	4.4	0.585	952
Osmonics	817-HR	10%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	3.7	0.247	844
Osmonics	817-PR	10%	*	*	<5	*	*	3 - 7	*	0.2 - 2	*	6.0	1.755	1030
Osmonics	817-SR	10%	6.9	*	<5	*	50	3 - 7	*	0.2 - 2	*	4.7	0.642	1070
Osmonics	811-HF (PA)	15%	*	*	<5	*	*	3 - 11	*	< 0.1	*	5.6	0.106	1075
Osmonics	811-HR (PA)	15%	*	*	<5	*	*	3 - 11	*	< 0.1	*	5.2	0.098	895
Osmonics	813-HR (PA)	10%	*	*	<5	*	*	3 - 11	*	< 0.1	*	5.2	0.102	977
Osmonics	815-HF (PA)	15%	*	*	<5	*	*	3 - 11	*	< 0.1	*	5.7	0.106	1058
Osmonics	815-HR (PA)	15%	*	*	<5	*	*	3 - 11	*	< 0.1	*	5.1	0.097	881
Osmonics	817-HR (PA)	10%	*	*	<5	*	*	3 - 11	*	< 0.1	*	5.3	0.102	958
Osmonics	823-HR (PA)	*	*	*	*	*	*	*	*	*	*	*	*	*
Osmonics/Desal	Desal-11 AG8040F	15%	2.758	69	<3	<1	50	4 - 11	2-11.5	(41.5)	6.5 - 7	10.3	0.131	1241
Osmonics/Desal	Desal-11 Durasan ™ AG8040C	15%	1.379	69	<3	<1	50	4 - 11	2-11.5	(41.5)	6.5 - 7	10.3	0.078	1231
Osmonics/Desal	Desal-3 Low Pres. SG8040F	15%	1.379	69	<5	<1	50	2 - 11	1 - 11.5	(20.8)	5.5 - 7	7.6	0.096	910

Table 10.3. Brackish Water Membranes, 203 mm diameter or equivalent.

		Physical Characteristics					Performance at Test Conditions			Test Conditions				
Manufacturer	Model No.	Composition	Config.	Dia. (mm)	Len. (mm)	Eff. Area, (m^2)	Permeate Flow Rate, m^3/d	Permeate Flow Variation, %	Nominal Salt Rejection, %	Feed TDS, mg/L	Feed Pressure, MPa	Rec. (%)	Temp. °C	pH
Osmonics/Desal	Desal-3 Stand. Pres. SE8040F	Thin Film Membrane	SW	200.2	1.016	32.52	29.1 1	15	98.9	2,000	2.93	15	25	6.5
Osmonics/Desal	Desal-3 Durasan ™ SG8040C	Thin Film Membrane	SW	201.9	1,016	32.52	27.22	15	98.2	2,000	1.551	15	25	6.5
Osmonics/Desal	Desal-3 Duratherm™RO SC8040CZH	Thin Film Membrane	SW	201.9	1,016	32.52	19.44	15	99.0	2,000	2.93	15	25	6.5
Osmonics/Desal	Desal-3 Duratherm™RO SE8040CZH	Thin Film Membrane	SW	201.9	1,016	32.52	24.6	15	98.9	2,000	2.93	15	25	6.5
Osmonics/Desal	Desal-3 Duratherm™RO SG8040CZH	Thin Film Membrane	SW	201.9	1,016	32.52	24.6	15	98.2	2,000	1.551	15	25	6.5
Osmonics/Desal	Desal-CA CD8040F	CTDB	SW	200.2	1,016	32.52	22.68	15	98.0	2,000	2.93	15	25	*
Osmonics/Desal	Desal-CA CE8040F	CTDB	SW	200.2	1,016	32.52	30.24	15	97.0	2,000	2.93	15	25	*
Osmonics/Desal	Desal-CA CG8040F	CTDB	SW	200.2	1,016	32.52	27.59	15	87.0	2,000	1.551	15	25	*
TriSep Corp.	8040-ACM1-TSA	PA(ACM)	SW	202	1,016	33.5	28	15	99.5	2,000	1.55	15	25	8
TriSep Corp.	8040-ACM1-UWA	PA (ACM)	SW	202	1,016	33.5	34	15	99.5	2,000	1.55	15	25	8
TriSep Corp.	8040-ACM2-TSA	PA (ACM)	SW	202	1,016	33.5	34	15	99.5	2,000	1.55	15	25	8
TriSep Corp.	8040-ACM2-UWA	PA (ACM)	SW	202	1,016	33.5	39.7	15	99.5	2,000	1.55	15	25	8
TriSep Corp.	8040-ACM3-TSA	PA (ACM)	SW	202	1,016	33.5	39.8	15	99.0	2,000	1.55	15	25	8
TriSep Corp.	8040-ACM4-TSA	PA (ACM)	SW	202	1,016	33.5	53	15	99.2	2,000	1.55	15	25	8
TriSep Corp.	8040-ACM4-TSA	PA (ACM)	SW	202	1,016	33.5	21	10	98.0	2,000	0.69	15	25	8
TriSep Corp.	8040-SB10-TSA	CA Blend	SW	202	1,016	31.6	17	15	99.0	2,000	2.9	15	25	5-6
TriSep Corp.	8040-SB20-TSA	CA Blend	SW	202	1,016	31.6	26	15	98.0	2,000	2.9	15	25	5-6
TriSep Corp.	8040-SB50-TSA	CA Blend	SW	202	1,016	31.6	32	15	95.0	2,000	2.9	15	25	5-6
TriSep Corp.	8040-SB90-TSA	CA blend	SW	202	1,016	31.6	25	15	85.0	2,000	2.9	15	25	5-6
TriSep Corp.	8040-X201-TSA	PA (ACM)	SW	202	1,016	33.5	30.3	15	99.3	2,000	1.55	15	25	8
TriSep Corp.	8060-SB20-TSA	CA Blend	SW	202	1,524	47.4	51	15	98.0	2,000	2.9	15	25	5-6
TriSep Corp.	8060-SB50-TSA	CA Blend	SW	202	1,524	43.4	51	15	95.0	2,000	2.9	15	25	5-6
TriSep Corp.	8340-ACM1-TSA	PA (ACM)	SW	211	1,016	34.6	30	15	99.0	2,000	1.55	15	25	8
TriSep Corp.	8340-SB20-TSA	SA Blend	SW	211	1,016	32.6	26	15	98.0	2,000	2.9	15	25	5-6
TriSep Corp.	8340-SB50-TSA	CA blend	SW	211	1,016	32.6	32	15	95.0	2,000	2.9	15	25	5-6
TriSep Corp.	8540-SB20-TSA	CA blend	SW	216	1,016	35.3	30	15	98.0	2,000	2.9	15	25	5-6
TriSep Corp.	8540-SB50-TSA	CA blend	SW	216	1.016	35.3	38	15	95.0	2,000	2.9	15	25	5-6

Table 10.3. Brackish Water Membranes, 203 mm diameter or equivalent (continued)

		Operating Limits										Calculated Parameters		
Manufacturer	Model No.	Max Feed Flow, **L/min, Max %** Recovery or Min Brine Flow **(L/min)**	Max Oper **Pres., MPa**	Max Press **Drop, kPa**	Max Feed **SDI**	Max/Highly Rec. Turbidity, NTU	**Max. Oper** Temp. °C	**pH** Range; continuous operation	**pH** Range; cleaning	Chlorine Tolerance, ppm or (ppm **days)**	Optimum **pH** for Rejection and or Permeation	Water Transport **Coef.** A* **10^{12}, m^{3}m^{-2}Pa^{-1}sec^{-1}**	Salt Transport Coef. **B*10^6, m/sec**	**Efficienc m^{3}day^{-1}m**
Osmonics/Desal	**Desal-3 Stand. Pres. SE8040F**	15%	3.448	69	<5	<1	50	2-11	1 - 11.5	(20.8)	5.5 - 7	3.8	0.106	910
Osmonics/Desal	**Desal-3 Durasan^{TM} SG8040C**	15%	1.379	69	<5	<1	50	2 - 11	1 - 11.5	(20.8)	5.5 - 7	7 . 1	0.163	837
Osmonics/Desal	**Desal-3 Duratherm^{TM}RO SC8040CZH**	15%	4.131	69	<5	<1	70	4-9	- 11.5, T<	120.8, T<50)	6.5 - 7	2.5	0.064	598
Osmonics/Desal	**Desal-3 Duratherm^{TM}RO SE8040CZH**	15%	4.137	69	<5	<1	70	4 - 9	- 11.5, T<	(20.8, T<50)	6.5 - 7	3.2	0.089	756
Osmonics/Desal	**Desal-3 Duratherm^{TM}RO SG8040CZH**	15%	4.137	69	<5	<1	70	4 - 9	- 11.5, T<	(20.8, T<50)	6.5 - 7	6.4	0.147	756
Osmonics/Desal	**Desal-CA CD8040F**	15%	3.103	69	<5	<1	30	5 - 6.5	3 - 8	1	*	2.9	0.151	709
Osmonics/Desal	**Desal-CA CE8040F**	15%	3.103	69	<5	<1	30	5 - 6.5	3-8	1	*	3.9	0.306	946
Osmonics/Desal	**Desal-CA CG8040F**	15%	3.103	69	<5	<1	30	S-4.5	3-8	1	*	7.1	1.348	863
TriSep Corp.	**8040-ACM1-TSA**	300	4.1	69	5	1	45	4 - 11	2.5 - 11.5	<0.1	*	7.1	0.045	860
TriSep Corp.	**8040-ACM1-UWA**	300	4.2		5	1	45	4 - 11	2.5 - 11.5.	<0.1	*	8.6	0.054	1044
TriSep Corp.	**8040-ACM2-TSA**	300	4.1	69	5	1	45	4 - 11	2.5 - 11.5	<0.1	*	8.6	0.054	1044
TriSep Corp.	**8040-ACM2-UWA**	300	4.1	69	5	1	45	4 - 11	2.5 - 11.5	<0.1	*	10.0	0.063	1219
TriSep Corp.	**8040-ACM3-TSA**	300	4.1	69	5	1	45	4 - 11	2.5 - II.5	<0.1	*	10.0	0.128	ii22
TriSep Corp.	**8040-ACM4-TSA**	300	4.1	69	5	1	45	4-11	2.5 - 11.5	<0.1	*	13.4	0.136	1628
TriSep Corp.	**8040-ACM4-TSA**	300	4.1	69	5	1	45	4 - 11	2.5 - 11.5	<0.1	*	14.1	0.135	639
TriSep Corp.	**8040-SB10-TSA**	300	4.1	69	5	1	40	4 - 7	2.5 - 7.5	1	*	2.3'	0.058	522
TriSep Corp.	**8040-SB20-TSA**	300	4.1	69	5	1	40	4-7	2.5 -1.5	1	*	3.5	0.179	799
TriSep Corp.	**8040-SB50-TSA**	300	4.1	69	5	1	40	4-7	2.5 -7.5	1	*	4.3	0.567	983
TriSep Corp.	**8040-SB90-TSA**	300	4.1	69	5	1	40	4-l	2.5 - 1.5	1	*	3.3	1.485	768
TriSep Corp.	**8040-X201-TSA**	300	4.1	69	5	1	45	4 - 11	2.5 - 11.5	<0.1	*	7.6	0.068	931
TriSep Corp.	**8060-SB20-TSA**	300	4.1		5	1	40	4-7	2.5 - i.5	1	*	4.6	0.234	1044
TriSep Corp.	**8060-SB50-TSA**	300	4.1	69	5	1	40	4 - 7	2.5 - 7.5,	1	*	4.6	0.602	1044
TriSep Corp.	**8340-ACM1-TSA**	300	4.1		5	1	45	4 - 11	2.5 - II.5	<0.1	*	7.3	0.093	844
TriSep Corp.	**8340-SB20-TSA**	300	4.1		5	1	40		2.5 - 7.5	1	*	3.4	0.173	132
TriSep Corp.	**8340-SB50-TSA**	300	4.1	69	5	1	40	4-l	2.5 - 1.5	1	*	4.2	0.549	901
TriSep Corp.	**8540-SB20-TSA**	300	4.1	69	5	1	40	4 - 7	2.5 -7.5	1	*	3.6	0.184	806
TriSep Corp.	**8540-SB50-TSA**	300	4.1	69	5	1	40	4-7	2.5 - 7.5	1	*	4.6	0.603	1021

Table 10.4. **Nanofiltration** Membranes, 203 mm (8”) diameter or equivalent.

Physical Characteristics							Performance at Test Conditions			Test Conditions						
Manufacturer	**Model No.**	**Comp.**	**Config.**	Dia. (mm)	**Len.** (mm)	**Eff. Area, m²**	**Permeate** Flow **Rate, m³/d**	Typical Chloride **Rej.**, % **(R_c)**	Typical **Divalent Rej.**, **% (R_d)**	**NaCl Conc.** for **R_c, mg/L**	Divaient **Conc.** for **R_d, mg/L**	Test solution	Feed **Pressure, MPa (divalent)**	Recovery, %	**Temp. °C**	**pH**
Adv. Mem. Tech.	**ATF50-B-8040**	**Thin Film**	**SW**	**200**	**1,000**	**34.4**	**40**	**60**	**95**	**2,000**	2,000	**$MgSO_4$**	**0.5**	13	25	7
Adv. Mem. Tech.	**ATF70-B-8040**	**Thin Film**	**SW**	**200**	**1,000**	**34.4**	**35**	**70**	**95**	2,000	2,000	**$MgSO_4$**	0.5	**15**	25	**7**
Adv. Mem. Tech.	**ATF80-B-8040**	**Thin Film**	**SW**	**200**	**1,000**	**34.4**	**35**	**80**	**95**	**2,000**	**2,000**	**$MgSO_4$**	**0.5**	**15**	**25**	**7**
FilmTec	**NF45-8040**	**PA/PS/PE**	**SW**	**200**	**1,000**	*	**28.3**	**45**	**98**	2,000	2,000	**$MgSO_4$**	0.9	*	25	*
FilmTec	**NF70-345**	**PA/PS/PE**	**SW**	**200**	**1,000**	**32**	**40**	60	95	2,000	2,000	**$MgSO_4$**	0.5	**15**	25	8
FilmTec	**NF70-400**	**PA/PS/PE**	**SW**	**200**	**1,000**	**37**	47	60	95	2,000	2,000	**$MgSO_4$**	0.5	is	25	8
FilmTec	**NF90-400**	**PA/PS/PE**	**SW**	**200**	**1,000**	**37**	**39**	**90**	**95**	2,000	2,000	**$MgSO_4$**	O S	**15**	**25**	*
Fluid Systems	**8231LP Mag**	**CA Blend**	**SW**	**203**	**1,524**	**49.1**	49.2	75	**95**	2,000	*	typical BW	1.55	16	25	**5.7**
Fluid Systems	**8921S**	**TFC**	**SW**	**203**	**1,016**	**30.7**	2 9	85	9 s	**500**	*	**typical BW**	0.552	**10**	25	**7.5**
Fluid Systems	**8921S-400**	**TFC**	**SW**	**203**	**1,016**	**37.2**	35	85	95	500	*	typical BW	0.552	**10**	25	7.5
Hydranautics	**8040 PVDI**	**PVD**	**SW**	**201.9**	**1,016**	**33.9**	**41.6**	**80**	**95**	**500**	250	**San Diego** TW	**1** .a5	**15**	**25**	**7.5 - 8.5**
Hydranautics	**8040 ESNA1**	*	*SW*	*201.9*	*1,016*	*	**41.6**	**70**	*	500	*	*	0.52	**15**	25	6.5 - 7.0
Hydranautics	**8040 ESNA2**	*	**SW**	**202**	**1,016**	*	**56.8**	**50**	*	**500**	*	*	**0.52**	**15**	**25**	**6.5 - 7.0**
Osmonics	**811-NF100 (CA)**	**CDTB**	**SW**	**211**	**1,020**	**32.5**	31.3	60	94	*	2.000	**Na_2SO_4**	1.59	**10**	25	6-7
Osmonics	**811-NF300 (PA)**	**PA**	**SW**	**211**	**1,020**	**32.5**	**33.8**	**40**	**95**	*	2,000	**Na_2SO_4**	**0.93**	**10**	**25**	*
Osmonics	**813-NF300 (PA)**	**PA**	**SW**	**211**	**1,020**	**34.7**	36.4	40	95	*	2,000	**Na_2SO_4**	0.93	**10**	25	*
Osmonics	**815-NF300 (PA)**	**PA**	**SW**	**211**	**1,020**	**29.4**	30.3	40	**95**	*	2.000	**Na_2SO_4**	0.93	I 0	**25**	*
Osmonics	**817-NF300 (PA)**	**PA**	**SW**	**211**	**1,020**	**31.2**	32.2	40	95	*	2,000	**Na_2SO_4**	0.93	10	25	*
Osmonics/Desal	**Desal-5 DK8040C**	**Thin Film**	**SW**	**201.9**	**1,016**	**32.52**	36.24	*	98	*	**1,000**	**$MgSO_4$**	0.69	10	2 5	*
Osmonics/Desal	**Desal-5 DK8040F**	**Thin Film**	**SW**	**200.2**	**1,016**	**32.52**	30.24	*	98	*	1,000	**$MgSO_4$**	0.69	I0	25	*
Osmonics/Desal	**Desal-5 DL8040C**	**Thin Film**	**SW**	**201.9**	**1,016**	**32.52**	38.56	*	96		**1,000**	**$MgSO_4$**	0.69	IO	**25**	*
Osmonics/Desal	**Desal-5 DL8040F**	**Thin Film**	**SW**	**200.2**	**1,016**	**32.52**	38.56	*	96	*	[illegible]	**$MgSO_4$**	0.69	10	25	*
TriSep Corp.	**8040-TS20-TSA**	**Thin Film**	**SW**	**202**	**1,016**	**33.4**	**43.6**	**20**	*	**500**	*	**NaCl**	0.72	**15**	25	8
TriSep Corp.	**8040-TS40-TSA**	**Thin Film**	**SW**	**202**	**1,016**	**33.4**	41.7	40	98	500	500	**$MgSO_4$**	0.69	**15**	25	8
TriSep Corp.	**8040-TS60-TSA**	**Thin Film**	**SW**	**202**	**1,016**	**33.4**	**30.3**	**70**	*	**500**	*	**NaCl**	0.72	**15**	**25**	**8**
TriSep Corp.	**8040-TS80-TSA**	**Thin Film**	**SW**	**202**	**1,016**	**33.4**	30	80	98	500	**500**	**$MgSO_4$**	0.69	15	**25**	8
TriSep Corp.	**8040-TS90-TSA**	**Thin Film**	**SW**	**202**	**1,016**	**33.4**	**33.4**	**92**	*	2,000	*	**NaCl**	0.72	**15**	**25**	**8**
TriSep Corp.	**813-NF100 (CA)**	**CDA/CTA**	**SW**	**211**	**1,020**	**34.7**	34.5	60	94	*	2,000	**Na_2SO_4**	1.59	IO	25	6-7
TriSep Corp.	**815-NF100 (CA)**	**CDA/CTA**	**SW**	**211**	**1,020**	**29.4**	27.8	60	94	*	**2,000**	**Na_2SO_4**	1.59	**10**	**25**	**6-7**
TriSep Corp.	**817-NF100 (CA)**	**CDA/CTA**	**SW**	**211**	**1,020**	**31.2**	30.9	60	94	*	2,000	**Na_2SO_4**	1.59	10	25	6-7

Table 10.4. **Nanofiltration** Membranes, 203 mm (8") diameter or equivalent (continued)

		Operating Limits										Calculated Parameters		
Manufacturer	Model No.	Max Feed Flow, L/min, Max % Recovery or Min Brine Flow (L/min)	Max Oper. Pres., MPa	Max. Pressure Drop, kPa	Max Feed SDI	Max/Highly Recommended Turbidity, NTU	Max. Operating Temp. °C	pH Range; continuous operation	pH Range; cleaning	Chlorine Tolerance, ppm or (ppm days)	Opt. pH for Rej. and/or Permeation	Water Transport Coef. A*10^{12}, $m^3m^{-2}Pa^{-1}sec^{-1}$	Salt Transport Coef. B*10^6, m/sec	Mod. Eff. $m^3day^{1}m^{-3}$
Adv. Mem. Tech.	**ATF50-B-8040**	265	4.1	103	5	1	50	4 - 9	3 - 10	0	•	30.1	8.245	1273
Adv. Mem. Tech.	**ATF70-B-8040**	265	4.1	103	5	1	50	4-9	3 - 10	0	•	26.9	4638	1114
Adv. Mem. Tech.	**ATF80-B-8040**	265	4.1	103	5	1	50	4 - 9	3 - 10	0	•	27.5	2.165	1114
FilmTec	**NF45-8040**	245	4.1	•	5	•	45	3 - 10	2 - 11	<0.1	•	•	•	901
FilmTec	**NF70-345**	265	1.7	•	5	1	35	3 - 9	1 - 11	< 0.1	•	34.8	0.700	1273
FilmTec	**NF70-400**	265	I.7	•	5	1	35	3-9	1 - 11	< 0.1	•	35.4	0.71 I	1496
FilmTec	**NF90-400**	265	1.7	•	5	1	35	3 - 9	1 - 11	< 0.1	•	29.4	0.590	1241
Fluid Systems	8231LP **Mag**	20%	1.655	104	5	I / 0.2	40	3 - 7	2.5 - 8	1	5.1	a.2	3.530	997
Fluid Systems	**8921S**	17%	2.41	69	5	1 / 0.2	45	4 - 11	2.5 - 11			21.3	1.828	882
Fluid Systems	89215400	17%	2.41	69	5	1 / 0.2	45	4 - 11	2.5 - 11	0	8	21.2	1.821	1064
Hydranautics	**8040 PVDI**	280	4.14	69	4	1	40	2 - 8	2 - 8	1	•	13.95	1.864	1279
Hydranautics	8040 **ESNA1**	280	4.14	69	•	i	45	3-10	•	>0.1	•	•	•	1279
Hydranautics	**8040 ESNA2**	280	4.14	69	•	1	45	3 - 10	•	>0.1		•	•	1746
Osmonics	**811-NF100** (CA)	•	6.9	•	•	•	80	2-12	•	<1	•	7.5	0.674	818
Osmonics	**811-NF300 (PA)**	•	6.9	•	•	•	50	2 - 8	•	2	•	14.6	0.600	948
Osmonics	**813-NF300** (PA)	•	3.44	•	•	•	50	2-8	•	2	•	14.7	0.605	1021
Osmonics	**815-NF300 (PA)**	•	2.76	•	•	•	S 0	2-8	•	2	•	14.4	0.595	850
Osmonics	**817-NF300** (PA)	•	2.07	•	•	•	50	2-8	•	2	•	14.5	0.596	903
Osmonics/Desal	**Desal-5 DK8040C**	1 5%	3.448	69	•	•	50	1 - 11	1-II.5	(41.5)	•	16.6	0.208	930
Osmonics/Desal	**Desal-5 DK8040F**	15%	3.448	69	•	•	50	2 - 11	1 - 11.5	(41.5)	•	16.6	0.21	946
Osmonics/Desal	**Desal-5 DL8040C**	15%	3.448	69	•	•	50	1 - 11	1 - 11.5	(41.5)	•	21.2	0.542	1185
Osmonics/Desal	**Desal-5 DL8040F**	15%	3.448	69	•	•	50	2-11	1 - 11.5	(41.5)	•	21.2	0.542	1206
TriSep Corp.	**8040-TS20-TSA**	302	2.7	69	5	1	45	4-11	2.5 - 11.5	1	•	•	•	1339
TriSep Corp.	**8040-TS40-TSA**	302	2.1	69	5	1	45	4-11	2.5 - 11.5	1	•	21.8	0.271	1281
TriSep Corp.	**8040-TS60-TSA**	302	2.7	69	5	1	45	4 - 11	2.5 - 11.5		•	•	•	931
TriSep Corp.	**8040-TS80-TSA**	303	2.1	69	5	i	45	4 - 11	2.5 - 11.5	0	•	15.7	0.195	921
TriSep Corp.	**8040-TS90-TSA**	302	2.7	69	5	1	45	4 - 11	2.5 - 11.5	0	•	,	•	1026
TriSep Corp.	**813-NF100** (CA)	•	6.9	•		•	80	2-12	•	<1	•		0.696	967
TriSep Corp.	**815-NF100 (CA)**	•	6.9	•	•	•	100	0.5 - 13	•	25	•	7.4	0.662	779
TriSep Corp.	**817-NF100** (CA)	•	6.9	•	•	•		0.5 - 13	•	25	•	7.7	0.693	866

Table 10.5. **Ultrafiltration** Membranes.

		Physical Characteristics					Performance at Test Conditions					
Manufacturer	Model No.	Comp.	Comfig.	Dia. (mm)	Len. (mm)	Eff. Are;a, m^2	Nominal Permeate Flow Rate, m^3/d	Permeate Flow Variation, %	Productivity Test Fluid	Feed Pressure, MPa	Feed Temp. °C	Recovery, %
A/G Technology	UFP-10-C-85	polysulfone	HF	108	1,310	13	32	25	clean water	0.14	2 5	>90
A/G Technology	UFP-100-E-85	polysulfone	HF	108	1,310	8.8	50	25	clean water	0.14	25	>90
A/G Technology	UFP-500-E-85	polysulfone	HF	108	1,310	8.8	100	25	clean water	0.14	2s	>90
Adv. Mem. Tech.	AES10-B-8040-H	polyether sulfone	SW	200	1,000	34.4	26	25	clean water	1.0	25	10
Adv. Mem. Tech,	AES30-B-8040-H	polyether sulfone	SW	200	1,000	34.4	33:	25	clean water	0.28	25	10
Adv. Mem. Tech.	AF10-B-8040-H	PVDF	SW	200	1,000	34.4	26	25	clean water	1.0	25	10
Adv. Mem. Tech.	AF100-B-8040-H	PVDF	SW	200	1,000	34.4	65	25	clean water	0.2	25	10
Adv. Mem. Tech.	AF3C-B-8040-H	PVDF	SW	200	1,000	34.4	33	25	clean water	0.28	25	10
Adv. Mem. Tech.	AF500-B-8040-H	PVDF	SW	200	1,000	34.4	65	25	clean water	0.2	25	10
Adv. Mem. Tech.	AGS5-B-8040-H	polyether sulfone	SW	200	1,000	34.4	26	25	clean water	1.0	25	10
Adv. Mem. Tech.	AS100-B-8040-H	polysulfone	SW	200	1,000	34.4	65	25	clean water	0.2	25	10
App. Mem. Inc.	M-C2020M7SH	Polyolefin	Tubular	43	2,870	0.66	5.4	•	clean water	0.5	25	2
App. Mem. Inc.	M-U8040D15	Thin Film	SW	203.2	1,016	*	23	25	clean water.	0.3	25	I0
App. Mem. Inc.	M-U8040D3	Thin Film	SW	203.2	1,016	*	26	25	clean water	1	25	10
App. Mem. Inc.	M-U8040D4	Thin Film	SW	203.2	1,016	*	21	25	clean water	OS	2.5	10
App. Mem. Inc.	M-U8040D500	Polysulfone	SW	203.2	1,016	*	136	25	clean water	0.2	25	10
Aqua Source, NA	LIB35	CA Polymers	HF	320	1,300	55	150	IO	clean water	0.04	20	*
Aaµr Source., NA	MIA35	CA Polymers	HF	114	1,300	26	19	2	clean water	0.04	20	•
CeraMem Sep. Inc.	PMA-0005-G	Gamma Aluminia	Cmpnd Tub.	144	864	11.1	2.8	N/A	clean water	0.17	25	N/A
CeraMem Sep. Inc.	PMA-0010-G	Titania	Cmpnd Tub.	144	864	11.1	3.2	N/A	clean water	0.17	25	N/A
Fluid Systems	ROGA7 8131UF Magnum	CDA	SW	203	1,524	48.8	49.2	20	SDTW	0.345	25	*
Fluid Systems	Ultra7 8321UF Magnum	PES	SW	203	1,524	30.7	64.3	*	SDTW	0.482	25	*
FPI Sep. & Sys.	8040-U100-TSA	Polysulfone	SW	202	1,016	33.4	88.8	15	clean water	‘0.2	25	15
FPI Sep. & Sys.	KC-840-010-SL	Polyether sulfone	SW	200	1,000	25.5	110	10	DI water	0.34	25	90
FPI Sep. & Sys.	KC-840-030-SL	Polyether sulfone	SW	200	1,000	25.5	135	10	DI water	0.34	25	90
FPI Sep. & Sys.	KC-840-100-SL	Polyether sulfone	SW	200	1,000	25.5	165	IO	DI water	0.34	25	90
Hydranautics	8040-FFF-2120	Hydophilic polyolefin	SW	203	1,016	23.3	26.5	*	DI water	0.2	25	10 - 20
Hvdranautics	8040-FFF-P100	Polvsulfone	SW	203	1.016	23.3	51.1	*	DI water	0.2	25	IO-20

Table 10.5. **Ultrafiltration** Membranes (continued)

		Rejection Characteristics					Operating Limits				
Manufacturer	Model No.	Nominal Molecular Mass Cutoff (NMMC)	Test Solute	Test Solute Concentration, mg/L	Feed Pressure, MPa	Feed Temp. °C	Max. Operating Pressure, MPa	Max. Operating Temp. °C	pH Range; continuous	pH Range; cleaning	Chlorine Tolerance, ppm or (ppm days)
A/G Technology	UFP-10-C-85	10 K	PVP K-15	*	*	25	0.5 @ 10 - 25 °C	80	2 - 12	2 - 13	100
A/G Technology	UFP-100-E-85	10 K	PVP K-90	*	*	25	0.5 @ 10 - 25 °C	80	2 - 12	2 - 13	100
A/G Technology	UFP-500-E-85	10 K	PVP K-90	*	*	25	0.5 @ 10 - 25 °C	80	2 - 12	2 - 13	100
Adv. Mem. Tech.	AES10-B-8040-H	3.5 K	polymer marker	2,000	0.34	25	2.8	50	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AES30-B-8040-H	10 K	polymer marker	2,000	0.34	25	2.8	50	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AF10-B-8040-H	3.5 K	polymer marker	2,000	0.34	is	2.8	50	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AF100-B-8040-H	100 K	polymer marker	2,000	0.34	25	2.8	50	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AF3C-B-8040-H	10 K	polymer marker	2,000	0.34	25	2.8	so	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AF500-B-8040-H	500 K	polymer marker	2,000	0.34	25	2.8	50	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AGS5-B-8040-H	2.5 K	polymer marker	2,000	0.34	25	2.8	50	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AS100-B-8040-H	100 K	polymer marker	2,000	0.34	25	2.8	50	2 - 11	2 - 12	(16,600)
App. Mem. Inc.	M-C2020M7SH	20000	*	[illegible]	*	*	[illegible]	40	2 - 12	*	*
App. Mem. Inc.	M-U8040D15	15000	*	*	*	*	*	*	*	*	*
App. Mem. Inc.	M-U8040D3	2500	*	*	*	*	*	*	*	*	*
App. Mem. Inc.	M-U8040D4	3500	*	*	*	*	*	*	*	*	*
App. Mem. Inc.	M-U8040D500	[illegible]	*	*	*	*	*	*	*	*	*
Aqua Source, NA	LIB35	100 K	Dexlrane	2,000	0.2	20	0.18	30	4 - 8.5	4-8.5	50
Aaua Source. NA	MIA35	100 K	Dexirane	2,000	0.2	20	0.18	30	4 - 8.5	4 - 8.5	50
CeraMem Sep. Inc.	PMA-0005-G	250 K	Polyvinyl-pyrrolidone	1,000	0.2	25	0.35	121 (aq)	4-9	4 - 9	Tolerant
CeraMem Sep. Inc.	PMA-0010-G	250 K	Polyvinyl-pyrroiidone	1,000	0.2	25	0.35	121 (aq)	3 - 12	3 - 12	Tolerant
Fluid Systems	ROGA7 8131UF Magnum	10 - 40 K	Dextran	*	0.345	25	1,034	40	4-6	2.5 - 8	1
Fluid Systems	Ultra7 8321UF Magnum	6 - 10 K	PEG	*	0.482	25	1.034@25, .827@ 50	50		1 - 13 @ 25, 1 - 11 @ 50	150 @ 50, pH 7 - 8
FPI Sep. & Sys.	8040-U100-TSA	100 K	Polyethylene glycol	1,000	0.2	25	2.7	45	2 - 12	2 - 12.5	10
FPI Sep. & Sys.	KC-840-010-SL	10 K	Myogiobin	1,000	0.38	25	0.69	50	2 - 10	1.5 - 13	200 ppm cleaning
FPI Sep. & Sys.	KC-840-030-SL	30 K	Bovine Albumin	2,000	0.38	25	0.69	50	2 - 10	1.5 - 13	200 ppm cleaning
FPI Sep. & Sys.	KC-840-100-SL	100 K	Globulin	2,000	0.38	25	0.69	50	2 - 10	1.5 - 13	200 ppm cleaning
Hydranautics	8040-FFF-2120	20 - 50 K	PEG	*	0.2	25	1.1	45	2 - 12	1 - 13	150
Hydranautics	8040-FFF-P100	100 K	PEG	*	0.2	25	1.1	45	2 - 12	1 - 13	150

Table 10.5. Ultrafiltration Membranes.

		Physical Characteristics					Performance at Test Conditions					
Manufacturer	**Model No.**	Comp.	**Comfig.**	Dia. (mm)	**Len. (mm)**	Eff. Arert, m^2	Nominal Permeate Flow Rate. m^3/d	Permeate Flow Variation, %	Productivity Test Fluid	Feed Pressure, MPa	Feed Temp. °C	Recovery, %
Hydranautics	**8040-TFF-P100**	thin film	SW	200	1,000	29.8	66.2	•	water	0.2		10 - 20
Hydranautics	**NTU-3306-K6R**	•	S W	*	*	*	15.5	*	*	0.1	*	*
Millipore	**H26MP01-43**	Polysulfone	HF	0.5	1,092	2.4	345 - 460	402.5	DI water	0.14	*	*
Millipore	**H53P10-20**	Polysulfone	HF	0.5	1,092	5	23 - 126	75	DI water	0.14	•	*
Millipore	**H53P100-20**	Polysulfone	HF	0.5	1,092	5	36 - 68	52	DI water	0.14	*	*
Millipore	**H53P3-20**	Polysulfone	HF	0.5	1,092	5	9-50	29.5	DI water	0.14	*	*
Millipore	**H53P30-20**	Polysulfone	HF	0.5	1,092	5	137 - 256	197	DI water	0.14	*	*
Osmoniw	**815-PT1** CA	Cellulose di-tri acetate	SW	201	1,016	28.1	46.2	*	2000 ppm NaCl	0.69	25	10
Osmonics	**815-PT1 PS**	Homogenous Polysufone	SW	201	1,016	28.1	72	•	2000 ppm NaCl	0.69	25	10
Osmoniw	**815-PT1** VS	Homogenous fluorocarbon	S W	201	1,016	28.1	118	*	2000 ppm NaCl	0.69	25	10
Osmonics	**815-PT2 CA**	Cellulose di-tri acetate	SW	201	1,016	28.1	56.9	*	2000 ppm NaCl	0.34	25	10
Osmoniw	**815-PT2** PS	Homogenous Polysufone	SW	201	1,016	28.1	72	*	2000 ppm NaCl	0.34	25	10
Osmonics	**815-PT2 VS**	Homogenous fluorocarbon	SW	201	1,016	28.1	99.3	*	2000 ppm NaCl	0.34	25	10
Osmoniw	**815-PT3** CA	Cellulose di-hi acetate	S W	201 ,	1.016	28.1	62.2	*	2000 ppm NaCl	0.34	25	10
Osmonics	**815-PT3 PS**	Homogenous Polysufone	SW	201	1,016	28.1	90.2	*	2000 ppm NaCl	0.34	2s	10
Osmontw	P-Series **PW8040F**	Polyether sulfone	S W	200.2	1,016	32.5	119.07	25	clean water	0.207	25	10
Osmonics/Desal	**G-10 GH8040F**	Thin Film Membrane	SW	200.2	1,016	32.5	25.7	25	clean water	1.034	25	10
Osmonics/Desal	**G-20 GK8040F**	Thin Film Membrane	S W	200.2	1,016	32.5	21.17	25	clean water	1.034	25	10
Osmonics/Desal	**G-50 GM8040F**	Thin Film Membrane	SW	200.2	1,016	32.5	23.44	25	clean water	0.517	25	10
Osmonics/Desal	**G-80 GN8040F**	Thin Film Membrane	SW	200.2	1,016	32.5	32.81	25	clean water	0.276	25	10
TriSep Corporation	**8040-UA50-TSA**	thin film	SW	202	1,016	33.4	40.6.	15	clean water	0.72	25	15
TriSep Corporation	**8040-UE50-TSA**	Polyether sulfone	S W	202	1,016	33.4	49.1	15	clean water	0.2	25	15

Table 10.5. Ultrafiltration Membranes (continued)

Manufacturer	Model No.	Rejection Characteristics: Nominal Molecular Mass Cutoff (NMMC)	Test Solute	Test Solute Concentration, mg/L	Feed Pressure, MPa	Feed Temp. °C	Operating Limits: Max. Operating Pressure, MPa	Max. Operating Temp. °C	pH Range; continuous	pH Range; cleaning	Chlorine Tolerance, ppm or (ppm days)
Hydranautics	**8040-TFF-P100**	100 K	PEG	*	0.2	25	1.1	45	2 - 12	1 - 13	150
Hydranautics	**NTU-3306-K6R**	6 k	Protein	*	*	*	*	*	*	*	*
Millipore	**H26MP01-43**	0.1 micron	Blue Dextran*	*	0.07	*	0.18	60	1.5 - 13.0	*	*
Millipore	**H53P10-20**	10 K	PVP K-30'	*	0.07	*	0.18	60	1.5 - 13.0	•	*
Millipore	**H53P100-20**	100 K	PVP K-90*	*	0.07	*	0.18	60	1.5 - 13.0	*	•
Millipore	**H53P3-20**	3 K	Insulin*	•	0.07	•	0.18	60	1.5 - 13.0	*	*
Millipore	**H53P30-20**	30 K	Albumin*	*	0.07	*	0.18	60	1.5 - 13.0	*	•
Osmonics	**815-PT1 CA**	5-10 K	*	•	•	✉	•	•	*	•	•
Osmonics	**815-PT1 PS**	5-10 K	*	*	*	*	*	*	*	*	*
Osmonics	**815-PT1** VS	5-10 K	•	*	•	✉	•		*	*	*
Osmonics	**815-PT2 CA**	10 - 20 K	*	*	*	*	*	*	*	*	*
Osmonics	**815-PT2 PS**	10 - 20 K	*	*	*	*	*	*	*	*	*
Osmonics	**815-PT2 VS**	10 - 20 K	*	*	*	*	*	*	*	*	*
Osmonics	**815-PT3** CA	SO-100	*	*	•	*	*	*	*	*	•
Osmonics	**815-PT3 PS**	50 - 100	*	*	*	*	*	*	*	•	*
Osmonics	P-Series **PW8040F**	10 K	**>96%** rejection of **Cytochrome-**C, **MM=13,300**	*	*	25	1.379	50	2 - 11	2 - 11.5	(5,000)
Osmonics/Desal	**G-10 GH8040F**	2.5 K	PEG	1,000	0.83	25	2.758	50	2 - 11	2 - 11.5	(20 - 50)
Osmonics/Desal	G-20 **GK8040F**	3.5 K	PEG	1,000	0.827	25	2.785	50	2 - 11	2 - 11.5	(500)
Osmonics/Desal	**G-50 GM8040F**	8 K	PEG	1,000	0.827	25	1.379	50	2 - 11	2 - 11.5	(1,000)
Osmonics/Desal	**G-80 GN8040F**	10 K	PEG	1,000	0.827	25	1.379	50	2 - 11	2 - 11.5	(1,000)
TriSep Corporation	**8040-UA50-TSA**	5 K	Polyethylene glycol	1,000	0.72	25	2.7	45	4 - 11	2.5 - 11.5	1
TriSep Corporation	**8040-UE50-TSA**	100 K	Polyethyleneglycol	1,000	0.2	25	2.7	45	2 - 12	2 - 12.5	10

Table 10.6. Microfiltration Membranes.

Manufacturer	Model No.	Physical Characteristics							
		Config.	Comp.	Dia. (mm)	Len. (mm)	Eff. Area, m^2	Pore Size, micron	Bubble Point (IPA), kPa	Foam Point (IPA), kPa
Adv. Mem. Tech.	AF 1U-B-8040-H	SW	PVDF	200	1,000	34.1	0.04	*	*
Adv. Mem. Tech.	AF 3U-B-8040-H	SW	PVDF	200	1,000	34.1	3.0	*	*
CeraMem Sep. Inc.	PMA-0200-G	Cmpnd. Tubular	Alpha Alumina	144	864	11.1	0.2	*	*
CeraMem Sep. Inc.	PMA-0500-G	Cmpnd. Tubular	Mixed Oxide	144	864	11.1	0.5	*	*
Graver Sep. Inc.	Various	Tubular	polyester	65	250	0.68	0.6	*	*
Graver Sep. Inc.	Various	Tubular	polyester	65	500	1.36	0.6	*	*
Graver Sep. Inc.	Various	Tubular	polyester	65	750	2.04	0.6	*	*
Graver Sep. Inc.	Various	Tubular	polyester	65	1,000	2.12	0.6	*	*
Graver Sep. Inc.	Various	Tubular	316L Stainless steel	25 - 1200	600 - 6100	0.05 - 750	0.1	*	*
Osmonics/Desal	E-500 EW8040F	SW	polysulfone	200.2	1,016	32.52	0.04	*	*
Osmonics/Desal	JX8040F	SW	PVDF	200.2	1,016	32.52	0.3	138	152
Osmonics/Desal	K8040	SW	PTFE	200.1	1,016	18.58	.1, .2, .5, 1, 3	*	*
Setec, Inc.	SC050	Tubular	polypropylene	38	460	0.5	9.2	*	*
Setec, Inc.	SC170	Tubular	polypropylene	50	1,060	1.6	0.2	*	*
TriSep Corp.	8040-TM10-TSA	SW	PVDF	202	1,016	33.4	0.2	*	*
Water Equip Tech.	WBC-0.2P9-15	SW	CPVC/Polypropylene	114.3	247.1	1.11	0.2	*	*
Water Equip. Tech.	WBC-0.4P49-15	SW	CPVC/Polypropylene	114.3	247.7	1.11	0.4	*	*
Water Equip. Tech.	WBC-1.0-15	SW	Cotton/Polyester	114.3	247.6	1.39	1	*	*
Water Equip. Tech.	WBRS-5-15	SW	Cotton/Polyester	114.3	247.6	1.39	5	*	*
Water Equip. Tech.	WVSS-20-15	SW	Cotton/Polyester	114.3	247.6	1.39	20	*	*

Table 10.6. Microfiltration Membranes (continued)

		Performance at Test Conditions						Operating Limits				
Manufacturer	Model No.	Nominal Permeate Flow Rate, m³/d	Permeate Flow Variation, %	Productivity Test Solution	Feed Pressure, kPa	Feed Temp. °C	Recovery %	Max. Operating Press., kPa or (Pressure Drop)	Max. Operating Temperature, °C	pH Range; continuous	pH Range; cleaning	Chlorine Tolerance, ppm or (ppm days)
Adv. Mem. Tech.	AF 1U-B-8040-H	150	25	clean water	0.2	25	10	1,030.	50	2 - 11	2 - 12	(16,600)
Adv. Mem. Tech.	AF 3U-B-8040-H	300	25	clean water	0.2	25	10	1,030	50	2 - 11	2 - 12	(16,600)
CeraMem Sep. Inc.	PMA-0200-G	6.8	N/A	clean water	0.17	25	N/A	345	121	3 - 12	3 - 12	Tolerant
CeraMem Sep. Inc.	PMA-0500-G	12.8	N/A	clean water	0.17	25	N/A	345	121	3 - 12	3 - 12	Tolerant
Graver Sep. Inc.	Various	19.1	5	water	2.9	25	*	517	60	4 - 9.5	N/A	*
Graver Sep. Inc.	Various	38.2	5	water	2.9	25	*	517	60	4 - 9.5	N/A	*
Graver Sep. Inc.	Various	57.3	5	water	2.9	25	*	517	60	4 - 9.5	N/A	*
Graver Sep. Inc.	Various	76.4	5	water	2.9	25	*	517	60	4 - 9.5	N/A	*
Graver Sep. Inc.	Various	various	10	clean water	0 - 7000	0 - 300	0 - 99	unlimited	400	0 - 14	0 - 14	varies
Osmonics/Desal	E-500 EW8040F	136.08	25	clean water	0.207	25	10	1,030	50	2 - 11	2 - 11.5	(5,000)
Osmonics/Desal	JX8040F	160.9	25	clean water	0.207	25		1,034	50	2 - 11	1 - 13	(5,000)
Osmonics/Desal	K8040	*	*	*	*	*	*	138	50	1 - 11	1 - 13	(5,000+)
Setec, Inc.	SC050	72	10	clean water	70	21	10	209	40	1 - 14	1 - 14	0
Setec, Inc.	SC170	230	10	clean water	70	21	10	209	40	1 - 14	1 - 14	0
TriSep Corp.	8040-TM10-TSA	34	15	clean water	70.0	25	15	2,760	45	2 - 12	2 - 12.5	10
Water Equip. Tech.	WBC-0.2P9-15	102	*	clean water	*	*	*	20	90	12 @ 35 de	*	Tolerant
Water Equip. Tech.	WBC-0.4P49-15	109	*	clean water	*	*	*	20	90	12 @ 35 de	*	Tolerant
Water Equip. Tech.	WBC-1.0-15	245	*	clean water	0.275	*	*	34	90	- 12 at 35	*	Tolerant
Water Equip. Tech.	WBRS-5-15	272	*	clean water	0.275	*	*	27.5	90	- 12 at 35	*	Tolerant
Water Equip. Tech.	WVSS-20-15	272	*	clean water	0.275	*	*	20.6	90	- 12 at 35	*	Tolerant

10.2 What Does It All Mean?

There are so many products available with slightly different rejection and production rates that are tested under different conditions! How is one to decide which one to get? The bad news is that the most important information is not on the specification sheet. Cost and service are the most important aspects of the membrane. Cost could be a deciding factor, but it changes with time and depends on the quantity purchased and other nebulous site specific situations. Service contracts are crucial. Membranes are a large part of the cost of a treatment plant and they can be ruined in a matter of minutes if not used properly. Be sure that the supplier will back up the warranty and be sure that none of the exclusionary conditions occurs in your facility.

Here are a few tips for ensuring a successful membrane treatment experience.

- Make sure the company that supplies your membranes is going to help you get started and will be available later when problems come up.
- Make sure there is a good warranty on the membranes.
- Invest in your operators. Well trained and empowered operators save money and aggravation.
- Consider automatic data acquisition and controls so that operators can tune in on gradual performance trends rather than day to day changes. Membrane data needs normalization to detect real changes, shift to shift changes in the raw data are not as meaningful.
- Calibrate sensors regularly to ensure data is as accurate as possible.

10.2.1 Water and Salt Transport Coefficients Versus Rejection and Productivity

Water (A) and salt (B) transport coefficients are difficult to get a handle on. They are extremely small numbers calculated from the productivity, rejection, and test conditions. A is the volume of water produced per net unit of pressure, area, and time at test conditions. B is the quantity of salt that passes through the membrane per unit of difference in salt concentration across the membrane for a unit of area in one second under test conditions. As discussed in chapter 4, if one wants to produce a large quantity of low salt water, one would want a membrane with a relatively large A value and a small B value. Figure 10.1 compares the A and B values for seawater, brackish water RO and NF membranes. Figure 10.2 focuses on RO membrane specifically, which will be the basis of the rest of this discussion. Generally, seawater membranes have very small A and B values, brackish water membranes have slightly higher values, and NF membranes have large values. The entire range of desalination membranes is close to zero to 4x 10” m/sec for B and close to zero to $35x10^{-12}$ $m^3m^{-2}sec^{-1}Pa^{-1}$ for A.

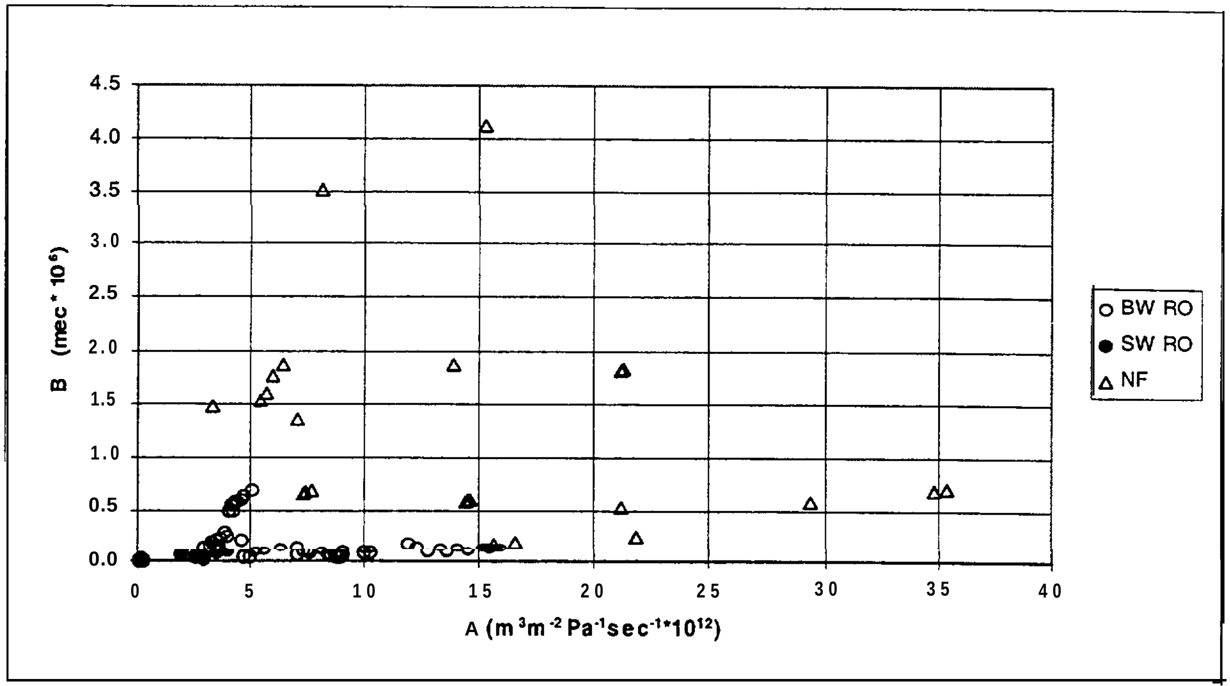


Figure 10.1.-Relation between salt and water transport coefficients for seawater, brackish water RO and NF membrane. There are two different optimization strategies apparent here: one group of membranes is for minimizing salt passage and concentrate volume, the other is for selective salt passage and minimizing concentrate TDS.

Though it might look from these graphs like there is a linear relationship between A and B, there is not. The value B/A is merely an artifact of the test conditions; it is the product of the salt passage for the test solution and the NDP under test conditions. Remember that these parameters are calculated from observed performance of the entire module, not from physical characteristics of the membrane as they should be. Still, the visual aid is helpful in portraying the performance of the a large population of membranes. A more informative graph would be A or B versus applied pressure and feed water TDS. Alas, this data is not available for all membranes. But what about the plain unprocessed specification sheet data? Salt rejection and productivity per unit area are explored in figure 10.3. As with A and B, there is absolutely no correlation between the two, except that in this figure it is readily apparent. It would be difficult to predict which of these membranes would be best for a particular application. There are too many variables that are not taken into consideration. Are the differences in productivity for a given rejection rate due to test conditions or could the module furthest to the top right actually have better circulation than the others? This is because productivity and rejection are extremely sensitive to the operating conditions. There must be a better way to determine the efficiency of a membrane module. One needs to be able to decide whether it is better to pay extra for a module with 99.5 percent rejection if it produces 6 m^3 less per day than the 99 percent rejection membrane.

Unfortunately, there is no way to be sure which membrane will be best for a particular application until it is tested on site. Until all membranes intended for a given purpose are

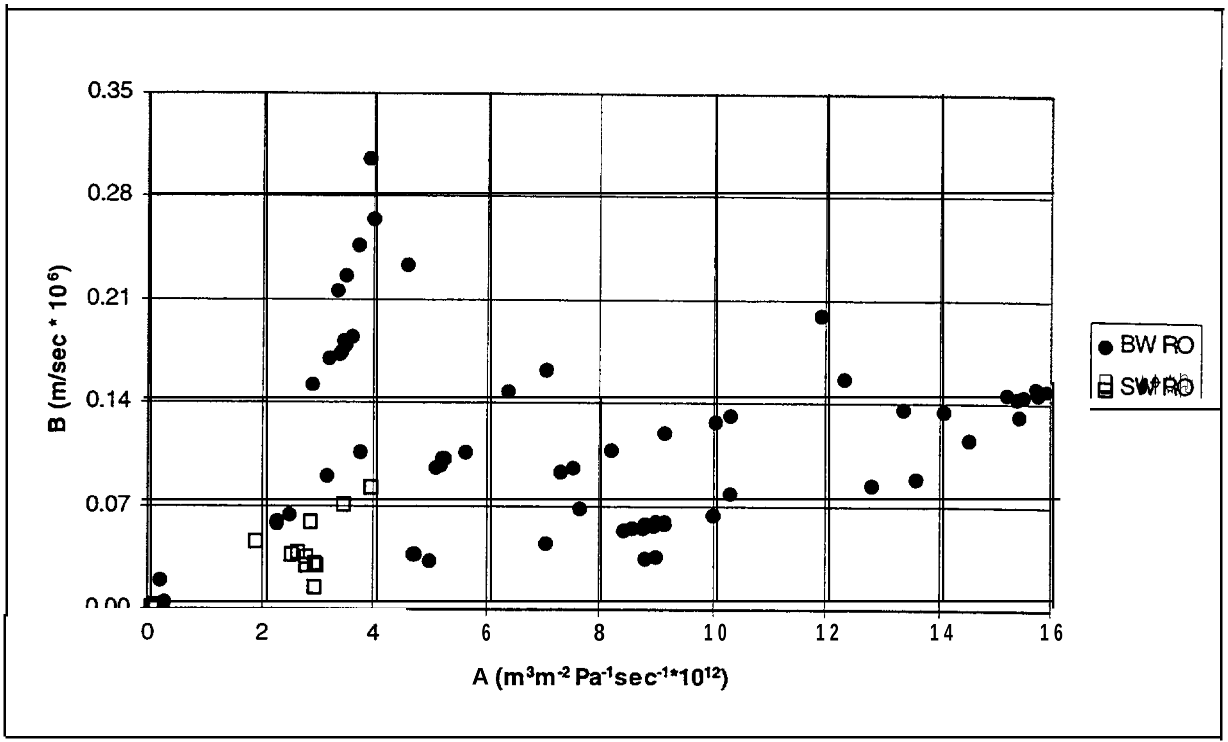


Figure 10.2.—Close up of relationship between sea and brackish water RO membrane transport coefficients.

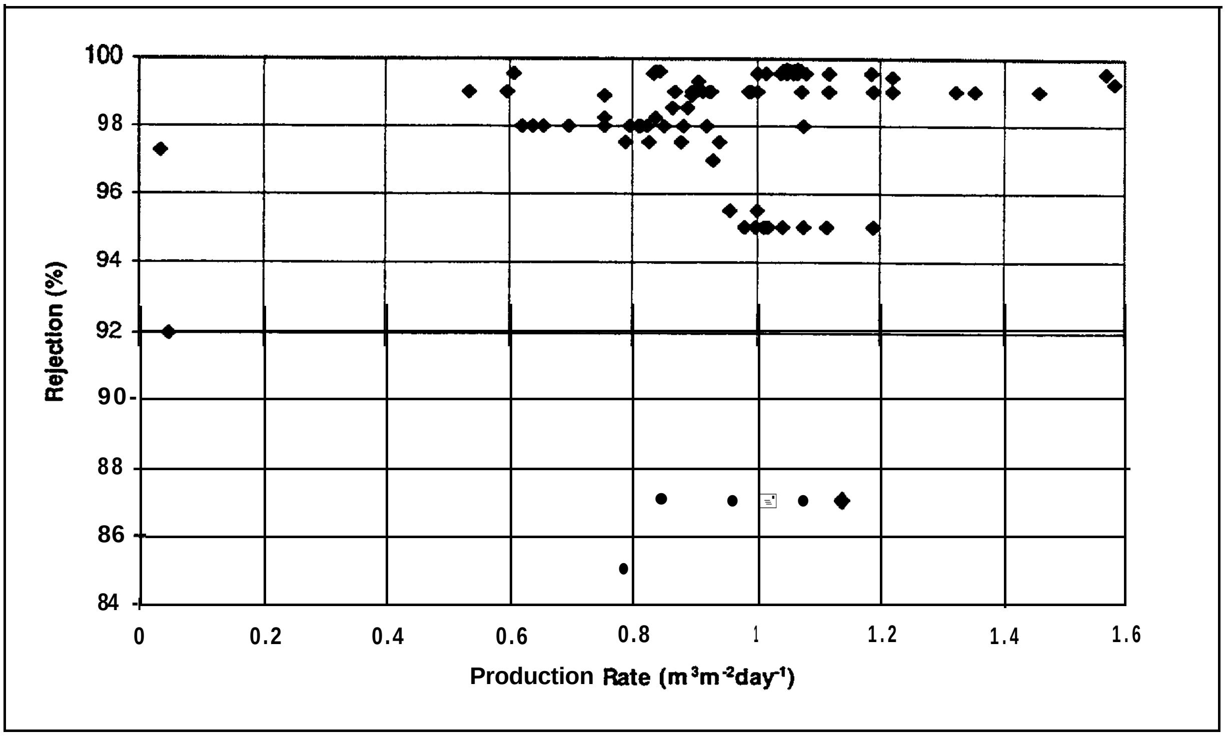


Figure 10.3.—No relationship between rejection rate and productivity per unit area.

under the same conditions, or at least with the same feed solution at the same pH and temperature, there is no meaningful way to compare them. With performance test data on many different membranes, though, one can make some comparisons between different configurations and testing methods.

10.2.2 Does Operating Pressure Make a Real Difference?

If a membrane will have the same rejection at a lower pressure with perhaps a small reduction in productivity, it may be cost effective to use the lower the pressure. However, there is a minimum pressure necessary to overcome the resistance due to osmotic pressure difference between the feed and permeate; the membrane resistance, which is a function of the polymer and its thickness; and the module resistance, which is a function of the flow channel width and surface roughness. There is a trade-off involved in optimizing any of these three resistances.

- Lowering the osmotic pressure difference means more salt in the product water,
- Decreasing membrane resistance usually is accompanied by increased salt transport, and
- Lowering module resistance could mean less contact time and lower permeability.

But, are the current module designs at the optimum point? Conveniently, a few manufacturers supply performance data for a membrane at more than one operating pressure Table 10.7 lists differences in performance parameters when tested at different operating pressures.

Table 10.7.—Change in performance statistics with operating pressure. Membrane X may have better performance at the higher pressure

Membrane	Net driving pressure (kPa)	Productivity (m^3/day)	Rejection (%)	A	B
X	861	34	98	11.9	0.20
	1,369	54	99	12.3	0.16
Y	481	20.1	98	15.8	0.15
	646	27.3	98.5	15.9	0.15
	996	41.6	99	15.7	0.15
Z	481	23.8	98	15.4	0.14
	646	32.2	98.5	15.5	0.14
	1006	49.2	99	15.2	0.15

The only real difference in performance at higher pressure is the decrease in salt transfer (B) for membrane X at higher pressure. Membranes Y and Z have essentially the same performance no matter what pressure they are operated at. They may have a slight improvement in efficiency at mid-range pressures, but the increase in productivity per unit NDP is too small to be significant. Membranes Y and Z cannot be compared fairly with membrane X because they were tested with only 500 mg/L NaCl solution while the latter was tested with 2000 ppm NaCl. Table 10.7 should be used only to compare each membrane to its own performance at a different pressure.

10.2.3 Does Increasing Membrane Area per Module Improve Productivity?

Several of the membranes listed are available in "high area" module designs. One way to measure the benefit of adding more area to a module is to look at the overall module efficiency. Module efficiency, as it is used here, is the production rate under standard conditions divided by the module volume. Figure 10.4 compares the module efficiency per square meter of added membrane area and performance statistics for 20 pairs of modules made with the same membrane, but constructed such that one has a greater surface area than the other. All increases in area resulted in a net increase in the modules production rate; otherwise, they would not be on the market. From figure 10.4, however, it is apparent that some increases are more effective than others. There are only a few ways to increase the area in a spiral wound element without increasing the volume.

- ***The membrane envelopes can be constructed such that the edge seam is very narrow.*** Hand glued envelopes can have edges that are 6 - 10 cm wide on three sides! Using a mechanical sealing machine could increase the area per envelope by 25 percent. This method should not change the A or B values. Small changes are probably due to the error in the area measurement of the regular modules. There is also an added value in that there are no longer dead spaces on the module edges that can trap particles and provide good places for fungal growth.

- ***Thinner spacer materials can be used.*** Thinner spacer material could improve or hinder overall performance depending on whether the new material is better at promoting mixing than the regular materials.

- ***The membrane envelopes can be wound more tightly.*** This method is sure to hinder performance which will show up as a decrease in A and an increase in B over the regular module. Tighter winding inhibits circulation and will promote fouling.

10.2.4 Are Longer Modules Better?

One recent change in module construction is to increase the length by one half so that two of the newer "MAG" modules should be equivalent to three regular size modules. The benefits of having longer modules are:

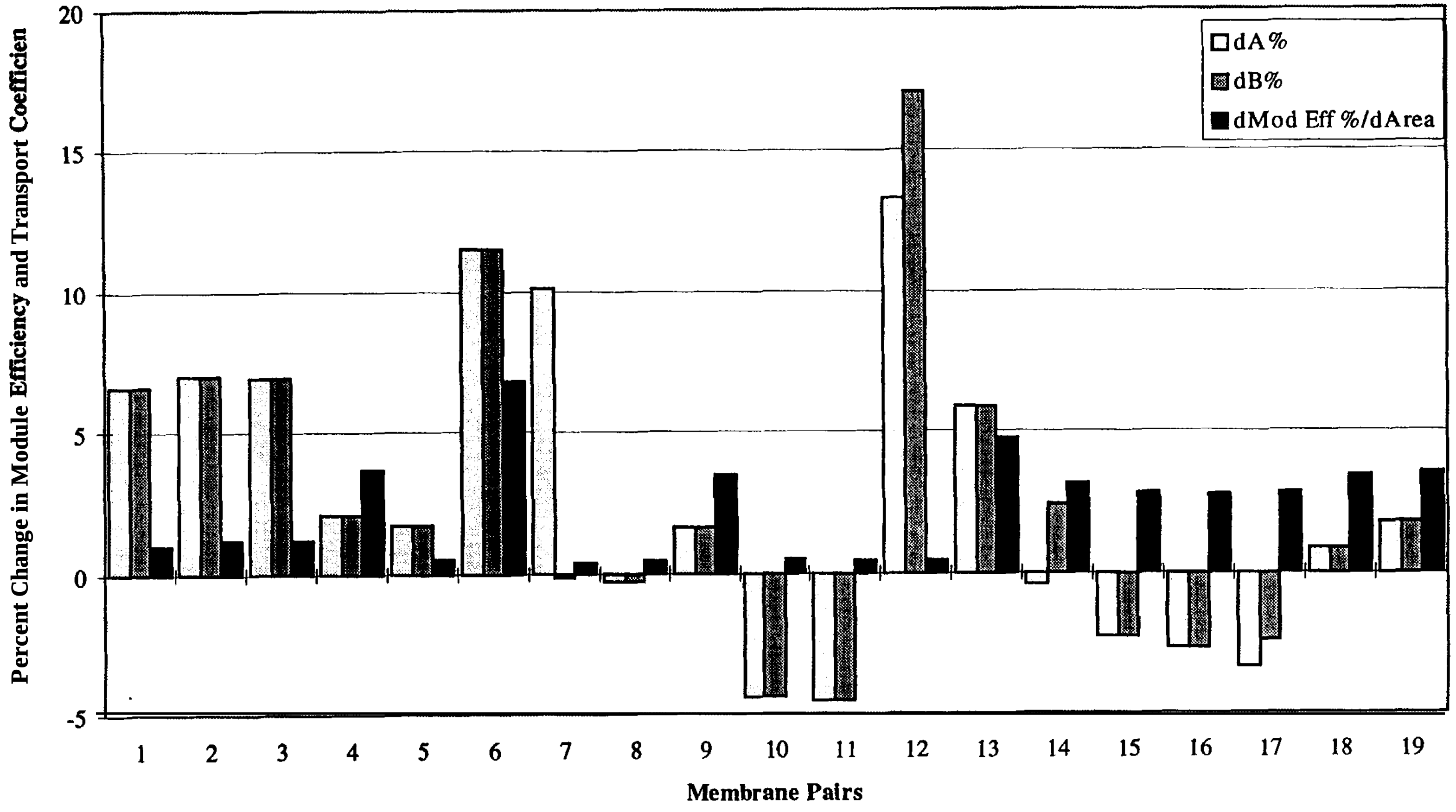


Figure 10.4.—Effect of increased module area on module efficiency per m^2 added area, water (A) and salt (B) transport coefficients. Change in module efficiency is the difference in production rate of high and standard area modules tested under the same conditions. Increases in B and decreases in A are counter productive. Both can be caused by hindering circulation within the module, which effectively decreases the useful area.

- Once the capital expense of the new equipment for manufacturing the modules has been realized, it should cut production expenses by about one third.
- Fewer connectors are needed, which means a small equipment savings and fewer opportunities for leaks.
- Greater active membrane area due to fewer seams and opportunities for an edge effect.

On the down side:

- The modules are heavier and will require an hydraulic lift to load them into the pressure vessels. They will also be more difficult to extract.
- Since the manufacturers had to up grade their equipment to produce these longer elements, they will probably be more expensive than the equivalent number of standard size elements.

How should a longer element perform in comparison to the standard size? Figure 10.5 compares five sets of modules that come in a 1 m length and also a 1.5 m length. Each pair was performance tested with the same feed solution, NDP, temperature, and pH. The recovery rate was higher for the 1.5 m module, as should be expected. Theoretically, A and B should not be affected by a change in the module length. The increase in productivity should be due to a reduction in edge effects caused by poor circulation in the membrane envelope comers. Since the specifications are based on single module performance, though, some of the improvement is probably the result of the higher feed flow necessary for a longer element. But if both size modules were tested in full pressure vessels, both would have the same feed flow, and the only differences would be due to the reduction in edge effects.

10.2.6 Conclusions

There are countless membrane products to choose from in a wide variety of materials, configurations, and sizes, with a wide range of rejection rates and production rates, and there will surely be many more in the future. They all operate under the same principles, however. As long as one can find out how they were tested to produce the advertized rejection and production rates, one can normalize the data for testing differences and compare the performance of different products. It is very important to look at more than two parameters, though, when the products are tested under different conditions or are of different sizes. While module efficiency may be not quite as important as $/kWhr, it is still helpful when deciding between modules of different configurations. Transport coefficients are useful by themselves only when products of the same configuration are tested under the same conditions.

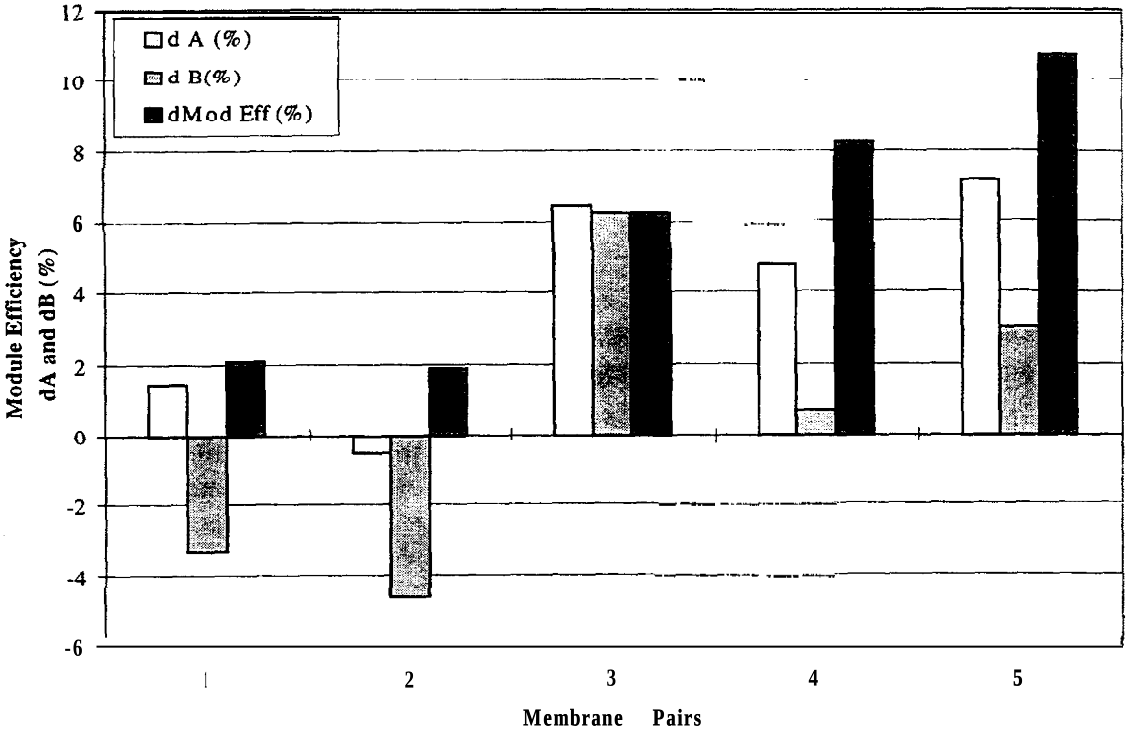


Figure 10.5.—Change in performance with longer modules.

No matter what the specifications are, service is still the ultimate consideration. Membranes are an expensive part of a water treatment plant and can be easily ruined. Whether you are purchasing membranes for your kitchen sink or a whole plant, your supplier should work with you to ensure you are getting the right product for your application.

Chapter 11
ESTIMATING THE COST OF MEMBRANE TREATMENT

The cost of membrane processes depends primarily on the quality of the source water, the method of concentrate disposal, and the quantity of water to be produced. The feed water quality and treatment objectives dictate the type of membrane to use, which, in turn, influence the operating pressure and the composition and volume of the concentrate stream. For this reason, concentrate disposal must be a part of the overall treatment objectives. There are two general approaches to disposing of concentrate. The first is dependent on volume and the second is dependent on concentration. A goal must be established at the start of the planning process: either to minimize concentrate volume or to minimize its salt content. Blending the membrane product water with filtered feed water is one way to help control concentrate characteristics, but first one needs to predict what the concentrate and product composition will be.

11.1 Predicting Product and Concentrate Composition and Volume

There are four important pieces of information needed to predict the product and concentrate composition and volume:

1) Recovery rate, (Rec) in decimal form. The recovery rate is limited by the concentration of sparingly soluble salts in the feed water. Lowering the pH and adding anti-sealants can increase the potential recovery rate. The other determining factor is the configuration of the membrane system. Each element can recover approximately 10 percent of the feed flow as product. Generally, 50 percent recovery is assumed for a 6-element vessel. Therefore, a 2: 1 array should be expected to achieve 75 percent recovery, and a 4:2: 1 array should be capable of 87.5 percent.

2) Rejection rate, Rej in decimal form. Manufacturers list a rejection rate for chloride and one for sulfate or other divalent ions for NF membranes. For greater accuracy, use a weighted average based on the feed water composition. For instance, if the feed water has a ratio of 3: 1 mono-valent to multi-valent ions and the rejection rates are 90 percent for chloride and 99.5 percent for sulfate, the weighted average rejection rate would be:

$$Rej_{ave} = \frac{0.75*0.900 + 0.25*0.995}{0.75+0.25} = 0\ 924 \qquad \text{Eq. 11.1}$$

If the goal is to minimize concentrate volume, choose a membrane with a very high rejection. However, if the goal is to minimize concentrate TDS, choose a membrane that will produce the target water quality. NF membranes are sufficient in many cases.

3) Feed water dissolved solids concentration, C_f in mg/L.
4) Target delivery water concentration after blending, C, in mg/L.

A very simple model of membrane separation is used here for the purpose of estimating costs. Accurate product and concentrate concentration prediction calculations that take concentration polarization into consideration can get quite complex, but do not provide that much more accuracy in a first pass cost estimate.

Product concentration, C_p in mg/L:

$$C_p = \frac{C_f \ (1 - Rej)}{Rec} \qquad \text{Eq. 11.2}$$

Concentrate concentration, C_c in mg/L

$$C_c = \frac{C_f \ Rej}{1 - Rec} \qquad \text{Eq. 11.3}$$

The maximum amount of blend water that can be mixed with the membrane product and still achieve the target water quality is calculated as follows, assuming filtered feed water is used for the blend water:

$$Q_b = \frac{Q_t \ (C_t - C_p)}{C_f - c_p} \qquad \text{Eq. 11.4}$$

Where Q_b is the maximum blend volume in m^3/day, Q_t is the target volume in m^3/day, and C_t is the target dissolved solids concentration in mg/L. If there is a component of the blend water that is more limiting than the total dissolved solids, there are two options. Either plan to remove that component from the blend water or use the concentration of that component in the blend water for C_f and the estimated remaining concentration of it in the membrane product water for C_p. As an example, consider the following situation:

C_f = 900 mg/L with 0.5 mg/L manganese
Rej = 0.95
Rec = 0.85
C_t = 300 mg/L with less than 0.05 mg/L manganese
C_p = $900*(1-0.95)/0.85 = 56$ mg/L
C_b = $(300-56)/(900-56) = 0.29$ or 29 percent blending with feed water.

But, when the manganese concentration is considered as the limiting component:

C_f = 0.5 mg/L manganese
Rej = *0.95*
Rec = *0.85*
C_t = Less than 0.05 mg/L manganese
C_p = 0.5*(1-0.95)/0.85 = 0.03 mg/L
C_b = (0.05-0.03)/(0.5-0.03) = 0.04 or 4 percent blending with feed water.

If the blend water is filtered with greensand or the manganese is removed in some other way, the higher level of blending is possible, otherwise not. However it is decided, once the blend volume has been established, the membrane process feed, product, and concentrate flows are set (all in m^3/day):

$$Q_p = Q_t - Q_b \qquad \text{Eq. 11.5}$$

$$Q_f = \frac{Q_p}{rec} \qquad \text{Eq. 11.6}$$

$$Q_c = \frac{Q_p\,(1 - Rec)}{Rec} \qquad \text{Eq. 11.7}$$

11.2 Estimating Membrane Cost Factors

The primary factors for estimating membrane system costs are the membrane area, operating pressure, pump horsepower, and the number of skids.

11.2.1 Membrane Area

Membrane productivity per day is normally measured under some standard operating conditions and listed as m^3/day or gal/day. At this point, most manufacturers have there own standards. The listed productivity will do for a preliminary estimate of membrane area requirements, but a pilot test will be needed to verify the productivity on site. Perusing the tables in chapter 10, you will notice quite a variation of productivity levels available even for the same kind of membrane. For a preliminary estimate, take a moderate value from the range listed. The number of modules needed is simply the required membrane capacity divided by the module productivity:

$$No.\ of\ modules\ =\ \frac{Q_p}{Q_m} \qquad \text{Eq. 11.8}$$

Where Q_m is the module productivity in m^3/day. Of course there should be a full compliment of modules in each vessel and enough vessels for a complete array. If a 2: 1 array is planned, then the number of modules should be evenly divisible by 3. Likewise for a 4:2: 1 array, the number of modules should be a product of 7.

11.2.2 Operating Pressure

The operating pressure is a function of the type of membrane used and the difference between the osmotic pressure of the feed water and product water. For this level of estimate, we will use the pressures recommended by the manufacturers. See chapter 10 for sample operating pressures.

11.2.3 Pump Horsepower

Using the following assumptions:

- Feed water is being pumped from a tank of approximately the same height as the membrane skid,
- 10 meters of pipe, and
- Pipe is 10 cm (4 in.) in diameter for 20 cm (8 in.) modules and 5 cm (2 in.) for 10 cm (4 in.) modules.

$$hp\ =\ \frac{(Ah\ g\ +\ 0.5\ v^2\ +\ \Delta p)\ (1 - E_{rec})\ Q_f\ 1000}{746\ E_{eff}} \qquad \text{Eq. 11.9}$$

Where:

Ah = height difference between top of tank and membrane inlet in m,
g = gravitational constant, 9.81 m/s^2,
v = velocity = Q_f/pipe area, m/s,
Ap = pressure difference between tank and operating pressure in kPa,
E_{rec} = energy recovery in decimal, 0.20 - 0.30 depending on concentrate pressure,
Q_f = membrane feed flow, m^3/sec,
1000 = mass of one m^3 of water in kg,

746 = conversion factor from J/s to hp,
E_{eff} = combined pump and motor efficiency in decimal, try 0.75.

11.2.4 Number of Skids

There are several cost items that are directly tied to the number of skids in the system. The most important item is the high pressure pump. The flow rate used in the horsepower calculation above is the flow per skid. If all the flow is assumed to go to one skid for large flowrates, the cost of the pump skyrockets. When choosing the number of skids, there are three important considerations:

1) There must be a complete array on each skid. If a 2: 1 array is planned, then a complete 2: 1 array should reside on each skid.

2) It is good to have two medium sized skids rather than one large skid. This is true especially if the system is in operation 24 hours a day. If there is a problem with one system, the other can still provide some water while the problem system is being maintained.

3) It is less expensive to build fewer large skids.

As always, there is a trade off between expense and flexibility. The designer has to balance the pumping costs with the other per skid costs such as chemical feed systems, controls and data acquisition and the extra lab costs for testing the water quality of each skid. Figure 11.1 shows that, for smaller capacity systems, one skid is most economical, but very quickly the extra cost associated with multiple skids is more than outweighed by the lower pump costs.

11.2.5 Chemical Feed Rates

Chemicals needed for RO and NF pre- and post-treatment and an estimate of the dose rate:

- Acid for pH adjustment to prevent scaling or for membrane preservation when using CA membrane. Dosage depends on the buffering capacity of the water and the target pH. Jar tests are best.

- Anti-scalants for scaling prevention. A typical dosage is about 5 mg/L.

- Chlorine, for disinfection after treatment. 5 mg/L is a good estimate for RO or NF product water. Blend waters will require more, depending on the organic content.

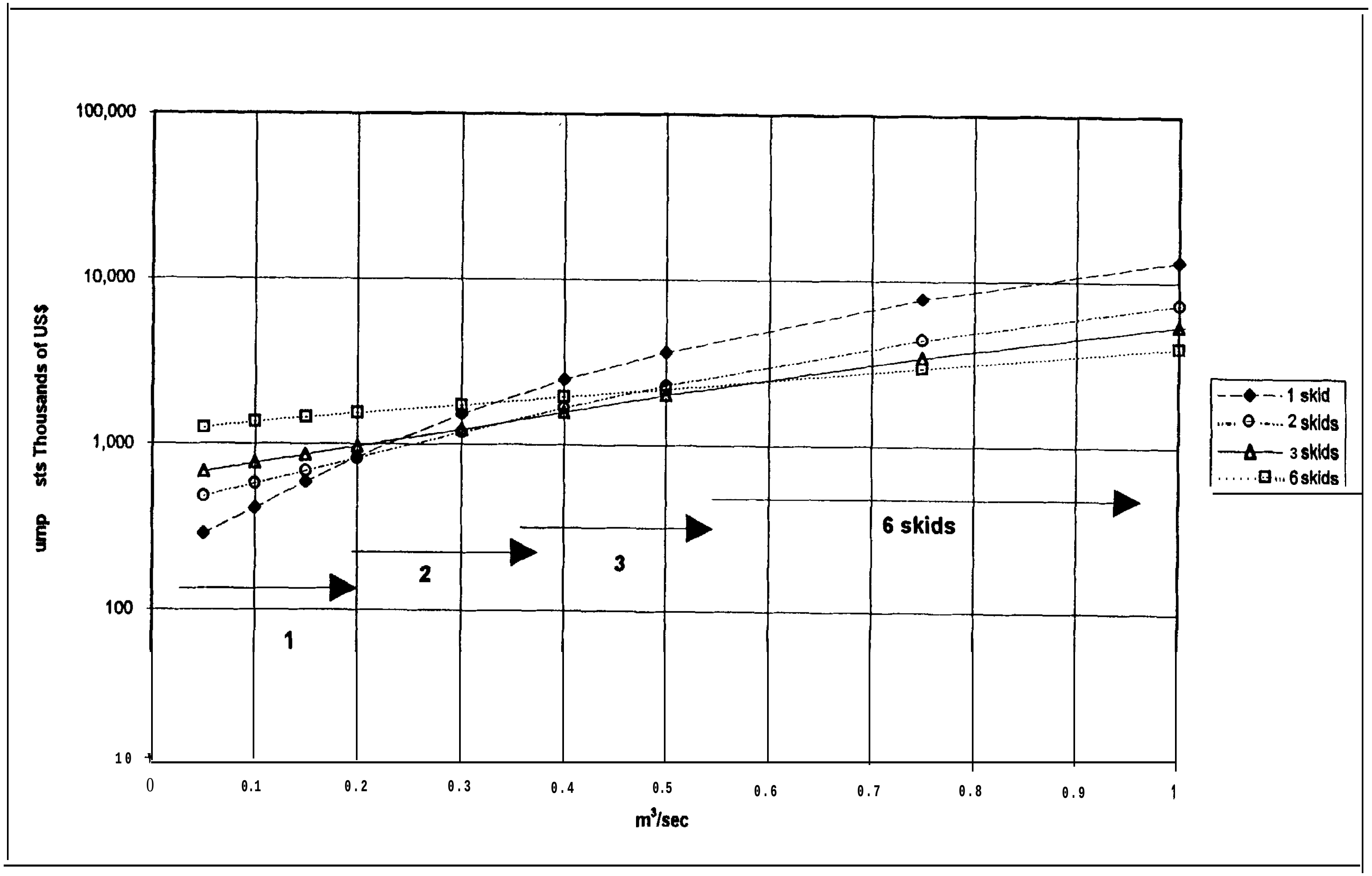


Figure 11 .1 .-Optimizing pump cost and the number of skids for a range of flow rates (0.1 m³/sec = 2.2 Mgal/d) Operating pressure is 1,350 kPa; pump efficiency is 75 percent.

11.3 Calculating Costs

The cost relations presented here for direct and indirect construction costs and O&M costs have been adapted from a paper by William B. Suratt of Camp Dresser and Mckee (Suratt, 1995). They apply particularly to RO and NF systems; with modifications, they could also apply to UF and MF as will be discussed below.

Table 11.1.-Direct construction costs

Cost item	Equation	Comments
Membrane modules	$\$*\text{number of modules}$	Use $700 for 20.2 cm (8") diameter, $500 for 10.1 cm (4") diameter and $250 for 6.4 cm (2.5") diameter vessels
Skids with plumbing	$\$*\text{ number of vessels}$	Assume 6 or 7 elements per vessel, use $5,000 for 20.2 cm diameter, $2,000 for 10.1 cm diameter and $900 for 6.4 cm diameter vessels
Building	(21 m^2 • number of skids + 50 m^2 admin. space)*$\$1,000/m^2$	Based on number of skids
Electrical hookups	$\$614*m^3/day^{0.65}$	Based on capacity
Instrumentation and controls	$\$100,000+\$80,000*$ number of skids	Based on number of skids
Booster pumps	$\$*(hp/100)^{0.65}$	Use $85,000 for vertical turbine 316L stainless steel with variable speed motor, $58,000 for vertical turbine 316L stainless steel with single speed drive and $35,000 for a single speed ANSI stainless steel horizontal pump
Product water degasifiers	$1.5006 * m^3/day + 3766$	Based on capacity
Odor control	$320.9*(m^3/day)^{0.6}$	Based on capacity
Process piping	$15.852*(m^3/day)$	Based on feed flow (capacity/recovery)
Yard piping	$100*(m^3/day)^{0.75}$	Based on feed flow (capacity/recovery)
Chemical feed systems	Storage (L) $+\$*\text{skids}*$ chemical injection points/skid	Based on chemical dosage rates and $1/L of storage + $30,000 for acid or caustic metering systems and $20,000 for antiscalant and chlorine metering systems.
Cartridge filters	$135,000*\text{skids}*(m^3/sec)^{0.8}$	Based on the capacity per skid and total number of skids. includes filters, filter housing, and plumbing.
Membrane cleaning equipment	$67,000	Two 4,000 m^3 tanks, recirculation pump, cartridge filter, mixer, piping, and electrical.
Concentrate treatment	$\$15*(m^3 \text{ concentrate/day})$	Conservative estimate for in-line aeration and surface water discharge.
Generator	$\$150,000*MW^{0.85} + \$50,000+$	Based on building and pumping power costs. Assume 4 W per m^3/day feed capacity plus 746 W per hp for each pump.
Site work	$\$15 * (m^3/day)$	Based on total plant capacity (membrane product + blend)
Contract engineering and training	$50,000	

Table 11.2.—Indirect construction costs

Cost item	Equation	Comments
Mobilization, insurance and bonds	0.05 * total direct cost	
Contingencies	0.40 * total direct cost	

Table 11.3.—Operation and maintenance costs

Cost item	Equation	Comments
Electricity	$/kWhr . kW . hr/yr	Normal estimates are $.04 - $.08 per kWhr
Labor	$/yr * Employees	$40,000 per year average
Chemicals	$/L . L/yr	
Cleaning supplies	1.3 . vol of system(m^3) * cleanings/yr * $50	Assuming an acid and a caustic cleaning for each cleaning session at a cost of $50/$m^3$ cleaning solution
Membrane replacement	$\frac{\text{\# modules . \$/module}}{\text{life expectancy}}$	Assume 3-year life expectancy
Cartridge filters	(23,100 . m^3/sec-6.25) . skids * 12 times/yr	Based on skid feed capacity
Repairs and replacements	0.005 . DCC	Based on direct construction cost
Insurance	0.002 * DCC	Based on direct construction cost
Lab fees	800 * skids * 12	$800 per skid for each set of analyses, 12 analyses per year
Capital recovery cost	Use payment function in your favorite spreadsheet program	Based on given interest rate and amortization time

Figures 11.2 and 11.3 show how construction and O&M costs change with capacity and with total dissolved solids. Assumptions involved are listed in table 11.4. These costs are very sensitive the number of vessels per skid. There are many cost items that depend on the number of skids: pumps, instrumentation, chemical feed systems, and building size. O&M costs are not as sensitive to skid size. For example, energy costs would be similar if used on one large pump or two smaller ones. Figure 11.4 illustrates the effect on cost of unit size or the number of vessels per skid.

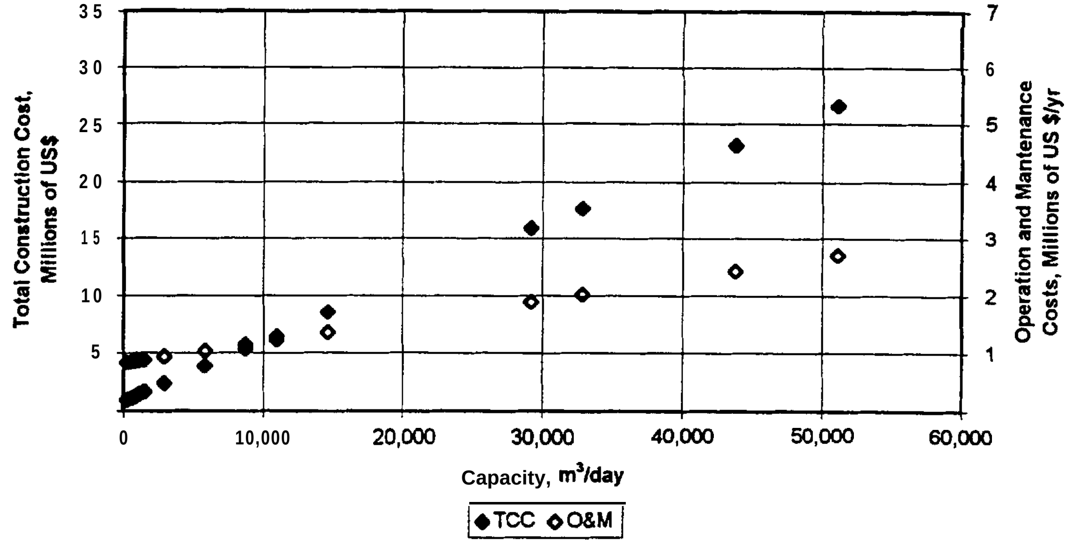


Figure 11.2.—Membrane system costs with increasing capacity. Assumed TDS is 1,800 mg/L, blending allowed, target concentration is 500 mg/L.

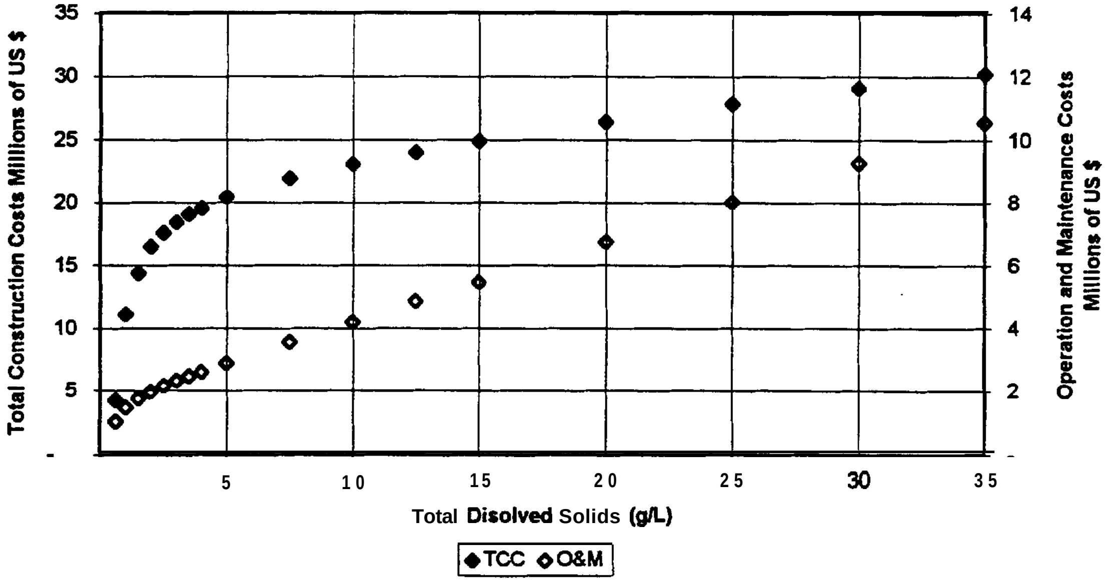


Figure 11.3.—Membrane system costs with increasing TDS. Capacity is 40,000 m^3/day, blending is used to the extent possible, target concentration is 500 mg/L.

Table 11.4.—Assumptions used in comparing membrane treatment costs

Cost item	Assumed value
Total dissolved solids for capacity comparison	1,800 mg/L
Capacity for total dissolved solids comparison	40,000 m^3/day
Target dissolved solids	500 mg/L
Average molecular mass	38.8 g/mole
Chemical addition and pretreatment	3.3 mg/L antiscalant, cartridge filtration,
Post treatment	Aeration, 3 mg/L hypochlorite
Recovery rate	85%
Blending?	Yes, if possible to achieve target quality
Membrane average rejection	99.1%
Membrane productivity	20 m^3/day for 20.3 cm diameter module
Membrane life expectancy	3 years
System life expectancy	40 years
Interest rate	7 %

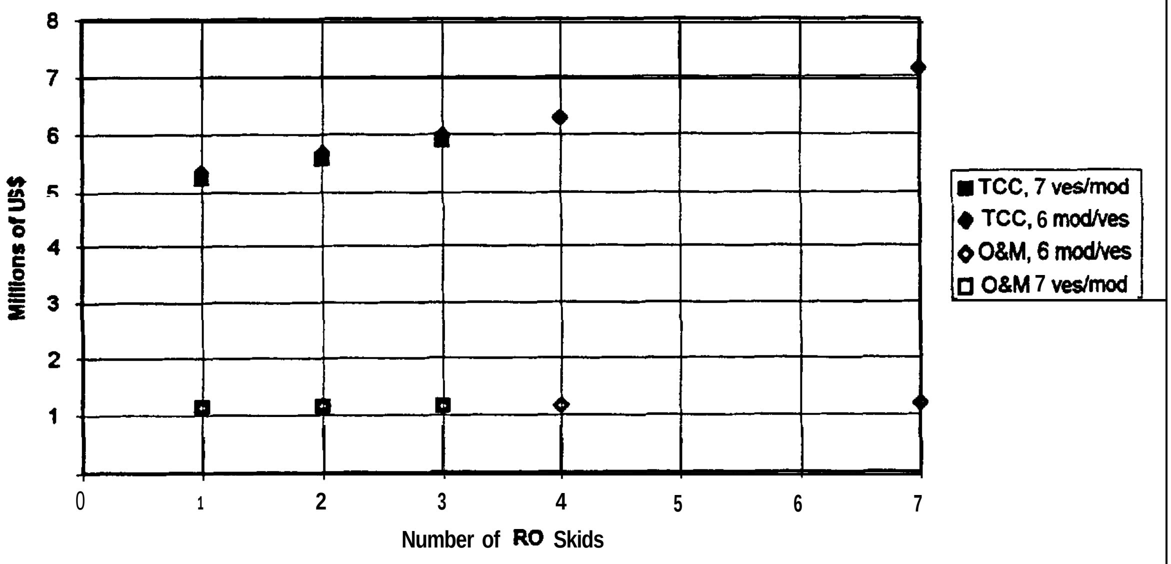


Figure 11.4.—Effect of skid size on construction and O&M costs for a system to produce 1,200 m^3/day (3.17 Mgal/d). Assumptions are the same as those listed in table 11.4.

11.4 Ultrafiltration Costs

The model presented above could be used to estimate UF costs, with a few modifications. Unfortunately, because of the wide variety of matter retained by UF membrane, pilot studies are necessary to estimate UF performance.

- Use retention rate instead of rejection rate. Manufacturers list nominal MWCOs which are the size of particles retained at the 90 percent level. Pilot studies are necessary to determine the actual retention rate of the substances of interest at a particular site.

- UF systems are often operated in a recycle mode. To estimate the number of modules needed, the on site productivity, at the optimum recycle rate, needs to be determined through pilot testing.

- UF systems are operated at much lower pressures than RO and NF. Therefore, the pumping costs would be driven by the piping configuration more than the system operating pressure.

- Pretreatment is not necessary with UF systems. Chemical feed systems would be unnecessary.

- Cartridge filters are not necessary. An in-line screen filter would be sufficient.

11.5 Microfiltration Costs

Cost estimating for MF systems is as particular as UF systems. Cost is still dependent on the membrane area required. While MF membrane has high permeability, it can be fouled quickly. The pilot study must focus on backwash duration and frequency. If the water requires further processing, a portion of the backwash stream may be recycled to the feed. However, if the water is to be for potable uses, it is best to discard the backwash to prevent build up of pathogens in the system. Other differences between MF and the RO model:

- Rejection or retention is not considered in the MF cost model. A membrane must be tested to ensure that it will perform adequately; the flow rate, backwash duration, and frequency are determined so as to best keep the membranes working.

- There is no blending with MF. All water must be filtered.

- Operating pressure is minimal. Pump requirements are determined by the skid capacity.

- No pretreatment is necessary, though coagulants can be added to enhance filtration.

- Cartridge filters are unnecessary.
- Waste stream is minimal; may be a sludge with some types of MF or some kinds of feedwater.

11.6 Conclusions

It is possible to estimate RO and NF membrane system costs from the feed water quality and the capacity. The model presented here includes default values for all other parameters. The model is very sensitive to unit size, as well as several other parameters used in the example. Accurate information on all the variables, but especially rejection, productivity and energy cost, will produce more accurate results. UF and MF costs can be estimated from the same model but require pilot testing to determine the cost parameters of membrane area and pumping requirements.

Chapter 12

PERSPECTIVES ON FUTURE OPPORTUNITIES

The water treatment industry has found increasing value in membrane processes during the last ten years. A snapshot from today's technology landscape presents some perspectives on infrastructure and technology developments that will possibly contribute to future growth. However, in a field as dynamic as membrane separation processes, it is likely that both users and technology developers will create innovation far beyond our current forecast horizon.

1. **Expanded education of users to better understand the differences between the membrane separation processes.**

The trend in water treatment applications is for combinations of membranes to be used in one system. For instance, MF or UF may be used as pretreatment for RO or NF membrane systems. As a result, it is very important that users understand the differences between these types of separations. It is very easy for operators to assume that what is good for RO is also good for UF systems. After all, the equipment looks the same! In actuality, the two processes involve very different separation mechanisms and require different O&M procedures.

2. **Standard methods of membrane characterization and testing must be developed to enable comparison of different membrane formulations and configurations.**

Membranes are complex; researchers of the desalting community are still trying to figure out exactly how they interact with feed water. However, a general understanding is not going to happen if manufacturers continue to describe membrane characteristics in their own separate ways. Users should not have to go through complex analyses to come up with equivalent operating conditions in order to compare membranes from different manufacturers. There needs to be some standardization within the industry. Sea water membranes, at least, are all tested on 35,000 mg/L simulated sea water, but they are tested at different pressures, production rates, and recovery rates. With brackish water membranes, the situation is even worse. Some are tested with 2,000 mg/L NaCl, while others are tested on "San Diego tap water." The feed solution should be adequate to test the general performance of the membrane as it is intended to be used. For instance, brackish water membranes could be tested with an industry standardized 5,000 mg/L NaCl solution, NF membranes could be tested with a combination of mono- and multi-valent ions, UF membranes could be tested with some predetermined macromolecule mixture (polyethylene glycol, for instance), and MF membranes could be tested with a specific type and loading of suspended solids.

At a minimum, procedures should specify testing pressures and feed crossflow velocity sufficient to produce a standard product flux, which should be consistent within a type of filtration. It also would be very useful to measure what the membrane's performance would be under different conditions. Detailed standard testing methods would facilitate preliminary application modeling and help end users narrow their field of potential candidate membranes much more quickly. A variety of groups are currently working on this issue, and we can be optimistic for progress in the future.

3. **Methods should be established to describe water composition, product water quality goals, and desired production capacity such that a few good membranes can be identified for the job from improved manufacturers specifications.**

Current practice in membrane selection is generally driven by the manufacturer's reputation and the cost. Since we know little about the membranes (see 2 above), we need to depend on the supplier for expert advice, support, and fair pricing. Hopefully, the supplier has experience from many pilot tests and knows which membrane will work best, or at least knows what pretreatment will be needed to make the system work with specific membranes. However, if the membrane/feed water interactions were understood well enough to select membranes compatible with the feed water, the pretreatment would not have to be quite as extensive. Though groundwater and surface waters are complex, it is possible that they fall into a finite set of categories. If these categories were described adequately and the membrane/water interactions were described in similar terms, it might be possible to tailor membranes specifically for major water types. Research addressing these goals is currently being undertaken in a number of academic, government, and industrial laboratories.

4. **There should be a way to predict membrane performance under any water condition if the membranes and the source water are well described.**

Accomplishing this final task would decrease the initial development costs for implementation of membrane processes.

5. **Correlations need to be developed among process monitoring data, membrane integrity, and the likelihood of biological contamination of the product water.**

One of the benefits of using membrane-based water treatment processes is that they provide a physical barrier to biological contamination. However, it is often very difficult to determine when that barrier has been compromised. The product water is so clean, particularly with UF, NF, and RO, that particle counts, turbidity, and other monitoring parameters give little indication of impending problems. Integrity testing studies currently are underway, and the ASTM has established a committee to develop consensus standard methods for RO and NF membrane integrity tests. Hopefully, the problem will be resolved within the next 10 years.

6. **Cost models need to be developed relating production capacity, water quality, membrane area, and power requirements for MF and UF to assist planners in estimating preliminary costs without pilot study data.**

Cost models for RO and NF are well developed, but cost models have not been developed for MF and UF. Recirculation rates, backwash duration and frequency, and recovery rates will need to be included as variables that affect the membrane area and power requirements.

12.1 Microfiltration

The potential for MF for removing suspended solids and biological contamination in municipal water treatment systems is great. There are many promising studies ongoing at this time. Recently published technical papers (as cited in chapter 2) show a possible economic advantage of MF over strictly conventional treatment and pretreatment systems for the following reasons:

- MF equipment takes up less space than conventional filtration or coagulation/ clarification equipment.

- Product water quality is independent of feed water quality.

- In some cases, chemical conditioning can be eliminated from the treatment train altogether.

- When MF alone is insufficient, it can be incorporated into conventional coagulation or biological processes to eliminate the large clarification basins or media filtration beds that are normally used.

12.2 Ultrafiltration

Tight UF can remove viruses, pesticides, and large organic molecules that cause color and odor problems and form DBPs (disinfection by-products). Since the use of UF and MF for water is still fairly new, there are only rough guidelines as to when MF should be used and when UF would be more effective. Current research results indicate that the membrane material and water chemistry interact in a complicated way that does not lend itself to simple rules of thumb. As case study documentation grows, it may be possible to establish more concrete rationale for choice.

Unlike MF, there have not been sufficient studies to say for sure that UF is a more economical treatment than the conventional processes of oxidation and media filtration. With UF, the choice is driven by increased awareness of water quality and the inability to achieve treatment goals with conventional processes. The many brands of UF membranes are more diverse than MF membranes, and characterization is even more difficult and critical to successful piloting. The case studies covered in chapter 3 demonstrated that the appropriate choice of membrane can determine the success of a study. In the future, manufacturers and suppliers may be able to provide enough information on the membranes for an intelligent choice to be made.

Fundamental understanding tells us that if UF membranes could be made more porous and with a more uniform pore size distribution, the task of classifying their solute removal characteristics, and therefore correctly choosing which membrane to use, would be easier. Advances in membrane materials development along these lines will likely increase the opportunities for UF in municipal water production.

12.3 NF and RO

NF and RO can be designed to accommodate a variety of treatment objectives when excessive dissolved salts are present. Either membrane type may be more appropriate, depending on the water composition and the concentrate disposal options available. Consequently, the concentrate disposal options should be examined at the beginning of the planning process before selection of the treatment method.

More cost effective and beneficial uses for NF and RO concentrate would increase the potential applications and have a large economic impact on membrane water treatment costs. There are currently four good long-term alternatives for concentrate use or disposal, as follows:

1) Mix it with a low TDS water or wastewater and disperse it into the ocean or other body of water with a similar TDS.

2) Inject it into a confined aquifer that already has a high salt concentration.

3) Evaporate it completely in evaporation ponds or brine concentrators to recover and sell the dried salt as a product, or dispose of it in a lined landfill.

4) Evaporate it partially and use it in a solar pond, providing an interim concentrate storage which produces process heat for additional desalting and brine concentration. Dispose of the dried salt in a landfill.

Improved robustness of membranes, modules, and systems would decrease maintenance costs for municipal water treatment systems. Improvements that are generally recognized and subject to current research and development are:

- Enhanced turbulence and transport of fouling matter out of the system so that excursions in the pretreatment system can be more easily tolerated.

- Increased chlorine resistance of thin-film composite membranes so that excursions in the dechlorination system can be tolerated.

- Inclusion of an on-line normalized performance trending display with instrumentation so that operators can tell at a glance whether performance is changing and take corrective action before irreversible problems occur.

Research is also needed into the source of counter-productive membrane/feed water interactions. Flux decline (especially the irreversible, or hard-to-clean kind) due to solute-membrane adsorption can be reduced when we have a broader range of membrane surface material choices and a more complete database for quantifying solute-material interactions.

12.4 The Next Generation

What will the next generation of membranes be like? Hopefully, they will represent such a paradigm shift that any predictions are woefully inadequate. Current research and development described in the public literature focuses on the following attributes:

- Highly controlled surface energy, pore size distribution, and solute transport coefficients
- High permeation rates
- Low fouling potential under intended operation conditions
- High resistance to any thermal or chemical degradation

The module advances include:

- Complete mixing within the feed flow channel
- Leak-proof seals
- Built-in sensors to monitor individual module performance

Advanced materials processing techniques could include tuning membranes for a particular water source through post-production membrane surface treatments. But implementation of this still requires the advances in describing and matching water composition and membrane characteristics mentioned earlier. Developments in facilitated transport are finally producing commercial applications. If this trend continues, and economical facilitated transport membranes become widely available, economical removal of specific undesirable components in individual processing steps can be realized.

Finally (and certainly more under our control), the third version of this manual may be published on the Internet, with color pictures and some real-time data.

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APPENDIX A
Manufacturer Applications
Source: Dupont and Fluid Systems, U.S.A.

Location	Capacity (1,000 GPD)	Capacity M^3/D	Application	startup
Arizona	680	2,574	boilerfeed	87
California	3,600	13,626	boilerfeed	87
California	100	379	boilerfeed	86
California	274	1,037	boilerfeed	86
California	18	68	boilerfeed	
Canada	400	1,514	boilerfeed	78
Canary Islands	26	98	boilerfeed	81
Canary Islands	26	98	boilerfeed	81
China	900	3,407	boilerfeed	79
India	1,000	3,785	boilerfeed	90
Iraq	17,000	64,345	boilerfeed	86
Iraq	1,300	4,921	boilerfeed	83
Iraq	1,200	4,542	boilerfeed	81
Mexico	100	379	boilerfeed	83
Mexico	158	598	boilerfeed	82
Morocco	400	1,514	boilerfeed	78
New Mexico	432	1,635	boilerfeed	
Pennsylvania	360	1,363	boilerfeed	76
Pensylvania	30	114	boilerfeed	
Singapore	400	1,514	boilerfeed	78
Spain	4	15	boilerfeed	86
Spain	48	182	boilerfeed	85
Spain	48	182	boilerfeed	85
Spain	8	30	boilerfeed	84
Spain	13	49	boilerfeed	84
Spain	26	98	boilerfeed	84
Spain	5	19	boilerfeed	80
Texas	500	1,893	boilerfeed	86
Texas	110	416	boilerfeed	
Texas	20	76	boilerfeed	
Tunis	254	961	boilerfeed	87
Tunis	254	961	boilerfeed	87
Utah	410	1,552	boilerfeed	87
Texas	144	545	boilerfeed	
TOTAL	30548	114,489	**BOILERFEED**	
Mexico	61	231	commercial - soft drink	
Mexico	60	227	commercial - soft drink	
Pennsylvania	272	1,030	commercial - textile waste reuse	
California	30	114	commercial - ultra pure	
West Virginia	60	227	commercial-acid mine drainage	
China	63	238	commercial-beverage	87
China	127	481	commercial-beverage	86
China	127	481	commercial-beverage	81
Maryland	24	91	commercial-beverage	86
Spain	190	719	commercial-beverage	87

Location	Capacity (1,000 GPD)	Capacity M^3/D	Application	startup
Japan	16	61	commercial-food	82
Florida	36	136	commercial-medical	85
Georgia	10	38	commercial-medical	85
Michigan	120	454	commercial-medical makeup	
South Carolina	66	250	commercial-pharmaceutical	85
Massachusettes	72	273	commercial-precious metal plating rinse	
California	15	57	commercial-process rinse water	
France	300	1,136	commercial-ultrapure	85
Ohio	200	757	commercial-ultrapure	85
Spain	7	26	commercial-ultrapure	86
Spain	4	15	commercial-ultrapure	80
Switzerland	320	1,211	commercial-ultrapure	77/87
Total	2,180	8,251	COMMERCIAL	
Arizona	320	1,211	electronic rinse	
California	50	189	electronic rinse	
California	30	114	electronic rinse	
California	30	114	electronic rinse	
Florida	260	984	electronic rinse	
California	300	1,136	electronic rinse	
San Salvador	50	189	electronic rinse water	
Texas	320	1,211	electronic rinse water	
Arizona	251	950	electronics	
Arizona	1,420	5,375	electronics	
Arizona	180	681	electronics	
Arizona	61	231	electronics	
Austria	50	189	electronics	87
Bulgaria	116	439	electronics	86
California	100	379	electronics	86
California	30	114	electronics	85
California	50	189	electronics	84
California	201	761	electronics	
California	201	761	electronics	
California	200	757	electronics	
California	42	159	electronics	
California	251	950	electronics	
California	100	379	electronics	
California	634	2,400	electronics	
California	124	469	electronics	
California	750	2,839	electronics	
California	15	57	electronics	
California	2,016	7,631	electronics	
California	151	572	electronics	
California	150	568	electronics	
California	61	231	electronics	
California	100	379	electronics	
California	450	1,703	electronics	
China	123	466	electronics	86
Colorado	600	2,271	electronics	
Erie	37	140	electronics	79
France	300	1,136	electronics	77
G.D.R.	I.000	3,785	electronics	82

Location	Capacity (1,000 GPD)	Capacity M^3/D	Application	startup
Germany	19	72	electronics	86
Idaho	750	2,839	electronics	
Idaho	71	269	electronics	
Italy	77	291	electronics	**85**
Maine	87	329	electronics	
Massachusettes	201	761	electronics	
Mexico	45	170	electronics	
New York	145	549	electronics	
New York	79	299	electronics	
New York	500	1,893	electronics	
Oklahoma	600	**2,271**	electronics	
Oregon	201	761	electronics	
Pennsylvania	34	129	electronics	
Poland	395	1,495	electronics	73
Spain	20	76	electronics	80
Texas	1,000	3,785	electronics	
Texas	**576**	2,180	electronics	
Texas	600	2,271	electronics	
Texas	58	220	electronics	
Texas	1,100	4,164	electronics	
Texas	341	1,291	electronics	
Texas	115	435	electronics	
Texas	1,200	4,542	electronics	
Utah	**885**	3,350	electronics	
Vermont	481	1,821	electronics	
Virginia	600	2,271	electronics	80
Maine	120	454	electronics - ultrapure	
TOTAL	21,424	81,090	ELECTRONICS	
Argentiana	100	379	industrial	83
Bahrain	600	**2,271**	industrial	79
California	4	15	industrial	84
California	100	379	industrial	83
Chile	160	606	industrial	86
Chilie	50	189	industrial	77
F.D.R.	4,640	17,562	industrial	77
Jordan	800	3,028	industrial	82
Mexico	80	303	industrial	83
Mexico	60	227	industrial	82
Mexico	140	530	industrial	82
Ohio	120	454	industrial	87
Peru	50	189	industrial	82
Peru	50	189	industrial	78
Saudi Arabia	1,044	3,952	industrial	86
Saudi Arabia	84	318	industrial	84
Saudi Arabia	1,500	5,678	industrial	82
U.S.S.R.	3,300	12,491	industrial	80
Venezuela	500	1,893	industrial	**85**
w. Germany	70	**265**	industrial	86
Conneticut	820	3,104	industrial-metal finishing waste	
Wisconsin	60	227	industrial-process	
Mexico	200	**757**	industrial-process	82

location	Capacity (1,000 GPD)	Capacity M^3/D	Application	startup
Mexico	300	1,136	industrial-process	77
Netherlands	300	1,136	industrial-process	75
New York	1,030	3,899	industrial-process	86
W. Germany	63	238	industrial-process	86
Algeria	1,200	4,542	industrial-ultrapure	79
Saudi Arabia	3,500	13.248	industrial/potable	85
TOTAL	20,925	79,201	INDUSTRIAL	
Kuwait	4	15	irrigation	
Saudi Arabia	134	507	irrigation	84
California	50	189	irrigation of seedlings	
Arizona	73,100	276,684	irrigation/waste treatment	
TOTAL	73,288	277,395	IRRIGATION	
California	1,300	4,921	potable-brackish	93
California	900	3,407	potable-brackish	88
California	500	1,893	potable-brackish	90
California	500	1,893	potable-brackish	89
California	250	946	potable-brackish	90
Canada	700	2,650	potable-brackish	88
Florida	4,000	15,140	potable-brackish	89/90
Florida	1,800	6,813	potable-brackish	89
Florida	1,500	5,678	potable-brackish	93
Florida	1,500	5,678	potable-brackish	93
Florida	864	3,270	potable-brackish	89
Florida	576	2,180	potable-brackish	90
Hawaii	500	1,893	potable-brackish	91
Ireland	1,000	3,785	potable-brackish	93
Massuchettes	1,152	4,360	potable-brackish	89
Minnesota	288	1,090	potable-brackish	87/88
New York	422	1,597	potable-brackish	88
New York	360	1,363	potable-brackish	88
North Carolina	3,000	11,355	potable-brackish	89
Ohio	432	1,635	potable-brackish	89
Spain	300	1,136	potable-brackish	89
Spain	240	908	potable-brackish	90
Tailand	300	1,136	potable-brackish	89
Texas	800	3,028	potable-brackish	93
Texas	670	2,536	potable-brackish	88
Texas	576	2,180	potable-brackish	87
Utah	800	3,028	potable-brackish	88
Utah	260	984	potable-brackish	88
TOTAL	25,490	96,480	POTABLE-BRACKISH	
California	1,300	4,921	potable-ground water	
California	500	1,893	potable-ground water	
Florida	12,000	45,420	potable-ground water	
Florida	5,000	18,925	potable-ground water	
Florida	3,000	11,355	potable-ground water	
Florida	2,000	7,570	potable-ground water	
Florida	2,000	7,570	potable-ground water	
Florida	1,800	6,813	potable-ground water	

location	Capacity (1,000 GPD)	Capacity M^3/D	Application	startup
Florida	1,500	5,678	potable-ground water	
Florida	1,500	5,678	potable-ground water	
North Carolina	3,000	11,355	potable-ground water	
South Carolina	600	2,271	potable-ground water	
TOTAL	34,200	129,447	POTABLE-GROUNDWATER	
Algeria	900	3,407	potable	83
Bahamas	7	26	potable	
Bahrain	12,000	45,420	potable	90
Bahrain	12,000	45,420	potable	84
Balearic Island	132	500	potable	85
Canary Islands	925	3,501	potable	89
Canary Islands	264	999	potable	89
Canary Islands	662	2,506	potable	87
Canary Islands	80	303	potable	87
Canary Islands	32	121	potable	87
Canary Islands	1,325	5,015	potable	86
Canary Islands	132	500	potable	84
Canary Islands	160	606	potable	84
Canary Islands	132	500	potable	83
Cape Verde Island	150	568	potable	83
Caribbean	150	568	potable	85/86
Caribbean	400	1,514	potable	90
Caribbean	130	492	potable	89
Caribbean	50	189	potable	87
Caribbean	400	1,514	potable	86
Caribbean	100	379	potable	86
Caribbean	100	379	potable	85
Equador	10	38	potable	
Florida	300	1,136	potable	86
Florida	4,500	17,033	potable	82
Florida	300	1,136	potable	80
Florida	2,000	7,570	potable	79
Florida	600	2,271	potable	74
Florida	1,000	3,785	potable	74
Florida	100	379	potable	
Florida	110	416	potable	
Florida	6	23	potable	
Florida	1,000	3,785	potable	
Florida	10	38	potable	
Florida	930	3,520	potable	
Florida	10	38	potable	
Florida	72	273	potable	
Greek Island	80	303	potable	90
Greek Island	320	1,211	potable	89
Greek Island	320	1,211	potable	89
Greek Island	132	500	potable	83
Greek Island	132	500	potable	81
Iran	1,400	5,299	potable	76
Italy	135	511	potable	85/86
Italy	340	1,287	potable	85
Japan	79	299	potable	81

Location	Capacity (1,000 GPD)	Capacity M^3/D	Application	startup
Jordan	150	568	potable	83
Kuwait	264	999	potable	84
Libya	130	492	potable	86
Libya	50	189	potable	81
Madieira Island	132	500	potable	80
Malta	5,600	21,196	potable	90/91
Malta	1,100	4,164	potable	88
Malta	3,700	14,005	potable	86
Malta	1,585	5,999	potable	83
Malta	5,300	20,061	potable	83
Mexico	500	1,893	potable	82
Mexico	50	189	potable	
Mexico	50	189	potable	
Oman	125	473	potable	81
Puerto Rico	150	568	potable	82
Saudi Arabia	1,600	6,056	potable	87
Saudi Arabia	3,600	13,626	potable	87
Saudi Arabia	7,900	29,902	potable	86
Saudi Arabia	150	568	potable	86
Saudi Arabia	4,500	17,033	potable	86
Saudi Arabia	6,900	26,117	potable	86
Saudi Arabia	100	379	potable	85
Saudi Arabia	260	984	potable	85
Saudi Arabia	60	227	potable	85
Saudi Arabia	100	379	potable	84
Saudi Arabia	130	492	potable	84
Saudi Arabia	150	568	potable	84
Saudi Arabia	150	568	potable	84
Saudi Arabia	189	715	potable	84
Saudi Arabia	2,600	9,841	potable	84
Saudi Arabia	5,000	18,925	potable	84
Saudi Arabia	150	568	potable	83
Saudi Arabia	600	2,271	potable	83
Saudi Arabia	600	2,271	potable	83
Saudi Arabia	1,000	3,185	potable	82
Saudi Arabia	1,000	3,785	potable	81
Saudi Arabia	1,300	4,921	potable	81
Saudi Arabia	2,100	7,949	potable	81
Saudi Arabia	1,300	4,921	potable	80
Saudi Arabia	5,300	20,061	potable	79
Saudi Arabia	7,900	29,902	potable	79
Saudi Arabia	10,200	38,607	potable	79
Saudi Arabia	16,000	60,560	potable	78
Saudi Arabia	75	284	potable	76
Saudi Arabia	400	1,514	potable	76
Spain	132	500	potable	82
St. Thomas	39	148	potable	84
U.A.E.	300	1,136	potable	90
U.A.E.	300	1,136	potable	90
U.A.E.	300	1,136	potable	90
U.A.E.	1,200	4,542	potable	90
U.A.E.	1,200	4,542	potable	90

Location	Capacity (1,000 GPD)	Capacity M^3/D	Application	startup
U.A.E.	1,200	4,542	potable	90
U.A.E.	1,200	4,542	potable	90
U.A.E.	2,400	9,084	potable	90
U.A.E.	2,400	9,084	potable	89
U.A.E.	1,200	4,542	potable	88
U.A.E.	1,200	4,542	potable	88
U.A.E.	2,400	9,084	potable	86
Argentina	53	200	potable-sea water	
Australia	106	400	potable-sea water	
Bahamas	30	114	potable-sea water	
Bahamas	100	380	potable-sea water	
California	16	60	potable-sea water	
California	21	80	potable-sea water	
California	581	2,200	potable-sea water	
Canary Islands	251	950	potable-sea water	
Egypt	1,000	3,785	potable-sea water	
Indonesia	**793**	3,000	potable-sea water	
Korea	151	570	potable-sea water	
Kuwait	264	1,000	potable-sea water	
Lybia	2,642	10,000	potable-sea water	
Mexico	300	1,135	potable-sea water	
Mexico	48	180	potable-sea water	
Ras AI **Khaima**	132	500	potable-sea water	
Saudi Arabia	75	284	potable-sea water	
Saudi Arabia	1,321	5,000	potable-sea water	
Saudi Arabia	3,170	12,000	potable-sea water	
Saudi Arabia	317	1,200	potable-sea water	
Saudi Arabia	100	380	potable-sea water	
Spain	1,321	5,000	potable-sea water	
Spain	201	760	potable-sea water	
Tahiti	16	60	potable-sea water	
Turkey	66	250	potable-sea water	
U.S. Virgin Islands	80	303	potable-sea water	
Venezuela	151	570	potable-sea water	
TOTAL	**173,093**	**655,158**	**POTABLE-SEA WATER**	
Shipboard	65	246	potable/ship	86
Shipboard	75	284	potable/ship	86
Shipboard	63	238	potable/ship	85
TOTAL	**203**	768	**POTABLE/SHIP**	
U.S.A	3,000	11,355	potable	80
U.S.A	125	473	potable	85
Utah	30	114	potable	
Algeria	12,200	46,177	potable-process	79
Algeria	300	1,136	potable-process	83
Canary Islands	50	189	potable-process	
Saudi Arabia	2,400	9,084	potable-process	79
Argentina	20	76	potable/arsenic removal	
Chilie	132	500	potable/arsenic removal	
Saudi Arabia	1,200	4,542	potable/industrial	76
Venezuela	1,000	3,785	potable/power	83

Location	Capacity (1,000 GPD)	Capacity M^3/D	Application	Startup
Algeria/Mobile Unit	77	291	potable/process	86
Venezuela	100	379	potable/process	
Bahamas	240	908	potable/resort	81
U.S.A	160	606	potable/resort	86
TOTAL	21,034	79,613	POTABLE / VARIOUS	818
TOTAL POTABLE	2,010	7,606	POTABLE - ALL TYPES	85
California	412	1,559	power	84
Canada	310	1,173	power	84
Texas	970	3,671	power	85
Venezuela	1,000	3,785	power	80
Colorado	432	1,635	power - cooling tower **blowdown**	
New Mexico	2,300	8,706	power plant waste water	
California	6	23	power-gas turbine injection	
California	30	114	power-gas turbine injection	
California	25	95	power-gas turbine injection	
TOTAL	5,485	20.761	POWER	333
California	50	189	waste - sewage reclamation	
California	168	636	waste - sewage reclamation	
California	225	852	waste - sewage reclamation	
California	5,000	18,925	waste-process tertiary effluent	
California	30	114	waste water treatment	
California	750	2,839	waste water treatment	
Colorado	125	473	waste water treatment	
Saudi Arabia	2,600	9,841	waste water treatment	
TOTAL	8,948	33,868	WASTE WATER	
not included in TOTALs:				
U.S. Army	50,198	190,000	potable-sea water	

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The second edition of the membrane manual provides an overview of microfiltration, ultrafiltration, nonfiltration reverse osmosis, and electrodialysis processes as they are used for water treatment. Performance evaluation, cost models, cleaning, and concentrate issues are presented as well as an extensive listing of membrane products currently available. Membrane listing includes physical characteristics, performance data, operational limits, and comparative statistics for each.

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