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THE DESALTING AND WATER TREATMENT MEMBRANE MANUAL:

A GUIDE TO MEMBRANES FOR MUNICIPAL WATER TREATMENT

Water Treatment Technology Program Report No. 1

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16. ABSTRACT

The Bureau of Reclamation prepared this manual to provide an overview of microfiltration, ultrafiltration, nanofiltration, reverse osmosis, and electrodialysis processes as they are used for water treatment. Membrane composition, the chemical processes, and the physical processes involved with each membrane type are described and compared. Because care and maintenance of water treatment membranes are vital to their performance and life expectancy, pretreatment, cleaning, and storage requirements are discussed in some detail. Options for concentrate disposal, also a problematic feature of membrane processes, are discussed. The culmination of this wealth of knowledge is an extensive comparison of water treatment membranes commercially available at this time. The tables cover physical characteristics, performance data, and operational tolerances.

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

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IMPORTANT NOTICES

We hope that you will find this manual a useful document that you refer to on occasion. It is intended as a beginning; feel free to add useful information from other sources. The membrane industry is producing new, improved membranes much faster than we can tell you about them in a formal report. In the future, updates to the manual will be distributed as inserts and page replacements to this publication. To further reduce the amount of paper involved, we hope to have copies available on disk in the future.

KEEP YOUR COPY OF *THE DESALTING AND WATER TREATMENT MEMBRANE MANUAL* UP TO DATE!

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WATER TREATMENT COST ESTIMATION PROGRAM

One important item missing from this manual is cost information. Cost of membrane systems is highly site specific, as any manufacturer will tell you. The Water Treatment Cost Estimation Program is in progress at this time. This program will take your water analysis, the amount of water you have access to, and the amount you need as input. The output will be two or three alternative treatment systems that can be used to bring your water source to EPA Drinking Water Standards, and an estimate of the cost for each system.

Yes, cost estimation programs do exist, but they use an outdated cost basis, are difficult to use, do not consider many water treatment alternatives, do not consider waste disposal options, and/or are expensive. The Water Treatment Cost Estimation Program will consider all pretreatment, membrane processes, and disposal options discussed in this manual. At this point, the program will be packaged as an EXCEL or LOTUS 123 spreadsheet.

Send inquiries to the above address.

SI Metric Conversions

From	To	Multiply by
ft	m	*3.048 000 E - 01
in	m	*2.540 000 E - 02
ft^2	m^2	*9.290 304 E - 02
kgal	m^3	3.785 412
Mgal	m^3	3.785 412 E + 3
acre-ft	m^3	1.233 489 E + 3
lb/in^2	Kpa	6.894 757
°F	°C	$t_{°C} = (t_{°F} - 32)/1.8$

* Exact Conversion

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1. INTRODUCTION

Membrane separation processes have been used for years, but they have not come to play an important role in producing potable water supplies until the past 10 years. RO (reverse osmosis) and ED (electrodialysis) are replacing phase change desalting technologies for supplying water to coastal and island communities all over the world. Nanofiltration is becoming an economical alternative to the traditional water softening processes. As membrane separations become more widely understood, more applications will be found for them. This document will shed some light on these processes: how membranes are made; what makes them different; what membranes are available today and how they can be compared; how they are cleaned and stored; what pretreatment is required; and what is done with the waste.

Traditionally, membrane applications in municipal water treatment have been limited to microfiltration and desalination. The newer membrane types are being developed primarily for high return processes such as chemical separations, enzyme concentration, and beverage purification. These processes are also useful in municipal water treatment, but the membrane systems are still new compared to traditional water treatment processes. The usual reluctance to try anything new and unfamiliar has been encountered. The cost to retrofit water treatment plants may be prohibitive in some cases. However, as the Environmental Protection Agency adds to its list of regulated water contaminants, many communities will find that they need a treatment process that will remove a variety of contaminants. And, as in the computer industry, the economics of membrane separations may improve as the market expands. Table 1.1 summarizes the operational differences between the four major membrane separation processes: microfiltration, ultrafiltration, nanofiltration, and reverse osmosis. Table 1.2 compares substances removed in each process and their traditional counterparts in municipal water treatment systems.

Table 1.1. – Comparison of membrane separations for municipal water treatment.

Separation process	Rating, pore size, or molecular weight cutoff	Operating pressure range (Kpa)	Productivity (L/cm^3*day)
Microfiltration	0.05 – 2.0 μm	140 – 5000	90 – 100% recovery
Ultrafiltration	0.001 - 0.1 μm 10 – 1000 Å, MWCO*(Daltons): 1000 – 500,000	200 – 1000	1 – 5
Nanofiltration	8 – 80 Å MWCO (Daltons): 180 – 10,000	550 – 1380	1 – 6
Reverse Osmosis	1 – 15 Å	1380 – 6890	0.1 – 2.3

* MWCO: Molecular Weight Cut Off is the molecular weight of species rejected by the membrane.

Table 1.2. – Municipal membrane separations and their traditional counterparts.

Membrane separation technology	Substances removed	Comparable traditional water treatment methods
Microfiltration	Bacteria, Virus, Larger Colloids, Separation of Precipitates and Coagulates	Ozonation/UV Chlorination Sand Filters Bio Reactors, Coagulation/Settling Tanks
Ultrafiltration	High MW Proteins, Organics, Pyrogens	Sand Filters, Bio Reactors, Activated Carbon
Nanofiltration	Divalent Ions, Larger Mono-valent ions, Color and Odor	Lime/Soda Softening, Ion Exchange
Reverse Osmosis	All of the above and monovalent ions	Evaporation, Freezing, Electrodialysis

2. MICROFILTRATION

2.1 Membrane Production Processes – Symmetric Membranes

Membranes used for microfiltration range from string wound depth cartridges to polymer films to "high-tech" titanium oxide tubes coated inside with zirconium. All the types of membranes used in ultrafiltration, nanofiltration, and reverse osmosis also are represented in microfiltration. Therefore, this chapter will go into the details of how the different membranes are made. One basic distinction that is made between membranes is in their symmetry. Symmetry, in part, determines the characteristics of the membrane. To develop membranes with improved flux, or water passage, fouling resistance, and rejection, the effects of different production processes on the membrane symmetry must be understood.

Symmetric membranes are of uniform composition; the two sides of the membrane have equal characteristics. The two primary methods for producing symmetrical membranes are solution casting and melt pressing.

2.1.1 *Solution Casting.* – In solution casting, the polymer is dissolved in a highly volatile solvent at a rate of 15 to 20 percent polymer. The solution is drawn with an adjustable casting knife over a plate of glass or other suitable smooth surface. As the solvent evaporates, the polymer is precipitated, forming a thin film.

The choice of solvent is critical to the production of a uniform polymer film. The solvent must have a boiling point low enough to evaporate at room temperature before the polymer has a chance to absorb moisture from the atmosphere. Additional water in the casting solution will be incorporated into the polymer film, producing an inferior product. If the desired polymer does not dissolve in a solvent with a very low boiling point, a variation on the above process called spin casting, may work.

With spin casting, the solution is spread by centrifugal force on the inside of a cylinder. As the surface thickens, centrifugal force pushes the higher density solution back against the cylinder, bringing fresh solution to the surface. This cycling of the casting solution prevents a skin from forming on top of the film and produces a uniform cell structure throughout the depth of the film (Strathmann, 1990).

2.1.2 *Melt Pressing.* – Melt pressing does not involve solvents. The dry, powdered polymer is pressed between sheets of teflon coated foil or cellophane with heated plates at pressures of 13,800 to 35,000 kPa. The temperature should be just high enough to fuse the polymer into a film. Shims or forms are used to control the thickness of the membrane. Porosity depends on the particle size of the powdered polymer or other material used (Strathmann, 1990).

2.1.3 *Pore Formation.* – Microporous membranes can be formed by irradiating symmetrical membranes with charged particles from a nuclear reactor. The particles pass through the membrane, leaving regularly sized sensitized tracks. The membrane is then passed through an etching bath that dissolves damaged membrane material. The sensitized tracks of the particles are dissolved, producing the pores. The longer the membrane is in the etching bath, the larger the pore diameter will be. Pores produced in this manner are called capillary pores. Nuclepore polycarbonate microfilters are the only commercial microfiltration membranes made by this process.

Variation in capillary pore size is minimal because of the uniform size of the charged particles; however, spacing is random. The longer the membrane has been irradiated, the greater the number of pores, and the greater the probability of pore overlap.

Stretching is another method used to create pores. Partially crystallized polymer membranes are stretched perpendicular to the direction of extrusion. Pores created in this way are linear lesions in the polymer matrix held together by fine strands. Overall pore distribution is much more regular than with irradiation, which means that greater porosities are possible without increasing the variation in pore diameter. At low porosities however, pores produced by stretching exhibit a higher level of variability than capillary pores.

One last method of pore formation is template leaching. If a soluble component is mixed with the casting/polymer solution, the component can be leached out from the finished membrane, leaving pores behind. The size of the pores and their distribution depends on the leached component and the level of mixing of the soluble component in the polymer solution (Porter, 1990).

2.2 Membrane Production Processes – Asymmetric Membranes

Asymmetric membranes have a change in pore structure with depth. Typically, a thin permselective layer forms on the top of the film that is underlain by an open porous support layer. The dense layer at the top face of the membrane increases rejection rates, and the porous support layer increases water flux rates. This type of membrane is produced through thermal phase inversion. In this process, differential polymer precipitation is caused by a rapid change in temperature at the film surface, or by exposure of the polymer to a precipitant.

2.2.1 *Thermal Phase Inversion Membranes.* – Phase inversion is a process whereby the polymer is precipitated in phases so that it is present both as a solid and a liquid. Membranes made through phase inversion may be symmetric or asymmetric, depending on the exact procedure used.

One way to induce phase inversion with a polymer that is insoluble in a particular solvent at room temperature is to dissolve it in hot solvent. The solution is cast as described above. As the solvent cools, the polymer develops a spongy uniform cell structure. The size of the cells depends on the cooling time. Longer cooling times produce a finer cell structure. As the polymer precipitates, the surface of the membrane is partially protected from atmospheric vapor by the solvent, producing an even, symmetrical structure.

Asymmetric phase inversion membranes result when the polymer-solvent mixture is exposed to a precipitant, such as water. As the precipitant is absorbed into the film the polymer precipitates at a faster rate than it would if only caused by the evaporation of the solvent.

The precipitant is introduced either as a vapor or as a wash. If a wash is used, the surface of the film quickly precipitates, causing a skin to form. The skin limits the absorption of precipitant into the interior of the film. After this point, precipitation is controlled by the evaporation of solvent through the skin. When a vapor is used to introduce the precipitant, it is absorbed into the interior of the film without skin formation, but cell structure near the surface will be finer than at depth. The structure of the film interior is governed by the concentration of polymer in the casting solution, the solubility of the polymer in the solvent, and the gelation time or precipitation rate. Generally, the longer it takes to precipitate the polymer completely, the finer the cell structure will be.

When a skin is formed on the surface of the membrane and a low concentration of polymer is present in the casting solution (<20% by weight), fingerlike structures tend to form within the film. Because the skin restricts the escape from the surface, the evaporating solvent accumulates in pockets and eventually breaks through the skin. The pocket is lined with a skin of precipitated polymer which further isolates the interstices. There, the polymer precipitates slowly, forming a fine cell sponge texture.

These finger formations, or rather the spaces between them, allow the membrane to have high water flux rates. The tight, thin, dense surface layer gives the membrane high salt rejection rates. Table 2.1 outlines the major factors affecting membrane structure development and their effects. For further information on membrane production see M.C. Porter's *Handbook of Industrial Membrane Technology* (1990).

Table 2.1. – Cause and effect of membrane production.

Factors affecting membrane structure	Effect
Solubility of Polymer in Solvent	Lower Solubility → Faster precipitation → Development of finger structures
Presence of Viscous Additives in the Solvent	High Viscosity → Slows reaction of polymer with precipitant → Finer pore structure
Concentration of Polymer	Higher concentration → Fine, spongelike pore structure
	Low concentration → Open cell structure, Finger structure development
Polymer Used	Affects operating parameters
Precipitant	Lowers solubility of polymer → Faster precipitation → Tends toward finger structures
Method of Precipitant Introduction	Slow introduction → Fine, spongelike pore structure Fast emersion → Skin formation

From Strathmann, H. "Synthetic membranes and their preparation" in *Handbook of Industrial Membrane Technology*, M.C. Porter, ed. 1990.

2.2.2 *Ceramic and Metallic Membranes.* – Ceramic and metallic membranes are used mostly for microfiltration and ultrafiltration. They tend to be highly inert, withstanding high temperatures and pH from 0 to 14. Three production processes are outlined here and to a greater extent in R.W. Baker's chapter on "Membrane and Module Preparation" in *Membrane Separation Systems.*

Leaching is used to produce microfilters with pore sizes from less than 10 µm to 100 µm. A leachable component is mixed with the membrane material (e.g., glass), then dissolved out with acid after the membrane is formed. Pore size is determined by the type and concentration of the leachate. However, high variability results from difficulty in controlling leachate distribution.

The Sol-Gel process is used with alumina, silica, and titanic membranes. This process involves polymerization of a hydrolyzed metal alkoxide dissolved in a water-alcohol solution. The resultant gel is dried and heat treated to form a membrane with average pore diameters of 50 Å or less.

Carbon membranes are made by heating polymeric hollow fibers to 500-800 °C in an inert atmosphere or vacuum. In this environment, the polymer is reduced to carbon with pore diameters between 2 to 5 Å.

2.3 Physical Characteristics

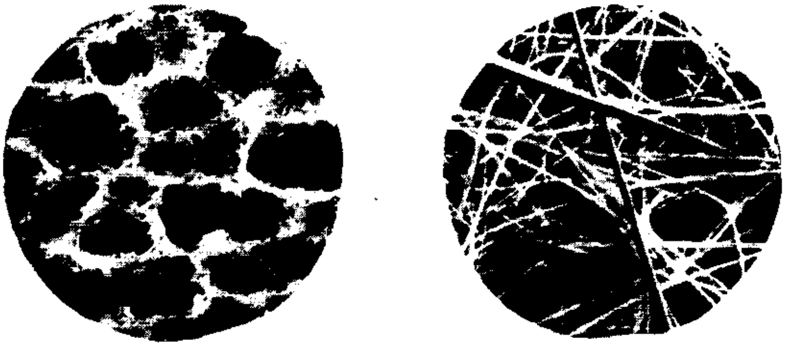
Microfilters have pore sizes between about 0.05 to 2 μm. The methods for creating pores of this range in a membrane are stretching, irradiation and etching, thermal phase inversion, melt pressing, and template leaching. Irradiation and etching produce a clean-edged, distinct pore known as a "capillary" type pore structure. A "tortuous," spongelike, porous structure can be formed through thermal phase inversion or with thermally bonded microfibers. Pores formed by stretching exhibit traits of both capillary and tortuous pore membranes. Figure 2.1 illustrates the differences between these three types. Table 2.2 lists the functional differences between capillary and tortuous pore membranes.

Stretched membranes are similar to capillary pore membranes in that they have distinct and regularly sized pores; they are also much thinner than tortuous pore membranes. They differ in that stretched membranes have regularly spaced pores and irradiated membranes have random pores. With increasing pore size or porosity, capillary pore membranes exhibit increasing occurrences of double and triple pores. This characteristic is shown on figure 2.1.

Asymmetric microfiltration membranes are a cross between microfilters and ultrafilters. The inside of the membrane has an open, tortuous pore structure that gradually closes in to form a tight pore structure on the outside. These membranes are not reversible. If used with the open side upstream, the membrane will become clogged and, because the fine surface layer has no support in that direction, feed water may break through into the product stream.

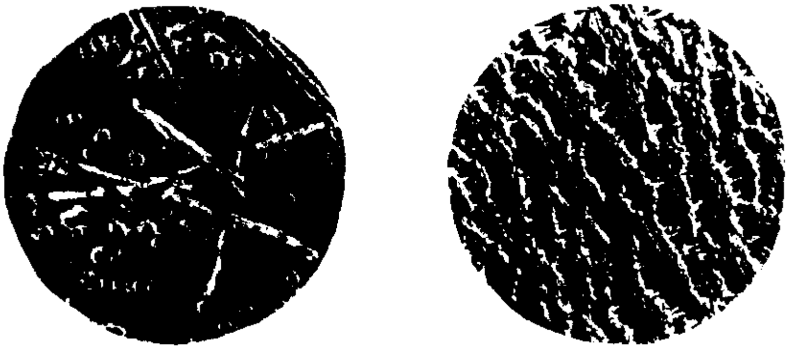
Table 2.2. - Comparison of tortuous and capillary pore membranes assuming equivalent pore sizes.

Tortuous pore membranes	Capillary pore membranes
Regular, spongelike pore structure	Randomly spaced, distinct pores
Membrane thickness ~ 10 μm	Membrane thickness ~150 μm
Porosity – 75%	Porosity – 5%
Capillary pore membranes require lower applied pressure for same flow rate as same pore size tortuous pore membrane	
Bubble point testable	Not bubble point testable
Retain particles smaller than pore size	Do not retain particles smaller than pore size



Thermal Phase Inversion

Thermally Bonded Microfibers



Capillary Pore Membrane

Stretched Membrane Pores

Figure 2.1. – Microfiltration pore types.

Capillary pore membrane photo from H. Stratham, 1990, *Handbook of Industrial Membrane Technology* stretched membrane pore membrane photo from M.C. Porter, 1990, *Handbook of Industrial Membrane Technology* reprinted with permission from Noyes Publications. Thermal bonding and thermal phase inversion from Hytrex Division, Osmonics, Inc. *Pleated Cartridge Filters Catalog*, 1986.

2.4 Pore Size Determination

Pore size determines what particles will be filtered out of feed water. Capillary pores can be measured with a scanning electron microscope, but the size of tortuous pores is not so apparent. The test used to determine the maximum pore size is called the bubble point test. The bubble point of a membrane is the pressure required to overcome capillary forces holding a wetting liquid within the pores and let a bubble of air escape. The bubble point is inversely proportional to the maximum pore size, which is why capillary pore membranes cannot be bubble point tested accurately. The bubble point would correspond to a double or triple pore even if only one pore was present.

Another method for determining pore size is a challenge test. The membrane is challenged with a solution of a particular organism. The beta ratio is then calculated from the number of organisms in the feed water divided by the number passing through.

As an example, AG Technology Corp. tests their 0.2-µm microfilter using *Pseudomonas dimunitia* and *E. coli* at challenge levels of 2.5×10^7 and 6.0×10^9, respectively. At these levels, the organisms were undetectable in the permeate, resulting in an LRV (Log Reduction Value) of 7 and 9 for those organisms (AGT Technical Bulletin, 1992). The LRV is the log of the beta ratio assuming an insignificant positive number of organisms did pass through the membrane.

The LRV is only an indication of the variation in pore size, and should not be confused with other parameters using micro-organisms. A titer reduction level is similar to the LRV except that the total number of organisms given is per cartridge rather than per milliliter of feed water. For example, Ionpure gives a LRV of >8 with *Pseudomonas diminuta* for their 0.2-µm PV filter cartridge and a titer reduction of $>10^{12}$ of the same micro-organism per cartridge. When comparing either of these parameters, be sure the terms and organisms are the same.

Although the LRV method is a good indication of effectiveness against organisms in the real world, it is not an accurate measure of the pore size of the membrane. Any organism will vary in size within a given population. Consumers have no way to tell if the filters have been tested with "fat" microbes or not. This problem is of concern to manufacturers as well; they would not want their filters to be tested with "thin" microbes. To minimize the size diversity problem, some manufacturers use monodisperse latex spheres of known diameter to pinpoint the pore size of their products.

2.5 Configurations

Microfilters come in four different configurations: plate and frame; pleated cartridge modules; spiral wound modules; and bundled tubular membranes. Figure 2.2 illustrates the differences between them. Plate and frame modules use membrane disks. Feed water flows into the pressure vessel and through several layers of membranes separated by spacers and isolated from each other by watertight seals. The product may be the filtrate, as in water purification, or the filter cake (as in cheese production, for instance). The larger particles are collected on the microfilter and disposed of with the filter as solid waste, or harvested. The seals between each membrane chamber must be intact to maintain pressure in the system and to avoid leakage of unfiltered feed water into the product stream.

Microfiltration cartridge modules come in three basic configurations: dead-end filtration, flow-through filtration, and cross-flow filtration cartridges. In dead-end filtration, the pressurized feed stream flows into a dead-end filter receptacle, passes through the filter, then exits from the inside of the filter. All of the water passes though the filter, and the particulates are left in the depths of the filter. For this reason, the microfilters are called depth filters. These tubular filters have walls of varying thickness depending on the effective pore size. Some are actually string wound around a perforated tube. Flow-through filtration is similar except the water flows through the module from the outside of the filter to the inside then exits at the opposite end. The filter itself is usually still a depth filter, and all of the water passes through the filter. Cross-flow filtration has more in common with the other membrane processes than with the two just described. Feed water flows under pressure across the membrane surface. Part of the flow is filtered through the membrane and part remains with the particulate load, exiting through a reject port. The reject can be recycled though the system or discarded.

Pleated membrane cartridges are mostly used in flow-through microfiltration. The cartridge is composed of an interior core support surrounded by pleated polypropylene support layers that enclose and support the membrane. A molded polypropylene protective cage surrounds the whole module. In flow-through filtration mode, feed water enters at one end of the module and flows through the membrane from the outside to the inside. Particulates are retained on or within the membrane. The membrane is disposed of when the flow rate decreases by an amount specified by the manufacturer. Some pleated membrane cartridges can be back-flushed at low pressure, considerably extending their useful life (Porter, 1990). Both sides of the membrane must have adequate support to withstand backflushing. Osmonics' line of pleated cartridge filters may be operated at two-thirds to one half the maximum forward differential pressure under reverse flow, depending on the membrane material (*Osmonics Technical Document*, 1986). With modification, some pleated membrane cartridges can be adapted for use in cross-flow filtration.

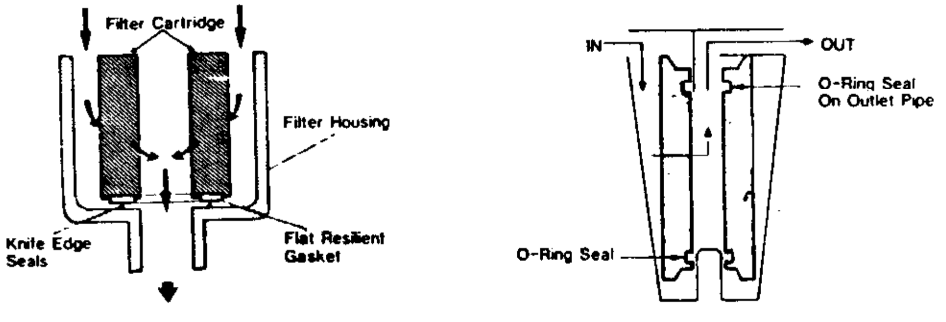
Table 2.3. – Hollow fiber and tubular membrane materials.

Membrane materials	Pore size (µm)	Tubes (I.D. in mm)	Hollow fibers (I.D. in mm)
Alumina	0.2 - 5.0	3 - 15	----
Cellulose Ester	0.2	----	0.37 - 0.61
Polypropylene	0.2	5.5	0.6 - 1.8
Polysulfone	0.1 - 0.4	----	0.5 - 1.0
Polyvinyl Alcohol	0.4	----	0.4
Polyvinylidene Difluoride	0.08	25.4	----

From *Handbook of Industrial Membrane Technology*; p. 113; Porter, 1990.

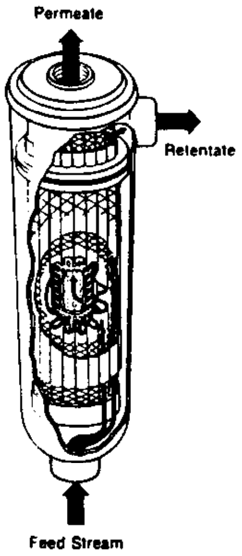
Tubular and hollow fiber membranes are made from a variety of materials. They are designed for cross-flow filtration. Table 2.3 lists some of the materials used and the typical pore sizes and internal diameter for that material. Although the pressure vessels for tubular and hollow fiber membranes look the same as those used for cross-flow pleated cartridges, they operate in the reverse direction. With tubular and hollow fiber membranes, feed water flows from inside the membrane tube or fiber and is filtered to the outside. The port on the side of the module is the permeate port and the reject comes out the end.

(a) Cartridge Microfilters

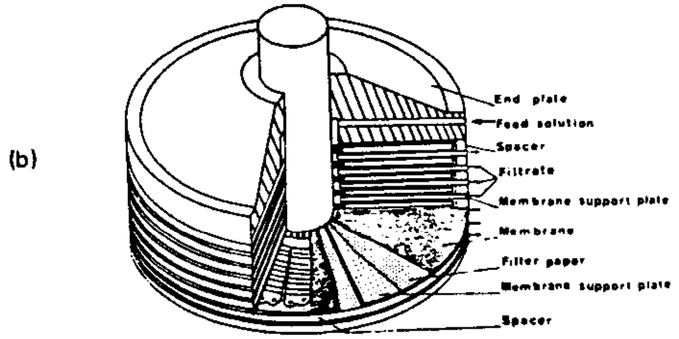


Flow-Through

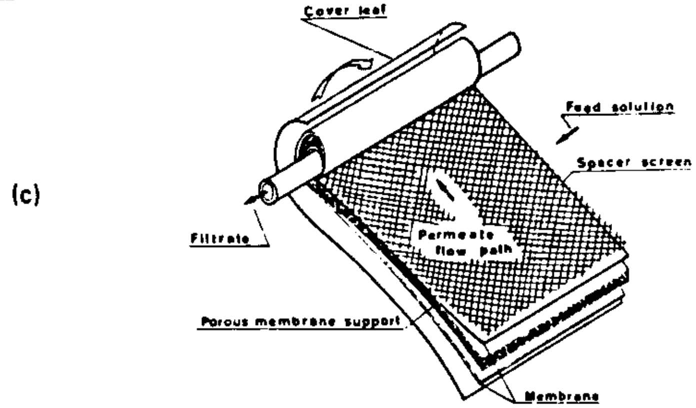
Dead-end



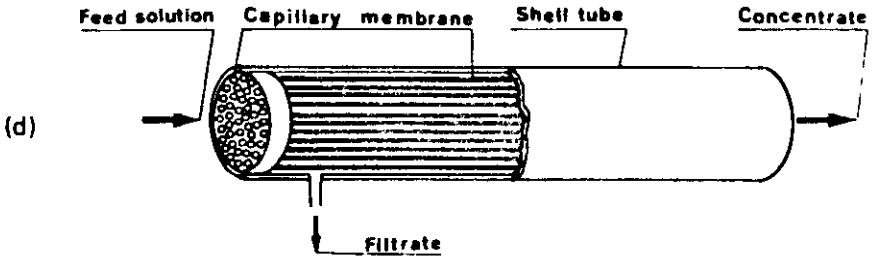
Cross-flow



(b) Plate and Frame Filtration



(c) Spiral Wound Membrane Element



(d) Tubular or Hollow Fiber Module

Figure 2.2. – Configurations normally used in microfiltration.

(From: *Handbook of Industrial Membrane Technology*, pp. 51-52, ed. Porter, 1991, reprinted with permission from Noyes Publications).

Spiral wound elements have four layers glued together and rolled around a tube that collects product water. Two layers of membrane are stacked back to back, forming an envelope with a product water spacer inside. Another spacer, called the feed channel, is placed between each set of membrane envelopes to improve circulation of the feed water to the membranes and to promote turbulence to reduce fouling. As water permeates into the membrane envelope, flow is directed inward toward the permeate tube at the center of the spiral. Feed water is forced through the membrane along the length of the module, and the brine exits through the feed channel on the other end. The product water exits from the center of the module through the permeate port

2.6 Filtration Rate vs. Throughput

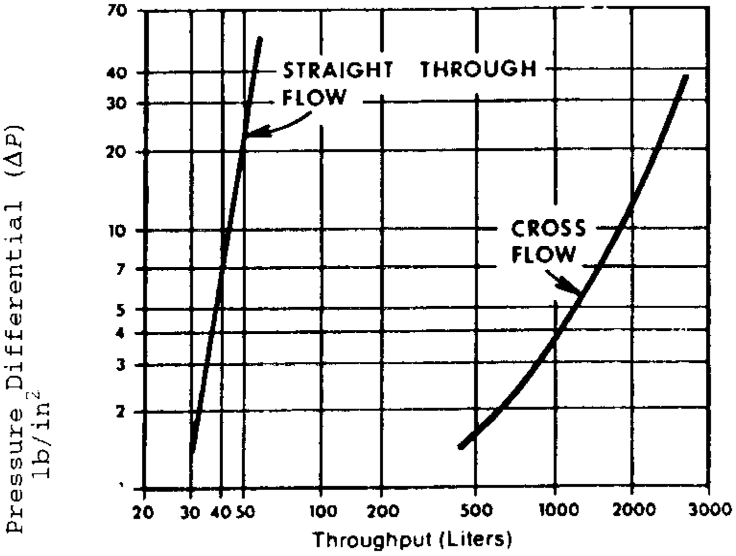
In flow-through microfiltration, water is forced under pressure through the membrane. The rate at which the filtrate exits the membrane module is the filtration rate. The amount of water that passes through the membrane before it plugs up is the throughput. Both depend on the pore size, the amount of suspended solids in the water, and the pressure. To a certain extent, filtration rate is proportional to pressure, but after a certain point, filtration rate begins to decline more rapidly with time than it would at lower operating pressure. The faster the particulate laden water is forced through a membrane, the faster the membrane plugs up. At high pressure, the filter cake is compacted and becomes impervious. At lower pressures, the filter cake may remain soft, allowing water to continue to pass through. So, although the filtration rate may be lower at lower pressure, the filtration time is longer, and therefore, more water can be processed before plugging (Porter, 1990). Figure 2.3 illustrates the increase in throughput with cross-flow microfiltration compared with through flow filtration. In a cross-flow mode of operation, flow across the membrane surface limits the buildup of filter cake. The higher the velocity of the feed water, the less buildup occurs. The residence time of the water in the filter is decreased, though, so some systems use recirculation to achieve a higher production rate (Porter, 1990).

The prevention of plugging with cross-flow filtration greatly increases the useful life of microfilters. The ability to backwash cross-flow microfilters also helps extend their life span. Cross-flow microfiltration is still novel, but experience indicates that with longer lifetimes, microfilters would begin to experience the same fouling and degradation problems as ultrafilters and reverse osmosis membranes.

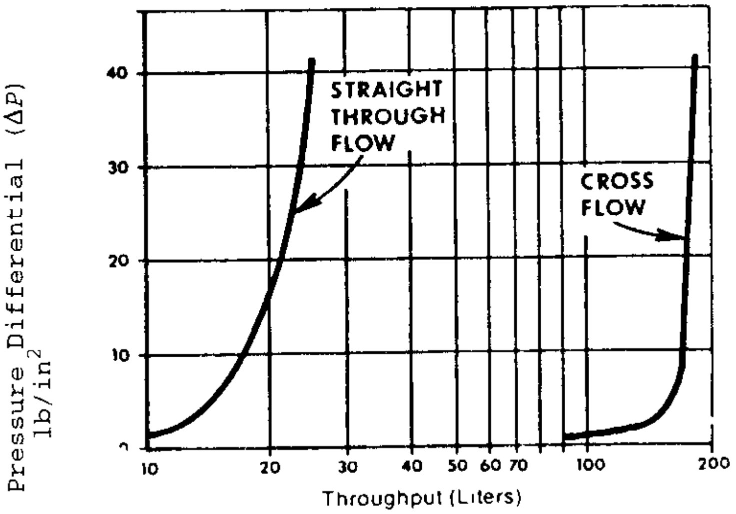
2.7 Applications

Microfiltration is used to remove particles, bacteria, and colloids from feed streams in water treatment systems. Ion exchange resins, ultrafilters, and RO membranes are all susceptible to fouling by micro-organisms and colloids in the feed. Even after ion exchange, feed water to RO membranes must be filtered with a microfilter to remove fine resin particles. Often a series of microfilters with progressively smaller pore sizes will be used in a line to extend the life of the smallest pore size filter. Some experimentation is required to arrive at the optimum pore size combination. For instance, if two sizes are used, they should each remove proportional percentages of the total particulate load over time. Neither should be removing the entire load.

Memtek produces a line of tubular microfilters designed to replace sand filters and settling tanks normally used to separate solids produced in precipitation and flocculation. Water from the flocculation tank is filtered from the solids leaving a waste stream with 2 to 5 percent solids. The waste stream is mixed with inflow to the filters in a concentration tank. Periodically sludge is removed and run through a filter press where it is reduced to 30 to 40 percent solids. These solids can be reclaimed or disposed of as solid waste (*Memtek Bulletin* PB09, 1989).



Increase in throughput for cross-flow microfiltration of 0.5 percent yeast with a modified 0.2-µm capillary-pore membrane pleated cartridge.



Increase in throughput for cross-flow microfiltration of 0.25 percent activated carbon with a modified 0.2-µm capillary-pore membrane pleated cartridge.

(Porter, 1990, reprinted with permission from Noyes Publications.)

Figure 2.3. – Comparison between cross-flow and through-flow microfiltration.

3. ULTRAFILTRATION

Ultrafiltration is a step smaller than microfiltration, for particles larger than 0.04 µm or with a molecular weight larger than 10,000 Daltons. The membranes used for ultrafiltration are anisotropic, or asymmetric. They are made by either a thermal inversion process, or by laminating a thin layer of polymeric membrane material onto a supporting porous substructure of polysulfone. Zirconium and aluminum or titanium oxide are used for ceramic ultrafilters. Because they are primarily used in cross-flow filtration mode, much of what applies to cross-flow microfilters also applies to ultrafilters, with some exceptions.

3.1 Characterization of Ultrafilters

Unlike microfilters, ultrafiltration membranes do not have pores made by mechanical processes and consequently, do not have a sharp cutoff in pore size. Ultrafilters are larger in area and are thinner than microfilters, which may result in a higher defect rate. For these reasons, the pore size of ultrafilters is difficult to pinpoint accurately.

By definition, the pore size of an ultrafilter ranges from 10 to 1000 Å (0.001 to 0.1 µm). Another way of indicating pore size is with the molecular weight cutoff, usually given in Daltons or gram-molecular weight. Instead of using micro-organisms, as with microfilters, globular proteins of various molecular weights are used to estimate the size of molecules retained by the membrane. The definition of molecular weight cutoff (MWCO or NMWCO, nominal MWCO) generally used is the molecular weight of globular proteins that are 90 percent retained by the membrane (Porter, 1990). Manufacturers produce ultrafilters with a range of "nominal" MWCOs and then specify the percent rejection of proteins with a similar range of molecular weights for each membrane. Table 3.1 is a typical pore size specification list for a range of ultrafilters. Rejection percentages decrease with increasing NMWCO.

Because the specified MWCO is only an estimate of the effective pore size, the bubble point method described for microfilters is used to determine the maximum pore size and give an indication of pore size variation. Broken fibers or bubbles in the seals will allow particles to get through the membrane. The shape of the particles being separated also affects the effective pore size. Figure 3.1 illustrates how straight chain or branched molecules with molecular weights substantially over the NMWCO can slip through where a much smaller globular shaped molecule cannot. This process depends on the alignment of molecules as they contact the membrane, and the flow of water through the pores.

Table 3.1. – Proteins used in pore size determination for ultrafiltration membranes.

Solute	Nominal molecular weight	Nominal molecular weight cutoff (NMWCO)					
		5,000	10,000	30,000	50,000	100,000	500,000
Bacitracin	1,400	---	76.5%	___	___	---	---
Bovine Trypsin Inhibitor	6,500	---	82.5%	---	---	---	---
Ribonuclease A	13,700	---	95.3%	---	---	---	---
Dextran-20	15,000 - 20,000	---	98.0%	---	---	---	---
Myoglobin	17,500	---	99.5%	---	---	---	---
PVP K-30	40,000 - 50,000	97.0%	---	---	---	---	---
Ovalbumin	43,000	---	---	---	---	78.5	---
BSA	66,000	---	---	---	99.5%	87.5%	---
Bovine IgG	166,000	---	---	---	---	99.4%	---
PVP D-90	630,000	---	---	---	---	---	<25%
Blue Dextran	2,000,000	---	---	---	---	99.5%	---

Adapted from AGT Technical Bulletin, RTB01-7/86, and Porter, 1990.

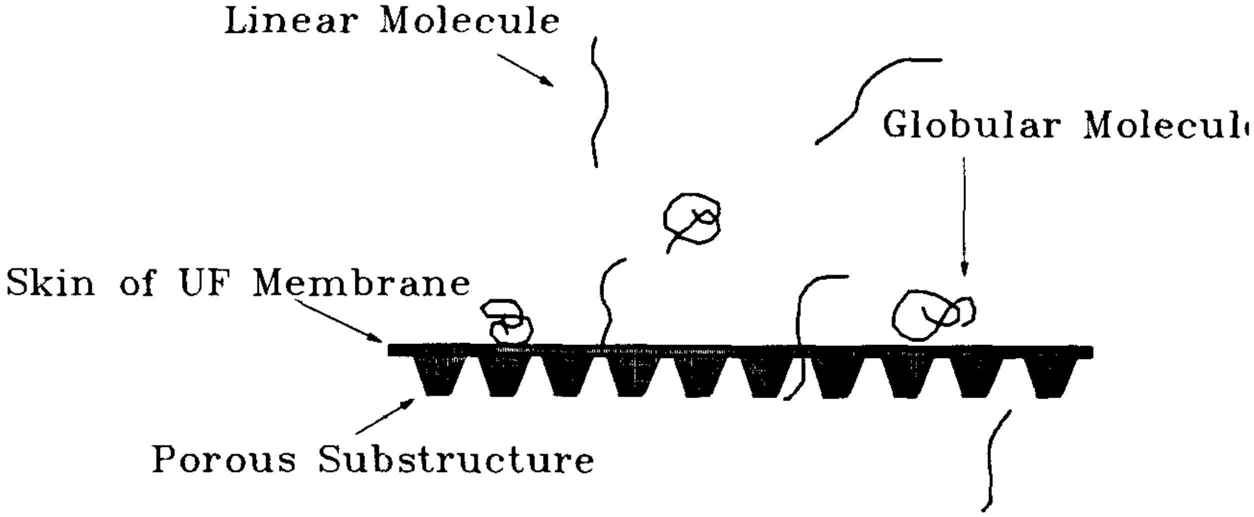


Figure 3.1. – Effect of shape on molecular retention.

3.2 Configurations

Ultrafilters come in the same base configurations as microfilters. Examples of plate and frame, tubular, and spiral wound configurations are shown on figure 2.2.

Depending on the application, plate and frame modules may be stacked flat or hung vertically from a support beam. The individual frames can be taken out and cleaned or replaced when needed without dismantling the whole assembly.

Tubular membranes are composed of a porous support tube with the membrane material deposited on the inside surface. Rhône-Poulenc produces Carbo-sep ultrafiltration membranes that use porous carbon support tubes plated on the inside with zirconium. The tubes have a diameter of 6 mm and hundreds can be packed in a single stainless steel or plastic pressure vessel. A/G Technology Corporation produces polysulfone hollow fiber ultrafiltration membranes with internal diameters of 0.5 and 1.0 mm. These tubes are true anisotropic membranes with a microporous substructure and finely porous internal surface layer.

Like tubular microfilters, both models have inside to outside flow paths. Feed water enters through the interior of the tubes and permeates to the outside. Product water is collected from permeate ports on the side of the module. Concentrate exits opposite the feed end.

Table 3.2 compares aspects of each configuration. Cost/area refers to installation cost relative to the membrane area within the module. Tubular membranes have a comparatively high installation cost per area because a given module size has less surface area than the other types. Plate and frame configurations are costly because of the hardware involved for each section of membrane. Replacement costs differ because of differences in the components that must be replaced when the membranes go bad. In tubular, hollow fiber, and spiral wound configurations, the whole module must be replaced. In the plate and frame configuration, only the sheet of membrane must be replaced. If labor costs are included, plate and frame modules are still most expensive to replace.

Comparing the water flux of different configurations is not easy. If all other factors are equal, the amount of water passing through the membrane depends on the membrane thickness. Because the actual membrane layer inside the tube is quite thin, tubular membranes can have as high a water flux as other configurations using thicker membranes. Unfortunately, different configurations of ultrafilters are not directly comparable from the specification sheets. Test parameters are different or may not even be specified. Therefore, the flux ratings given in the tables in chapter 10 should be taken as a very rough comparison of performance.

Packing density refers to the ratio of surface area to module volume. Tubes are not as compact as other configurations, but they have other advantages that compensate. Hold-up volume refers to the volume required to fill the system before filtration takes place. Since tubular filters have a greater volume to surface area ratio, they also have a greater hold-up volume. Their high hold-up volume also means tubular membranes require the most energy.

Table 3.2. – Comparison of ultrafiltration configurations.

	Tubular	Hollow fiber	Plate and frame	Spiral wound
Installation Cost/Area	High	Low	High	Low
Membrane Replacement Cost (not including labor)	High	Moderate	Low	Moderate/Low
Flux (GSFD)	Good	Fair/Poor	Excellent/Good	Good
Packing Density (m^2/m^3)	Poor	Excellent	Good/Fair	Good
Hold-up Volume	High	Low	Medium	Medium
Energy Consumption	*High*	*Low*	*Medium*	*Medium*
Fouling	Excellent	Poor	Good/Fair	Good/Fair

(From Porter, 1990)

However, tubular membranes excel in the maintenance department. The tubular configuration allows a greater volume of water to pass through which alleviates fouling problems. Membranes made of durable metallic materials can withstand harsh cleaning regimes. Some can even be cleaned during operation by introducing sponge rubber balls slightly larger than the diameter of the tube. Hollow fiber membranes are not easy to clean. Their diameters are so small that they can be clogged completely if larger particles get into the feed stream. If cleaning solutions cannot remove the clog, the fiber can be taken out of commission by blocking it off at both ends. Using cleaning solutions, spiral wound modules can be cleaned effectively, although they cannot be back-washed. Glued seams tend to absorb water and burst if clean water is introduced into the product side of the membrane. Plate and frame modules can be thoroughly cleaned, but they must be taken apart and each frame cleaned separately.

3.3 Ultrafiltration Separation Mechanism

Because ultrafilters do not have a sharp pore size cutoff, solute flux is mainly attributed to what is washed through the larger pores. Therefore, any parameter that increases solvent flux also increases solute flux. Pressure is proportional to flux up to a point, although pressure also contributes to surface concentration. At first, these dual effects—increased water flux and increased surface concentration—work together to increase solute flux. When surface concentration increases to the point where a gel layer forms at the surface, solute flux begins to stabilize. The gel buildup around the edges of pores shrinks their diameter. As a result, the water flux levels off, and solute passage declines. Further increases in pressure do not affect flux until compaction of the gel layer occurs. Then, both water and solute flux decline rapidly. Other parameters being equal, higher solute concentration causes the gel layer to build up sooner, and equalizes solute and water fluxes earlier after cleaning than does low solute concentration.

The cross-flow velocity of the feed stream has a positive effect on water flux. Higher velocity causes higher flux of water and solute. When operated with concentrate recirculation though, solvent flux decreases with increasing feed concentration until it reaches zero, when the feed concentration equals the gel concentration at the membrane surface.

The effect of pH on ultrafiltration flux depends on its effect on the solubility or structure of the solute. Lower solubility results in faster gel layer formation and thus lower water and solute flux. Lower temperatures decrease solubility and increase viscosity of the feed solution, which also speeds up the gel layer formation process.

Changing the pH or temperature of the feed water usually is not practical. Pressure and cross-flow velocity can be adjusted somewhat, but the surface area to volume ratio and pore size and variability are the factors that can have the most profound effect on water and solute flux. Filtration configurations that bring the feed water in contact with the membrane in the most efficient manner should have best all around performance.

3.4 Applications

Ultrafiltration is primarily used in industrial, pharmaceutical, and food processing applications where recovery of valuable waste products offsets the capital costs of ultrafiltration. Laundries, car washes, and other industries can reduce water and sewer costs by using ultrafiltration to recycle water. Metals can be recovered from electroplating and photographic rinse water. In the dairy industry, ultrafiltration is used to concentrate milk products. It is also useful for clarifying juice, beer, broth, and wine. The pharmaceutical industry uses ultrafiltration to remove pyrogens from injection water. Higher MWCO ultrafilters are used in biotechnology to harvest enzymes and other metabolic products.

In municipal water treatment, ultrafiltration can be used to filter water from sewage treatment bioreactors. Biomass is built up in a fermentation tank to a concentration of up to 50,000 mg/L. The bacteria digest the sewage, breaking down organics that cause color and odor. The ultrafilter keeps the bacteria in the tank while removing water, small molecular weight organics, and salts for further processing. The sludge in the fermentation tank can be left indefinitely with only small amounts being removed bimonthly. This process allows time for the metabolism of slightly biodegradable substances (Stavenger, 1971).

Low MWCO ultrafiltration membranes can be used to remove pesticides and other large organic compounds causing problems in the water supply. Watson and Hornburg (1989) found that a 500 Dalton MWCO UF removed 90 to 95 percent of THM's (trihalomethanes) from water. THM's are carcinogens regulated by the Safe Drinking Water Act. They are formed from reactions of naturally occurring humic and fulvic acids with chlorine used for disinfection. Humic and fulvic acids are large, rambling molecules. Larger MWCO ultrafilters possibly could be used to remove these acids before they are converted to THM during disinfection.

4. NANOFILTRATION AND REVERSE OSMOSIS

Nanofiltration and reverse osmosis are treated together here, as they are in many product catalogs, because they are similar processes. Nanofiltration membranes are even referred to as "loose" RO membranes. The difference between them is that nanofiltration membranes have lower rejection rates for monovalent ions than RO membranes. RO membranes generally have a 99.5-percent rejection rate for NaCl and a 99.99-percent rejection rate for $CaCl_2$. A nanofiltration membrane might have a 75-percent rejection rate for NaCl and 99.5 percent for $CaCl_2$ and other divalent salts (see table 10.7).

4.1 The Hyperfiltration Process

Nanofiltration and reverse osmosis are known as hyperfiltration because they are capable of separating dissolved ions from a feed stream. The membranes used in these processes do not have actual pores. Under pressure, water passes through spaces in the polymeric structure of the membrane. The apparent size of openings in the structure of a nanofilter is from 8 to 80 Å. RO membrane openings are 1 to 15 Å. The atomic diameters of some ions and molecules are listed in table 4.1 for comparison. As is apparent from the table, the size of spaces in the membranes does not prevent salt permeation. Ions are rejected even though they should have plenty of room to pass through. The reason for this phenomenon is that the ions of dissolved salts are held together in a matrix by weak, transient bonds between positively and negatively charged ions. The expanding force of this matrix is the osmotic pressure of the solution. Undissociated molecules, such as water and molecules of slightly soluble salts, do not contribute to the osmotic pressure. They are merely suspended within the ionic matrix.

In a system under atmospheric pressure such as on figure 4.1, water will be absorbed through the membrane from a low ionic concentration solution to a higher ionic concentration solution. The driving force behind this water transport is the difference in the chemical potentials of the two solutions. The low ionic concentration solution possesses a relatively high chemical potential compared to the higher ionic concentration solution. In effect, the water passing through the membrane is trying to dilute the higher ionic concentration solution to equalize concentrations on both sides of the membrane. If pressure is applied to the right side of the system, the migration of water will slow down. When it stops, the applied pressure will be equal to the osmotic pressure of the concentrated solution. Increasing the pressure beyond this point, like squeezing a sodden sponge, will cause water from the concentrated salt solution to permeate through the membrane to the other side. In general, the osmotic pressure of water is increased 69 kPa per 1000 mg/L TDS (total dissolved solids).

Table 4.1. – Atomic diameters of ions normally found in water supplies.

Species	Diameter (Å)
Sodium (Na^+)	0.95
Chloride (Cl^-)	1.81
Calcium (Ca^{+2})	0.99
Magnesium (Mg^{+2})	0.65
Hydrogen fluoride (HF)	1.73
Water (H_2O)	2.14

Because undissociated molecules are not an integral part of the ionic matrix, they are washed through the membrane with the water molecules. Flux of these components depends on their concentration at the membrane surface. Small polar molecules encounter resistance, but non-polar molecules are not rejected at all unless they are too large to fit through the spaces in the membrane matrix.

Some dissolved salts may pass through the membrane, but ionic charges on either side of the membrane must balance. An equivalent charge of anions and cations must pass at the same time. This process is more likely to happen with nanofiltration membranes because they have larger pores than RO membranes. Divalent cations must bring two monovalent anions with them (and vice versa), and so are less likely to pass through nanofiltration membrane pores.

Because of this close relationship between feed water composition and membrane performance, specifications are difficult to obtain from manufacturers without a water analysis. Typically, the best results obtainable are the performance records for a test run with a given salt concentration. Tests are usually run with sodium chloride solutions made with reverse osmosis permeate. Therefore, the performance characteristics listed in chapter 10 should be taken as a best case scenario.

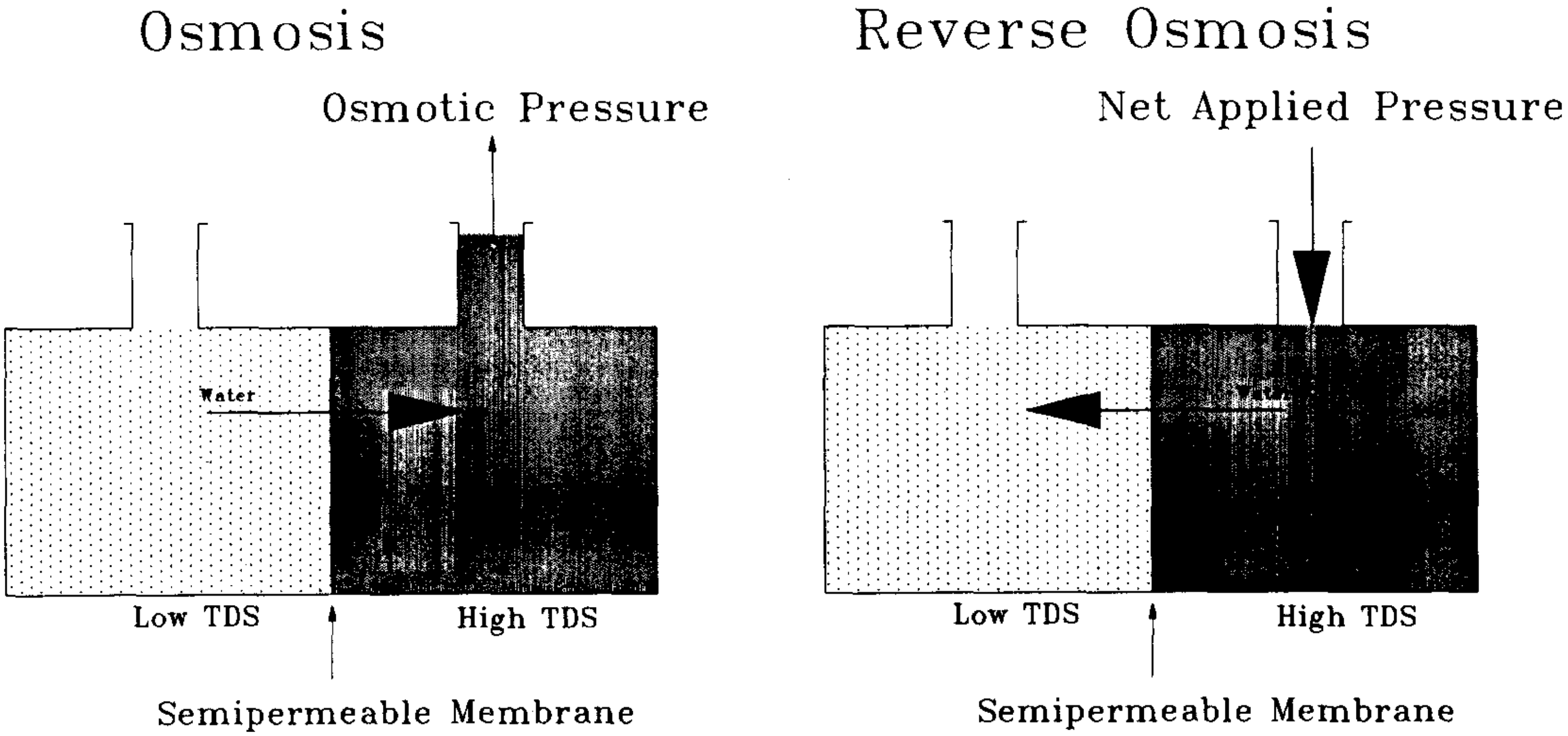


Figure 4.1. – Transport of water and salts in osmosis and reverse osmosis.

4.2 Hyperfiltration Membranes

Nanofiltration and reverse osmosis membranes are produced primarily through the thermal phase inversion process described in chapter 2. The difference between microfiltration membranes produced by this process and membranes used for hyperfiltration is the porosity of the membrane. Porosity is controlled by the optimizing the factors in table 2.1. The best form for a reverse osmosis membrane is a tight, thin surface layer covering a porous base. A nanofiltration membrane would have a slightly more porous surface layer. One type of membrane incorporates these two requirements by using two different membranes. Called thin film composites, these membranes join the best qualities of both asymmetric and symmetric membranes. By laminating an exceptionally thin, tight, highly selective membrane to an open, durable, microporous membrane, one obtains a membrane with a high water flux rate and a good salt rejection rate.

The trick to producing a composite membrane is to cast the 1- to 10-µm thick perm-selective barrier layer onto a microporous membrane that may be 100 µm thick. The barrier layer can be cast on water and then laminated to the support layer, but this process is difficult to do on a large scale with such a delicate film. The method most used for industrial production is polymerization of a reactive monomer on the surface of the support film. Other methods involve casting the barrier layer directly onto the support layer either by dip coating or gas phase deposition (Strathmann, 1990).

4.3 Membrane Characteristics

All reverse osmosis and nanofiltration membranes have four general characteristics no matter what they are made of: water flux, salt flux, salt rejection rate, and recovery rate. Flux is the amount of water or salt that passes through, or permeates, a unit area of membrane at a given salinity, pressure, temperature, and pH. The relationship between the flow of water through a membrane, applied pressure, and salinity of feed water is given by the equation:

$$F_w - A\ (\Delta P - \Delta\pi) \tag{1}$$

where:
F_w = water produced per unit area of membrane expressed as m^3/m^2 per day,
A = the water transport coefficient in m/s*kPa,
ΔP = difference in pressure across the membrane in kPa,
$\Delta\pi$ = difference in osmotic pressure across the membrane in kPa.

Salt flux, the amount of dissolved salt that passes through the membrane, is given by the equation:

$$F_s = B\ (C_1 - C_2) \tag{2}$$

where:
F_s = weight of salt passed per unit area and unit time expressed as moles/m*day,
B = salt permeability constant for the membrane in m/s,
$C_1 - C_2$ = difference in salt concentration across the membrane in moles.

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

$$R = (1 - C_p/C_f) * 100 \tag{3}$$

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

Recovery rate is the ratio of product water to feed water given as a percentage:

$$\text{Recovery Rate} = 100 * N_p/C_f \tag{4}$$

where:
N_f and N_p are the volume of feed water processed, and the volume of treated water produced respectively.

Equations (1) and (2) show that more efficient membranes will have a high "A" value, o water permeability coefficient, and a low "B" value, or low salt passage coefficient. Thes values are calculated for membranes compared in chapter 10 whenever possible.

4.4 Effects of Varying Operation Parameters

Membrane performance depends on the chemical make up of the feed water, but given standard feed water, performance will vary with pressure, water temperature, level of wate recovery, and the oxidation potential of the feed water. The first three of these factors ar related to the feed water composition. The last is related to the material used in th membrane. Figure 4.2 illustrates how applied pressure, feed water temperature, and wate recovery affect RO membrane flux and water quality. The graphs are generalized curve intended to show trends; they are not based on actual data points.

4.4.1 *Effect of Pressure.* – The graph on figure 4.2 (a) and equation (1) both show that wate flux is directly proportional to applied pressure. At higher pressures, the feed water is force against the membrane at a higher velocity, forcing a greater number of foulants in the fee stream to interact at the membrane surface.

Scaling problems increase at high pressures. Salts left at the membrane surface increase the local osmotic pressure. As the surface pressure exceeds the main stream osmotic pressure, ions diffuse away from the membrane surface. At higher operating pressures, salts have to become more concentrated before diffusion can take place, which can result in precipitation of slightly soluble salts.

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.

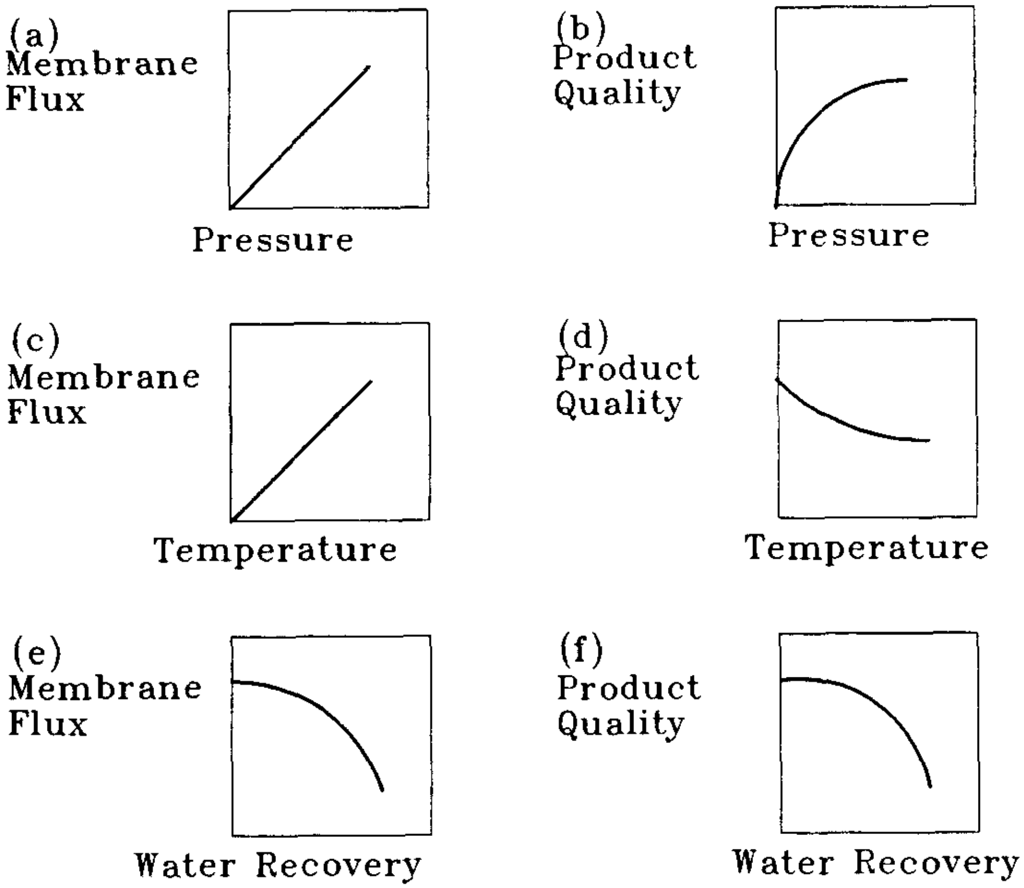


Figure 4.2. – The effects of applied pressure, feed temperature, and water recovery on membrane flux and product water quality of reverse osmosis membranes. (From Riley, 1990.)

4.4.2 *Effect of Temperature.* – The effects of temperature on RO and nanofiltration membranes are 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 generally provide a table or formula for determining TCF (temperature correction factors). The TCF is proportional to the change in pressure needed to maintain the 25 °C flux rate and is equal to 1.00 at 25 °C. The TCF increases at temperatures less than 25 °C and decreases at higher temperatures.

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 ground water 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; thus, higher pressure is needed to maintain a standard 25 °C flux rate.

4.4.3 *Effect of pH.* – The pH of the feed water can affect the membrane structure and the scale formation potential of the brine 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 10.5 and 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 by the brine 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 or the temperature. However, if the brine 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.

4.4.4 *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 cellulose acetate 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. Other methods of disinfection will be discussed later. Here, we will discuss how exposure to halogens, chlorine dioxide, and ozone affect cellulosic membranes and those made with synthetic polymers.

4.4.4.1 *Halogens.* – Polymer oxidation by halogens depends on pH as much as the presence of oxidants. The halogens chlorine, bromine, and iodine may be used for disinfecting water supplies. When mixed with water, they dissociate by the following reaction (X represents the halogen atom):

$$X_2 + H_2O \leftrightarrows HOX + H^+ + X^- \tag{5}$$

$$HOX \leftrightarrows H^+ + OX^- \tag{6}$$

Table 4.2 lists the proportion of each species, X_2, HOX, and OX^-, for chlorine, bromine, and iodine at pH 3.0, 5.6, and 8.6.

The HOX species is the most damaging to polyamide and polyether urea membranes. Cellulosic membranes are resistant to halogens within their normal operating pH range of 4 to 6. Polyvinyl alcohol membranes are resistant over a pH range of 2 to 8.

Two different manifestations of membrane oxidation occur in the presence of halogens. One is an increase in both salt and water flux. The other is a decrease in water flux in which the salt flux remains constant or even decreases. The first set of symptoms indicates breakdown of the polymer. In scanning electron micrographs, holes are shown scattered over the membrane, allowing feed water to enter the product side of the membrane. The second set suggests that halogen atoms may be forming cross-links in the polymer matrix that prevent both water and salt from passing through. If conditions continue, the membrane will become brittle, resulting in a dramatic increase in water and salt flux.

In McCutchan and Glater's study (1980), the homogenous aromatic polyamide membrane was susceptible to the second form of decay when exposed to chlorine and bromine. Under the same conditions, polyether urea and aromatic polyamide composite membranes demonstrated the first symptoms. Both conditions were exacerbated at lower pH levels.

Table 4.2. – Proportional distribution of halogen species.

	pH	X_2	HOX	OX^-
Chlorine	3.0	0.7	99.3	0.0
	5.6	0.04	98.5	1.4
	8.6	0.0	6.5	93.5
Bromine	3.0	63.1	36.9	0.0
	5.6	7.9	92.0	0.1
	8.6	0.0	0.0	Most all
Iodine	3.0	99.5	0.5	0.0
	5.6	91.0	9.0	0.0
	8.6	0.0	0.0	100.0

(From McCutchan and Glater, 1980)

4.4.4.2 *Ozone and Ultraviolet Light.* – 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). When used in RO systems, ozone is bubbled into the feed water allowing sufficient time for reaction. Ultraviolet light is then 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.

When a parts per million limit is set for oxidant tolerance, it is usually given as mg/L chlorine or equivalent tolerance. Chlorine (Cl_2) has a molecular weight of 70.9 and ozone (O_3) has a molecular weight of 48.0, a ratio of 1:0.67. Oxidative effects are proportional to the number of molecules available; therefore, membranes that can tolerate 0.1 mg/L chlorine can only tolerate 0.067 mg/L ozone.

4.5 Applications

Nanofiltration membranes are used in Florida for treating hard ground water. In tests with 100- to 500-MWCO nanofilters, organics, color, and hardness were rejected at 85 to 95 percent (Watson and Hornburg, 1989). Nanofilters can be operated at lower applied pressure and, if the sodium concentration is acceptable, are an option for removing hardness.

RO is used for desalting brackish water or seawater. Because demand for water is increasing, many coastal communities are beginning to hire companies to build desalting facilities and then buy water from them. In Santa Barbara, California, where prolonged drought has threatened the water supply, city officials contracted with Ionics, Inc. of

Watertown, Massachusetts, to supply freshwater from the ocean. Ionics owns and operates the RO plant, which is designed to supply up to 92.5 million m^3/yr to the city. The city pays a fee per 1200 m^3 which includes capital facilities and water delivery charges, based on whether the plant is in operation, and, if so, whether it is on short- or long-term standby. When the plant is paid for in 5 years the city has the option of taking over ownership and operation, or continuing the agreement with Ionics (Lizarraga and Brown, 1992).

RO has become an economical solution to water supply shortages. Even inland communities like Las Vegas, Nevada, and areas like the San Joaquin Valley in California, where reservoirs have become brackish through agricultural runoff and evaporation, are candidates for RO water treatment plants.

5. ELECTRODIALYSIS

Until now, the membranes that have been discussed operate under hydrostatic pressure. In ED (electrodialysis), an electric current is used to draw ions out of the feed solution. An ED unit has paired anion and cation exchange membranes (D and E on figure 5.1) stacked between positive and negative electrodes (C). As feed water (A) containing dissolved salts passes between pairs of membranes, negatively charged ions are attracted toward the positive electrode and are allowed to pass through the anion exchange membrane, and the positively charged ions are allowed to pass through the cation exchange membranes toward the negative electrode. A portion of the feed stream is used to carry the concentrated salts out of the system. The demineralized product water is collected from the feed channel (G), and the brine is collected from the concentrating compartments (F) and exits through the reject port.

Channeled spacers inserted between each membrane direct the feed water through the stack via a tortuous route that maximizes exposure to the membrane. The spacers separate the flow into many streams that move in parallel paths back and forth across the surface of the membrane. Baffles across each channel create turbulence that further assists in ion transport by mixing the water, which helps dissolve more of the slightly soluble salts.

The membranes are formed from polymers with charged groups incorporated into the membrane matrix. For instance, cation exchange membranes have fixed negative ion groups, such as SO_3^-, and positively charged, free moving, counter ions, such as Na^+. Anion exchange membranes have positively charged fixed groups and negatively charged counter ions. The fixed ion group repels like-charged ions in the feed solution while oppositely charged ions are attracted and allowed to pass through. The membranes are impervious to water under pressure. Only ions are transported. Dissociated water molecules can pass through the membrane; the hydroxyl ion crosses the anionic membrane and the hydrogen ion crosses the cationic membrane.

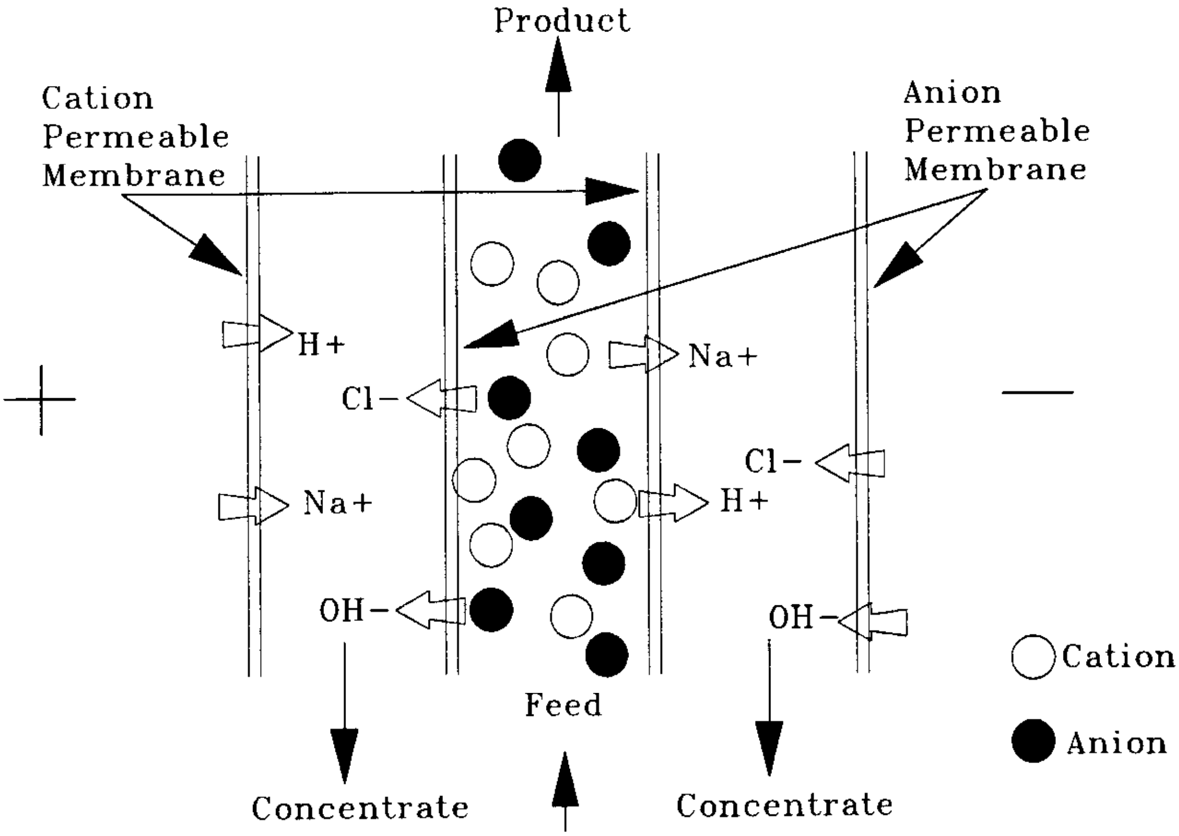


Figure 5.1. – Ion transport in electrodialysis.

The following are some other traits of both cationic and anionic ED membranes:

- Low electrical resistance,
- Insoluble in water,
- Semi-rigid for ease of handling during stack assembly,
- Resistant to change in pH from 1 to 10,
- Operate at temperatures in excess of 46 °C (115 °F),
- Resistant to osmotic swelling (water uptake),
- Long life expectancies,
- Resistant to fouling.

5.1 Factors Affecting ED Separation

Factors that affect the ED process are the composition of the feed water, the selectivity of the membrane, the electrical current supplied to the system, and the electrical resistance of the system.

5.1.1 *Composition of the Feed Water.* – Only ionized substances can be separated with ED Non-ionic species, such as undissociated molecules of slightly soluble salts, are unaffected by the electrical current. However, the ionized concentration of the salt will be separated, which in turn allows more of the undissociated molecules to ionize. As an example, magnesium hydroxide, $Mg(OH)_2$, dissolves in the following manner:

$$Mg(OH)_2 \leftrightarrow Mg^{2+} + 2\ OH^-$$

The solubility product of magnesium hydroxide is 1×10^{-11}, which means that the concentration of magnesium cations multiplied by the square of the hydroxide anion concentration must be equal to 1×10^{-11} moles/L. If the concentration of either of these components causes the solubility product to shift, magnesium hydroxide will either precipitate or dissolve. In an ED system where the Mg^{+2} and OH^- ions are in equilibrium with undissociated molecules of $Mg(OH)_2$, the Mg^{+2} and OH^- can be separated from the solution. Then molecules of $Mg(OH)_2$ will be ionized to bring the equilibrium back into balance. These ions can in turn be separated from the feed stream and so on until the feed stream becomes so dilute that the osmotic pressure of the concentrate stream is great enough to overcome the electrical attraction of the current running through the electrodes. At this point, ions begin to diffuse back into the feed stream.

5.1.2 *Membrane Selectivity.* – The selectivity of the membrane, or how well it transfers ions, depends on the number of ions embedded into its framework and the thickness of the membrane. Ionic's ED membranes have a selectivity of 90 percent (Meller, 1984). Other factors that affect selectivity are related to the concentration of salts in the feed and concentrate streams, and the resistance of the membrane. High salt concentration and low temperatures decrease membrane selectivity because of the higher osmotic pressure gradient and the lower mobility of ions at lower temperatures.

5.1.3 *Faraday's Law.* – Faraday's Law states that "the passage of 96,500 amperes of electric current for one second will transfer one gram equivalent of salt" (Meller, 1984). This number, 96,500 ampere-seconds, is known as a Faraday. The time factor is not as important as the total current; 26.8 amperes for one hour will still transfer one gram equivalent of salt. The gram equivalent of an ion is its molecular weight in grams divided by its charge and is expressed as N, for Normality, or gram equivalents per liter. A gram equivalent of sodium is 23 grams (MW 23/1 charge) and a gram equivalent of calcium is 20 grams (MW 40/2 charge). Unless the composition of the water is known, the gram equivalent is obtained by assuming that the TDS concentration is entirely sodium chloride. For instance, if the TDS is 5000 mg/L, then the normality of the solution is 0.086 N (5000 mg/L/58,400 mg per equivalent). The current required to remove a given number of gram equivalents is calculated with Faraday's Law as follows:

$$I = \frac{F \times F_d \times \Delta N}{e \times N}$$

(From Meller, 1984, p. 30)

where:

I = direct electric current in amperes,

F = Faraday's constant = 96,500 ampere seconds/equivalent,

ΔN = change in normality of demineralized stream between the inlet and outlet of the membrane stack,

F_d = flow rate of the demineralized stream through the membrane stack (L/s),

e = current efficiency,

N = number of cell pairs.

The voltage requirement is calculated from Ohm's law which states that "the potential (E) of an electrical system is equal to the product of current (I) and the system resistance (R)" (Meller, 1984). E is expressed in volts, I in amperes, and R in ohms. The resistance of the membrane is made up of four components: the resistance of the cation membrane, the resistance of the anion membrane, the resistance of the concentrate stream, and the resistance of the demineralized stream. Overall resistance decreases with higher temperature and solution concentration, and with increasing percentage of sodium chloride in the solution.

5.2 Electrodialysis Reversal

In EDR, the polarity of the electrodes is switched periodically. The concentrate stream is then converted to the feed stream and the feed stream becomes the concentrate stream. This process requires more involved plumbing and electrical systems than ED. Reversing the flow increases the life of the electrodes and helps clean the membranes. When the membranes are operated in the same direction all the time, precipitant can build up on the concentrate sides. Switching these compartments to demineralization compartments helps dissolve the scale build up.

6. COUPLED TRANSPORT

Coupled transport is similar to electrodialysis in that metallic ions are removed from the brine and concentrated in a reject solution rather than removing the water from the brine solution, as in reverse osmosis. Like ED, coupled transport is much more specific than reverse osmosis or nanofiltration. Coupled transport is also more specific than ED. This specificity occurs because the force driving the separation in coupled transport is chemical, not electrical. The most obvious difference however, is that the membranes used in coupled transport are liquid. These membranes actually have two parts. The first is the membrane itself, which is a water insoluble liquid that contains a complexing agent which combines with the metal ion that is to be removed. The other part is a stripping agent that removes the ion from the complexing agent at the inner surface of the membrane, exchanging it for a counter ion, such as a hydrogen cation.

A good correlation for coupled transport is a miner who looks for gold in a stream. If he finds some, he brings it back into town and gives it to the bartender, who in turn gives him a bottle of whiskey. When the miner runs out of gold he goes back to the stream to find some more. Now, if the miner cannot find any gold, the bartender will not give him any whiskey, and likewise, if the bartender runs out of whiskey, the miner will not leave him any gold. The complexing agent and the stripping agent are the miner and the bartender respectively. The gold and whiskey represent the metallic ion and the counter ion.

The complexing agent is dissolved in an organic solvent that does not mix with water. The stripping agent may be soluble in water, but not in the solvent for the complexing agent. In one coupled transport system, the stripping solution is emulsified into the complexing solution and this emulsion is then mixed into a tank of mineralized water to produce an emulsion within an emulsion. Bubbles of liquid membrane are formed with the stripping solution inside the bubble. Metal ions in the feed water are transported across to the interior of the bubble by the complexing agent. After a gentle mixing period, the bubbles are allowed to settle at the top of the tank (like oil), and the demineralized water is drained out from the bottom. The emulsion is broken with a surfactant to release the stripping solution, which now contains metal ions, from the complexing solution. Because the two components of the membrane do not mix, the stripping solution can be drained from the bottom of the tank. Both metals and complexing agent can be recycled.

The membrane can also be supported within the pores of microfilters. In this scenario, the microfilter support is placed between adjoining vats with the mineralized water on one side and the stripping solution on the other. Both sides are gently stirred to keep the separation going. Theoretically, it should go until the water is completely depleted of the metal ion.

Complexing agents are substances that bind chemically with a particular ion, or class of ions. The complexing site on the molecule is called the carrier (R). The carrier combines with a counter ion, such as H^+, or SO_4^{-2}, from the stripping solution. As the concentration of R-H complexes build up, they diffuse to the feed side of the membrane. At the feed side surface of the membrane, R-H groups react with metallic cations and release the hydrogen (or other counter ion). As the metallic carrier species concentration builds up at the feed side, they diffuse to the product side of the membrane where the metallic ion is exchanged for another counter ion. In this way, the metallic ions are eventually all moved to the product side in exchange for counter ions. Figure 6.1 diagrams the reactions taking place at the membrane surface and the transport process.

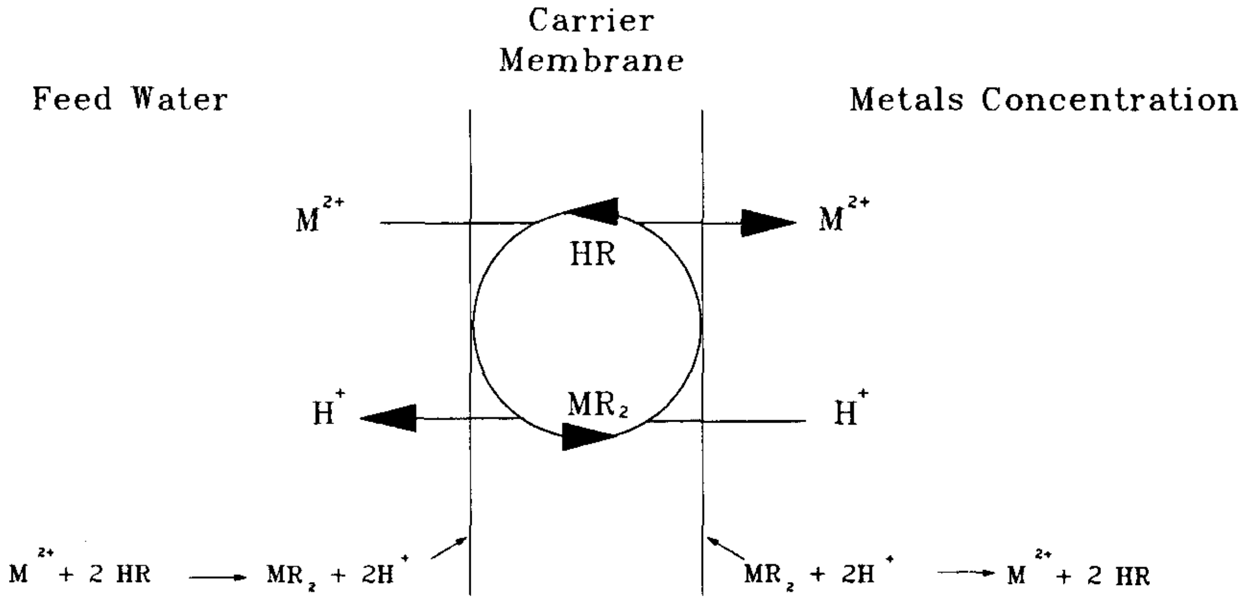


Figure 6.1. – Mechanism of coupled transport for generic metals.

The separation can go to completion because the concentration differential of the counter ion, not the metal ion, is driving the separation. Metal ions can therefore be concentrated on the product side without having back diffusion as in ED. Coupled transport continues in this way until the metal ion concentration gradient becomes greater than the counter ion's concentration, which will never happen as long as the stripping solution is replenished.

The coupled transport process is not regularly used in municipal water treatment systems, but it may be practical on a small scale for problems with heavy metal contamination. In this area, RO may be beneficial in concentrating the metals to a volume where coupled transport can be used economically. For more information on liquid membranes and coupled transport, see R. Baker and I. Blume's chapter, "Coupled Transport Membranes," in the *Handbook of Industrial Membrane Technology* (ed. M.C. Porter, 1990), or E.L. Cussler's chapter, "Coupled and Facilitated Transport," in *Membrane Separation Systems.*(U.S. DOE)

7. PRETREATMENT PROCESSES

The importance of water 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 reverse osmosis and nanofiltration 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 it is so important that a summary of common problems and technologies for remediation is offered here, with references in appendix A for further information.

7.1 Particulates

Cartridge microfilters are designed to remove at least 90 percent of particles larger than their rating 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 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. For the purity level required by membrane processes, the cost is $0.01 to $0.08 per cubic meter using twenty-five 5-µm filters (Parekh, 1991).

7.2 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 (see fig. 7.1). 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 hollow fine fiber membranes (O'Melia, 1972).

7.2.1 *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 sand as a filter, they reduce the SDI, the turbidity, and the salinity of the intake water.

7.2.2 *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 (see figure 7.1). 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).

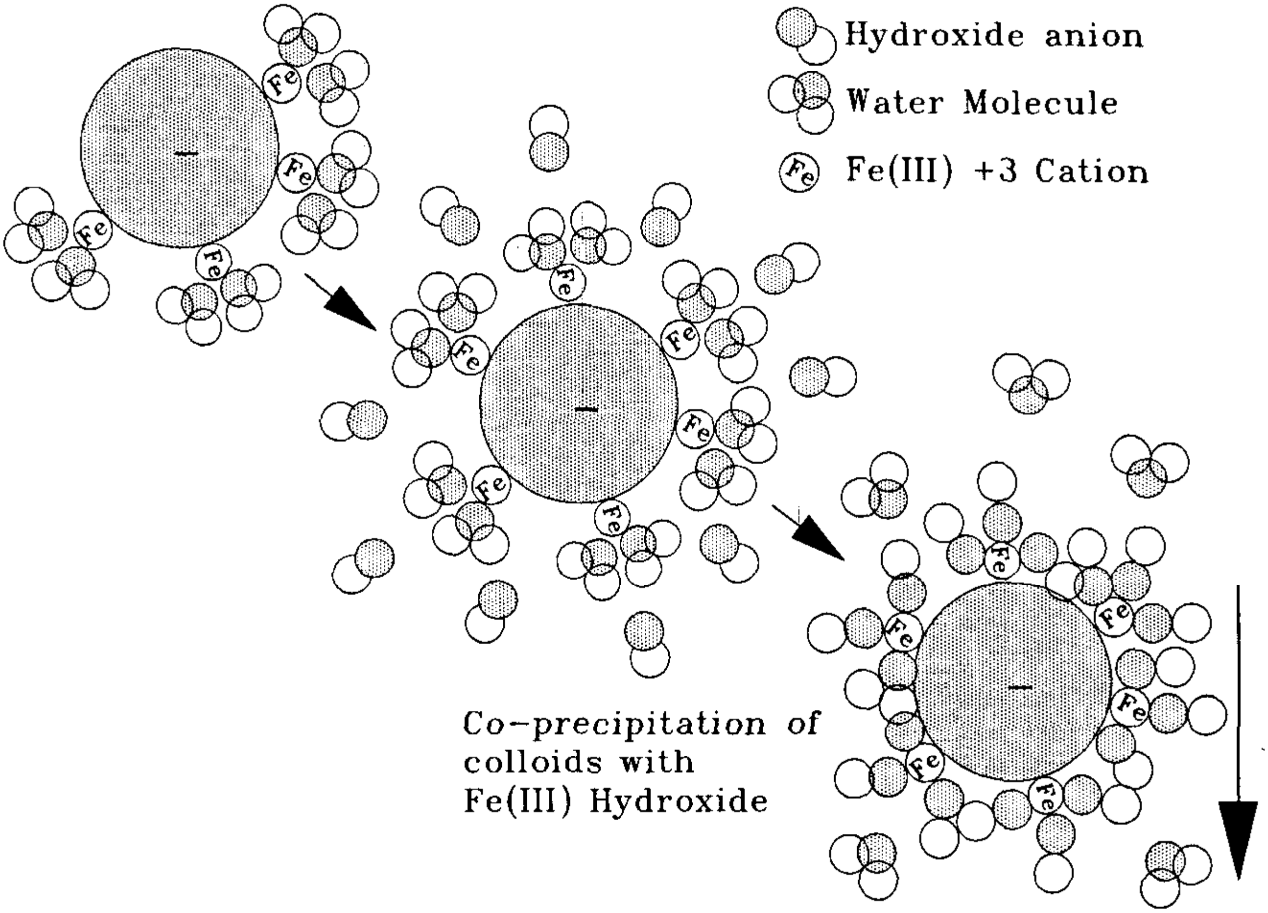


Figure 7.1. – Double charge layer formation around colloidal particles and co-precipitation with metal hydroxides.

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.

7.2.3 *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 has an SDI of 6 or higher (Permasep Engineering Manual #506, 1982).

7.2.4 *Ultrafiltration.* – Ultrafilters with molecular weight cutoffs 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 ultrafiltration as opposed to the traditional precipitation processes. Ultrafiltration 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 RO level.

7.3 Iron and Manganese

Iron is found in ground water and deep surface water low in dissolved oxygen. It may be in either the soluble ferrous state (Fe^{+2}); 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 (Degrémont, 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.

7.3.1 *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.

7.3.2 *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 $FeCO_3$.

7.3.3 *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: 0.5 mg/L
pH 6.3
Redox potential 100 mV

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

7.3.4 *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:

$$Mn^{2+} + MnO_2 \rightarrow MnO_2{\cdot}Mn^{2+} \quad \text{Fast Rate} \qquad MnO_2{\cdot}Mn^{2+} + O_2 \rightarrow 2MnO_2 \quad \text{Slow Rate}$$

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:

$$Mn_2O_3 + 2MnO_4^- + 2H_2O \rightarrow 5MnO_2 + 4H^+$$

Theoretically the required ratio of $KMnO_4$ to Mn is 1.9:1.0.

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 0.9:1.0. The addition of excess oxidant will produce MnO_4^-, which turns the water pink (Degrémont, 1991).

7.3.5 *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, 1991).

7.4 Scaling

Scaling is caused by the precipitation of slightly soluble salts such as $MgCO_3$, $CaCO_3$, $CaSO_4$, $BaSO_4$, $SrSO_4$, CaF_2, 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 so 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:

$$\text{LSI} = \text{pH}_R - \text{pH}_S.$$

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_3 to CO_2;
- Addition of lime and soda ash – causes precipitation;
- Ion Exchange – replaces cations with sodium; and
- Nanofiltration – removes dissolved divalent salts.

7.4.1 *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_2 bubbles out of solution and is no longer available for carbonate production. Bicarbonate is a potent buffer; however, 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 HCl 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.

7.4.2 *Lime softening.* – Lime, $Ca(OH)_2$, reacts with bicarbonate ions in the following manner:

$$Ca(OH)_2 + Ca(HCO_3)_2 \rightarrow 2CaCO_3\downarrow + 2H_2O \quad \text{and}$$
$$Ca(OH)_2 + Mg(HCO_3)_2 \rightarrow CaCO_3\downarrow + MgCO_3 + 2H_2O$$

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

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

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 310 g of carbonate solids. The solids are mixed with water though, after settling and pumping off the supernatant, 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.

7.4.3 *Ion exchange.* – Another approach to the carbonate cycle 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 brine may be used as regenerant solution. Cation exchange resins can be adequately regenerated at lower concentrations than 10 percent with lower flow rates and longer regeneration cycles (Haugseth and Bietelshees, 1974). However, precipitation o $CaCO_3$ may cause problems because the cations removed from the resin are reintroduced t the concentrated bicarbonate ions in the brine.

7.4.4 *Nanofiltration.* – Nanofiltration membranes preferentially reject divalent ions ove monovalent ions at a rate of about 95 percent to 75 percent. Chloride ions tend to pas through the membrane while divalent carbonate ion is retained. Whether bicarbonate ion are rejected at the same rate as chloride or not is unclear. The rate of rejection reall depends on the composition of the water. The calcium and magnesium must be balanced b an equivalent negative charge. If divalent anions are scarce then ion size will be th determining factor. Larger monovalent ions will tend to be retained and bicarbonate i almost twice as large as chloride (61:35).

The reject from nanofiltration contains 95 percent of the divalent (and higher) cations, 50 t 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_2 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, and pretreatment for bacteria.

7.5 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. However, exposure of the membranes to chloramines and free chlorine used to control bacteria growth did destroy the structural integrity of the membranes, resulting in loss of mineral rejection.

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.

7.5.1 *Oxidation.* – Several oxidants can be used for disinfection of drinking water. Table 7.1 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 UV is only effective 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.

7.5.2 *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. Single media filters are run as high as 122-163 L/min*m^2; dual and multi-media filters can be operated as high as 244-610 L/min*m^2 (Parekh, 1991).

Table 7.1. – 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 O_3 or UV	20-25 mW·s/cm^2
Conditions	No Mn^{2+}: oxidizes to MnO_2	Remove THMPs	Forms ClO_2^- in oxidation of OM	Used for post-treatment with O_3 and UV	Turbidity <1 NTU, no iron, OM, thin stream of water, clean equipment

OM is organic matter, mW·s/cm^2 is milliwatts per second per square centimeter, 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, 1991). 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).

7.5.3 *Ultrafiltration.* – As discussed earlier, ultrafiltration 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).

7.6 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_2, or filtering with granular activated carbon (GAC). Reduction with SO_2 is a fast reaction, which can work in the feed pipes on the way to the RO system. SO_2 can be sulfur dioxide gas, crystallized sodium sulphite, or sodium bisulphite. The reactions are as follows:

$$SO_2 + H_2O \rightarrow H_2SO_3$$

With free chlorine:

$$H_2SO_3 + HClO \rightarrow HCl + H_2SO_4$$

With monochloramine:

$$H_2SO_3 + NH_2Cl + H_2O \rightarrow NH_4Cl + H_2SO_4$$

GAC takes longer to deoxidize water than SO_2. 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.

7.7 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. Table 7.2 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.

7.7.1 *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 7.2. – 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 *Water Treatment Handbook*, Degrémont, 1991.

7.7.2 *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.

8. CLEANING AND STORAGE

8.1 Membrane Cleaning

Cleaning is just as important as good water pretreatment for extending the life and efficiency of the membrane. As a rule of thumb, membrane elements should be cleaned whenever any operating parameter changes 10 to 15 percent. If permeate flow drops, product water quality decreases, or applied pressure must be increased to maintain normal productivity more than

10 to 15 percent, then the elements need to be cleaned (Hydranautics, TSB 100.00, 1992). Some common foulants are mineral scaling, silica buildup, and bacterial slime. Fouling can occur when a breakdown occurs in the pretreatment system, or the quality of the feed water decreases, rendering the pretreatment system inadequate.

The cleaning process is complicated by being foulant dependent. If calcium carbonate scaling is a problem, the elements can be cleaned by just lowering the pH of the feed water for a few hours. If colloidal fouling occurs, a detergent with phosphate should be used. If bacteria fouling occurs, a biocide must be added to the cleaning solution.

How do you know what type of fouling you have? Without destroying the element, the only way to tell is by knowing the system thoroughly. Records of the feed water analysis, membrane pressure differential, water flux, salt passage, and applied pressure must be kept so that slight differences can be detected. Hydranautics offers some general symptoms for the different fouling plagues, but they are highly subjective. A "marked decrease in salt rejection and a moderate increase in ΔP between feed and concentrate," accompanied by a slight decrease in system production, may indicate carbonate scaling, and an acid cleaning procedure is used. However, a "marked decrease in salt rejection and a gradual increase in ΔP between feed and concentrate," accompanied by a gradual decrease in system production, indicates an organic deposit problem, and a basic cleaning solution is needed.

Manufacturers are very specific in recommending cleaning solutions, and for good reason. Cleaning is an area where membranes exhibit their differences. A solution that works for cleaning one membrane may damage another. Special attention must be paid to the recommended cleaning solutions, especially when using membranes from different manufacturers, or of different composition in one system. For instance, Hydranautics' PVD and PA membranes, and Fluid Systems' TFCL membranes, cannot be cleaned with cationic surfactants. Fluid Systems' TFC membrane cannot be cleaned with anionic surfactants. The Hydranautics's PVD cannot be cleaned with nonionic surfactants either, so mixing a PVD with a TFC would guarantee trouble. Included are detailed cleaning recommendations from as many manufacturers as possible to help in selecting a membrane type.

8.1.1 *A / G technology.* – In general, AGT hollow fiber and tubular ultra and micro filtration modules should be cleaned at low pressure and high velocity, at temperatures of 40 to 50 °C. The following procedures for use with water purification are excerpted from "*Hollow Fiber & Turbo Tube Ultrafiltration & Microfiltration Cartridge Operating Guide*" (AGT, OG10/92R1)

Flush. Before cleaning, flush residual feed water from cartridge with clean, warm (50 °C) water or saline solution. Flushing time is typically 30 minutes, though tighter membrane pore sizes and systems with large hold-up volumes may require a longer flushing time. After cleaning, flush residual cleaning agent from the cartridge. Flushing should be performed in a non-recirculating mode, such that the flush water does not reenter the system.

Water quality. Cleaning and flush water should contain <0.05 mg/L iron, <25 mg/L calcium and magnesium, and no colloidal silica. Water should be free of particulate matter, oil, and grease. Ideally, UF or RO permeate should be used.

Temperature. Cleaning at room temperature (i.e., 20 °C) is NOT recommended. Preferably, cleaning should be performed at 50 °C, to decrease the strength of foulant/membrane surface bonds and to improve solubility of residual feed constituents. Higher temperature cleaning (>60 °C) is not recommended due to potential cleaning chemical/membrane interactions.

Time. Recommended cleaning times are given in table 8.1. These times should be used as guidelines. Shorter or longer times may be required depending on the extent of the membrane fouling. In many cases, soaking the membranes overnight improves the effectiveness of the cleaning cycle. Flushing time depends on the cleaning chemical, the membrane pore size, and the total fold-up volume of the system.

Chlorine wash cycles. Chlorine dissipates with time and is rapidly depleted in very dirty operations. A chlorine test kit should be used to check chlorine levels and additional chlorine should be introduced as needed.

Safety. Caustic, acid, bleach, and other cleaning chemicals should be handled with care. Operators should take appropriate precautions to prevent contact with eyes and skin.

Table 8.1. – Cleaning recommendations for A/G Technology Hollow Fiber Membranes.

Foulants	Alternate cleaning procedures in order of preference
Proteins, cell debris	A. 1. Flush with clean water, buffer or saline solution at 50 °C. 2. Circulate 0.5 N NaOH at 50 °C for 1 hour. 3. Flush with clean water. *Optional* 4. Flush with H_3PO_4 at 50 °C, pH 4, for 1 hour. 5. Flush with clean water.
	B. 1. Flush with clean water, buffer or saline solution at 50 °C. 2. Circulate 500 mg/L NaOCl at 50 °C, pH 10-11 for 1 hour. 3. Flush with clean water. *Optional* 4. Flush with H_3PO_4 at 50 °C, pH 4, for 1 hour. 5. Flush with clean water.
	C. 1. Flush with clean water, buffer or saline solution at 50 °C. 2. Circulate 0.1% Tween 80® at 50 °C, pH 5-8 for 1 hour. 3. Flush with clean water.
Iron complexes	A. 1. Flush with clean water. 2. Circulate Citric Acid at 50 °C, pH 2-2.5 for 1 hour. 3. Flush with clean water. *Optional: if low water flux,* 4. Circulate a low foaming alkaline cleaner for 20 min. 5. Flush with clean water.
Mineral scale	A. 1. Flush with clean water 2. Circulate HNO_3 at 50 °C, pH 4 for 1 hour. 3. Flush with clean water. *Optional:* 4. Repeat Step 2 and leave soaking overnight. 5. Flush with clean water.
	B. 1. Flush with clean water 2. Circulate H_3PO_4 at 50°C, pH 4 for 1 hour. 3. Flush with clean water. *Optional:* 4. Repeat Step 2 and leave soaking overnight. 5. Flush with clean water.
Oil, grease, colloids	A. 1. Flush with clean water 2. Circulate 0.2% Micro® at 50°C, pH 9-10 for 1 hour. 3. Flush with clean water. *Optional:* 4. If iron fouling is suspected, wash with Citric Acid, pH 2-2.5, as noted above.
	B. Substitute alternate detergent cleaners for Micro®. Increase detergent concentration.

(A/G Technology Corporation, pp. 10-12, 1992)

For sanitization, one of the following solutions should be circulated through the system for 30 to 60 minutes after cleaning and rinsing:

- Up to 200 mg/L sodium hypochlorite. If properly cleaned, 10 mg/L should be sufficient.
- Up to 3 percent formalin.
- 100 to 200 mg/L peracetic acid.
- Up to 0.5 Normal sodium hydroxide.
- Up to 70 percent ethanol in water; or
- Autoclave.

All AGT UF and MF cartridges with Housing Sizes 3, 4, 4x2TC, 5, 6, 8, and 9 are autoclavable. Larger UF and MF cartridges that have the suffix "A" in their model number are also autoclavable (e.g., UFP-100-E-35A). The procedure for autoclaving is as follows:

- Wet the cartridge with water by filling through the permeate ports.

- Place the cartridge in an autoclaving bag.

- (Optional) Run autoclave with three pre-vac/steam purge cycles.

- Autoclave at 121 °C for 30 min. Be certain autoclave temperature does not drift above 123 °C. A 30-min cycle is typically sufficient; up to 2 h is acceptable.

- Slow exhaust without vacuum for 30 min.

- Allow SLOW cool down of cartridge, ideal case is overnight. If not possible, allow at least 3 h to cool.

Note: Cartridge life is a function of autoclave temperature, cycle time, and number of cycles. MF and UF cartridges with housing sizes 35A and 55A are guaranteed for two autoclave cycles. Other autoclavable MF cartridges are guaranteed for three cycles and other UF cartridges are guaranteed for two cycles. In practice, strict attention to proper autoclave procedures may permit well over 10 cycles per cartridge. It is prudent to have a spare cartridge available should cartridge life be exceeded or autoclave cycle temperature drift too high. (*AGT Hollow Fiber & TurboTube Ultrafiltration & Microfiltration Cartridge Operating Guide*, OG01/92RO, 1992)

Other considerations when autoclaving:

- Flush preservative glycerol solution from new UF cartridges. At least 10 L/m^2 of membrane area should be permeated without recycle and 10 L/m^2 of membrane area should be flushed through on the reject side without recycle.

- Do not shock cartridges when fully wet or while hot.
- Never expose a warm cartridge to cold process fluids. Always allow cartridge to cool completely prior to shutdown.

8.1.2 *CUNO separations systems.* – CUNO Separations Systems produces a line of polyethersulfone ultrafiltration membranes. They recommend either a 1.5-percent hydrogen peroxide solution in deionized water or a 200 mg/L sodium hypochlorite solution in deionized water. The hydrogen peroxide solution should be circulated through the system for 4 h. Hydrogen peroxide levels should be checked every hour and adjusted as needed to maintain the 1.5-percent concentration. Sodium hypochlorite solution should be circulated for 30 min. The concentration of sodium hypochlorite should be checked after 15 min and more added as needed to maintain a 200 mg/L concentration.

After the sanitizing solution is drained out, the system should be flushed with deionized water for at least 2 h or until trace amounts of solution have reached an appropriate level.

8.1.3 *Desalination Systems, Inc.* – The following cleaners and sanitizers have been evaluated by Desalination Systems, Inc., personnel and are compatible with the DESAL element types listed.

Table 8.2. – Cleaning recommendations for Desalination Systems RO membranes.

CELLULOSE ACETATE RO ELEMENTS	
Foulant	Cleaners/Sanitizers
Mineral Scale Calcium Carbonate Calcium Sulfate Barium Sulfate Strontium Sulfate Iron or Manganese	• HCl, pH 3 • Pfizer Floclean 303 • Citric acid, 2%, adjust to pH 3 with NH_4OH
Organics, Silt, Bacterial Slime	• Trisodium Phosphate, 1% Sodium Tripolyphosphate, 1,% Sodium Dodecyl Sulfate, 0.1% Tetrasodium EDTA, 1% Adjust to pH 9.0 with HCl (See Note) • Pfizer Floclean 307
Bacteria (Sanitizers)	• Sodium Bisulfite, 0.1% • Formaldehyde, 0.1% • Chlorine, 30 mg/L for 30 min

CAUTION: Cleaning at prematurely frequent intervals will hydrolyze the element.
(From DESAL Cleaners/Sanitizers Bulletin E-22, April 1991)

DSI does not approve of using sanitizers containing hydrogen peroxide and peracetic acid such as Renalin, Mincare, and P3-Oxonia for their thin film membranes. These cleaning agents are mainly hydrogen peroxide and peracetic acid. If used improperly, they can destroy a thin film composite membrane instantly. To avoid such severe consequences for operator error, DSI personnel suggest using sodium bisulfite.

Table 8.3. – Cleaning recommendations for Desalination Systems UF membranes.

DESAL-1, DESAL-3, DESAL-5, G-SERIES UF ELEMENTS	
Foulant	Cleaners/Sanitizers
Mineral Scale Calcium Carbonate Calcium Sulfate Barium Sulfate Strontium Sulfate Iron or Manganese	• HCl or Citric Acid, pH 2 • Pfizer IPA 403
Organics, Silt, Bacterial Slime	• Sodium Hydroxide, 0.5% Sodium Dodecyl Sulfate, 0.1% Adjust to pH 11.5 with HCl if necessary. • Trisodium Phosphate, 1% Sodium Tripolyphosphate, 1% Sodium Dodecyl Sulfate, 0.1% Tetrasodium EDTA, 1% Adjust to pH 11.5 with HCl if necessary. • Pfizer IPA 411
Bacteria (Sanitizers)	• Sodium Bisulfite, 0.1% • Formaldehyde, 0.1% • Chlorine Dioxide, 30 mg/L

(From DESAL Cleaners/Sanitizers Bulletin E-22, April 1991)

If Renalin must be used, they highly suggest the following guidelines:

- Maximum strength of 0.1 percent by weight;
- Maximum cumulative exposure time less than 1,000 h over the life of the element;
- Temperature during disinfection must be below 25 °C.

For use of P3-oxonia Active, they recommend:

- Concentration of 1 mL/L once a week;
- Maximum contact time to the membrane of 10 min;
- Temperature should not exceed 25 °C;
- Make up water must be free from transition metals like iron.
(Liberatore, 1992)

8.1.4 *du Pont Permasep Permeators.* – du Pont has a good technical bulletin on cleaning procedures for their hollow fine fiber modules (du Pont Bulletin No. 4040, 1992) that is available on request. The procedures are similar to those listed for A/G Technology Hollow Fiber membranes. The bulletin includes safety precautions, equipment, detergent flushing procedures, procedures for cleaning with citric acid and other solutions, and a diagram of the cleaning process flow. Their table of recommended cleaning solutions for particular foulant problems clearly demonstrates the difficulties involved in selecting a cleaning method for variable water sources (see table 8.5).

Table 8.4 is a list of other compatible cleaning agents that may be used as an alternative to Biz, the original cleaner found effective with RO membranes. These products were found to have no significant effect on membrane fibers at a concentration of 0.25 to 0.50 percent by weight during a 3-mo test. They were not tested for effectiveness (Permasep Engineering Manual Bulletin No. 4040).

Table 8.4. – Cleaning agents compatible with du Pont Permasep permeators.

Biz (17.6% phosphorous)	Procter & Gamble, U.S.A.
Alconox	Alconox Inc., U.S.A.
Alcojet	Alconox Inc., U.S.A.
Drewsperse 738	Drew Chemical, U.S.A.
Decon 90	United Kingdom
Dobanol 91/6	United Kingdom
Tergitol 15/5/7	Union Carbide Co, USA
Biox	Unibound, U.S.A.
Microdetergent	International Product Group, U.S.A.
DMCA-14A	Al Kawther, Saudi Arabia
P.3 Ultrasil 53	Henkel, Germany
P.3 Ultrasil 30 (pH11)	Henkel, Germany
MT 1000	B.F. Goodrich, U.S.A.
MT 2000	B.F. Goodrich, U.S.A.
MT 3000	B.F. Goodrich, U.S.A.
MT 5000	B.F. Goodrich, U.S.A.
Floclean 103	FMC, U.S.A.
Floclean 107	FMC, U.S.A.
Floclean 403	FMC, U.S.A.
Floclean 411	FMC, U.S.A.
Elgalite RF 11	Elga Corp., United Kingdom
Kemazur 2166	Degremont, France
Kemazur 2160	Degremont, France
Kemazur Cleaner	Degremont, France

Table 8.5. – Chemical cleaning agents for du Pont Permasep permeators.

CHEMICALS	FOULANTS					
++ = Preferred + = Effective	Foulants Sparsely Soluble inorganic salts	Metal Oxides	Inorganic Colloids	Biological Matter	Organics	Silica
0.5% (V) HCl, pH 2.3 min. [1]	++		+			
0.5% (W) H_3PO_4, pH 2.3 min. [1]	+	+				
0.2% (W) NH_2SO_3H pH 2.3 min. [1]	+	+				
2.0% (W) Citric Acid pH 4 (w/NH_4OH)	+	+	+			
2.0% (W) Citric Acid pH 8 (w/NH_4OH)	+					
2.0% (W) Citric Acid 2.0% (W) Na_2EDTA pH 4.0 (w/NH_4OH)	+	+				
1.0% (W) STP 1.0% (W) TSP 1.0% (W) Na_2EDTA				+	+	
1.0% (W) Na_2EDTA 0.1% (W) NaOH pH 11.0 max. [2]				++	+	+
1.0% (W) $Na_2S_2O_4$	+	++				
0.3% (W) Na Perborate 0.25% (W) Na-DBS pH 10.0			++	+	+	+
2.0% (W) Na STP 0.25% (W) Na-DBS pH 10.01				+	+	+
1.0% (W) NaHMP		+	+		+	
0.5% (V) NaOH pH 11.0 max [2]						++

1. The pH in the permeator must be ≥4.0 for continuous operation. The maximum allowable exposure time during operation and cleaning for solutions with pH 2.3 to 4.0 is 100 h over the life of the permeators. Permeators must not be exposed to pH <2.3.

2. The pH in the B-10 permeator must be ≤9.0 for continuous operation. For cleaning, the permeator can be exposed to pH 9.0 to 11.0 for an unlimited period of time. For the B-9 permeator for continuous operation the pH must be ≤11.0. The maximum total allowable exposure time during cleaning for solutions with pH 11.0 to 9.0 is 500 h over the life of the permeator. Permeators must not be exposed to pH >11.9.

Abbreviations: HCl = Hydrochloric Acid; H_3PO_4 = Phosphoric Acid; NH_2CO_3H = Sulfamic Acid; Citric Acid is $C_3H_4(OH)(CO_2H)_3$; NH_4OH is Ammonium Hydroxide; Na_2EDTA = Disodium Salt of Ethylenediamine Tetraacetic Acid; Na-DBS = Sodium Salt of Dodecylbenzene Sulfonic Acid; Na-perborate = $NaBO_3 \cdot 4H_2O$; STP = Sodium Triphosphate $Na_5P_3O_{10}$; TSP = Trisodium Phosphate $Na_3PO_4 \cdot 12H_2O$.

8.1.5 *Filmtec.* – Filmtec has a variety of recommended cleaning solutions for the FT30 aromatic polyamide TFC membranes. Table 8.6 is from their technical bulletin on *Cleaning Procedures for Filmtec FT30 Elements.* Table 8.7 is a relative rating of the different cleaning solutions for various purposes. Cationic surfactants are not recommended for Filmtec elements.

Table 8.6. – Cleaning solutions for the FT30 membrane.

SOLUTION	COMPOSITION	SOLUTION	COMPOSITION
A.	0.1% NaOH 0.1% Na·EDTA pH 12, 30°C max	E.	0.5% H_3PO_4
B.	0.1% NaOH 0.05% Na·DSS pH 12, 30°C max	F.	2.0% Citric Acid
C.	1.0% STP 1.0% TSP 1.0% Na·EDTA	G.	0.2%NH_2SO_3H
D.	0.2% HCl	H.	1.0% $Na_2S_2O_4$

Percents are by weight of active ingredient. Chemical compounds: NaOH is Sodium Hydroxide; Na·EDTA is the tetra-sodium salt of ethylene diamine tetraacetic acid; Na·DSS is the sodium salt of dodecylsulfate; STP is sodium triphosphate ($Na_5P_3O_{10}$); TPS is trisodium phosphate ($Na_3PO_4 \cdot 12H_2O$); HCl is hydrochloric acid; H_3PO_4 is phos-phoric acid; citric acid is $C_3H_4(OH)CO_2H)_3$; NH_2SO_3H is sulfamic acid; $Na_2S_2O_4$ is sodium hydrosulfite. (*Technical Bulletin - FILMTEC® Membranes – Cleaning Procedures for FILMTEC® FT30 Elements*, 7/91.)

8.1.6 *Fluid Systems.* – Fluid Systems recommends the following cleaning solutions for temperatures up to 40 °C (104 °F) for 45 min at the appropriate rate for the tube diameter.

2.5 in	0.7 m^3/h/tube (3 gal/min)
4 in	2.3 m^3/h/tube (10 gal/min)
6 in	4.6 m^3/h/tube (20 gal/min)
8 in	9.1 m^3/h/tube (40 gal/min)

The system pressure should be the minimum required to achieve the specified flow but the pressure drop across any vessel should not exceed 414 kPa. Pressure tubes should only be cleaned in parallel so that the reject from one tube is not used to clean another. If stages of pressure tubes are used, each stage must be cleaned separately. Contact the Technical Services Department at Fluid Systems if cleaning does not restore system to expected capacity or water quality, or for advice on alternative cleaning solutions.

Table 8.7. – Performance of recommended cleaning solutions for different types of fouling.

	A.	B.	C.	D.	E.	F.	G.	H.
Slightly Soluble Inorganic Salts				BEST	OK	OK	OK	
Metal Oxides (iron)					GOOD		OK	GOOD
Inorganic Colloids (Silt)		GOOD						
Silica	OK							
Biofilms	BEST	GOOD	GOOD					
Organics	OK	GOOD	GOOD					

Examples of slightly soluble inorganic salts are $CaCO_3$, Calcium Carbonate; $CaSO_4$, Calcium Sulfate; and $BaSO_4$, Barium Sulfate. Consult a FILMTEC representative if a more effective cleaner is needed for silica. (*Technical Bulletin – FILMTEC® Membranes – Cleaning Procedures for FILMTEC® FT30 Elements,* 7/91.)

8.1.6.1 *Polyamide TFCL elements.* Polyamide TFCL elements may be cleaned with anionic or non-ionic surfactants. Cationic surfactants are not recommended because they may cause irreversible fouling. Fluid Systems recommends two cleaning solutions that must be prepared with water free of residual chlorine or other oxidizing agents:

Solution L1: used to remove acid soluble substances such as metal hydroxides and calcium carbonate. It should generally be used before using Solution L2 if both solutions are used. Per cubic meter of water (264.2 gal)

1. Citric acid	10.0 kg (22.1 lb)
2. Adjust pH to 2.5 with NH_4OH (ammonium hydroxide)	

Solution L2: used to remove organic substances and microbiological slimes.
Per cubic meter of water (264.2 gal)

1. STPP (Sodium tripolyphosphate)	10.0 kg (22.1 lb)
2. Sodium salt of EDTA (Ethylenediaminetetraacetic Acid)	
as powder	10.0 kg (22.1 lb)
as 39-percent solution	25.6 kg (56.5 lb)
3. TSP (Trisodium phosphate)	10.0 kg (22.1 lb)
4. Adjust to a pH of 10 to 11 with HCl (hydrochloric acid.)	

(From Fluid Systems *Cleaning Instructions for TFCL Elements*, 1992)

8.1.6.2 *Polyether urea TFC membranes.* Polyether urea TFC elements may be cleaned with cationic or non-ionic surfactants. Anionic surfactants are not recommended because they may cause irreversible fouling. Fluid Systems recommends two cleaning solutions, Y and Z, that must be prepared with water free of residual chlorine or other oxidizing agents. RO permeate should be used for seawater application, but treated feedwater may be used for brackish water applications.

Solution Y: used to remove acid soluble substances such as metal hydroxides and calcium carbonate. It should generally be used before using Solution Z if both solutions are used.

Per cubic meter of water (264.2 gal)

1. Citric acid	10.0 kg (22.1 lb)
2. Adjust pH to 4.5 with NH_4OH (ammonium hydroxide)	

Solution Z: used to remove organic substances and microbiological slimes.
Per cubic meter of water (264.2 gal)

1. Borax	10.0 kg (22.1 lb)
2. Sodium salt of EDTA (Ethylenediaminetetraacetic Acid)	
as Powder	10.0 kg (22.1 lb)
as 39 percent solution	25.6 kg (56.5 lb)
3. TSP (Trisodium phosphate)	10.0 kg (22.1 lb)

4. The pH of this solution will be in the 10 to 11 range; it may have to be adjusted to the recommended cleaning pH range for TFC elements of 4.5 to 11.

Formaldehyde may be added to either cleaning solution to aid in disinfection or removal of biological growths. Fluid Systems recommends a 0.5- to 1-percent solution that is 13.5 to 27.0 L/m^3 of water (3.6 to 7.2 gal/264.2 gal).
(Fluid Systems *Cleaning Instructions for TFC Elements*, 1991)

8.1.6.3 *Cellulosic ROGA elements.* Cellulosic membranes have more tolerance to oxidants than the thin film composites. The recommended limit is 1 mg/L chlorine residual or the equivalent. Treated feedwater or RO permeate are recommended for preparing cleaning solutions. The following are recommended by Fluid Systems:

Solution A: Used to remove acid soluble substances such as metal hydroxides and calcium carbonate. It should generally be used before using **Solution B** if both solutions are used.

Per cubic meter of water (264.2 gal)

1. Citric acid	20.0 kg (44 lb) **or**
Phosphoric acid 75 percent	55 L (14.7 gal)
2. Triton X-100	1 L
3. Adjust pH to 2.5 with NH_4OH (ammonium hydroxide)	

Solution B: used to remove organic substances and microbiological slimes.

Per cubic meter of water (264.2 gal)

1. TSP (Trisodium phosphate)	20.0 kg (44 lb)
2. Sodium salt of EDTA (Ethylenediaminetetraacetic Acid)	
as powder	8.0 kg (17.5 lb)
as 39 percent solution	20.0 kg (56.5 lb)
3. Triton X-100	1 L

4. Adjust pH with H_2SO_4 (sulfuric acid) or HCl (hydrochloric acid.) as follows after mixing chemicals with product water.

HR elements should be cleaned at 7.0.
SD, S, and LP elements should be cleaned at a pH of 8.0.

Formaldehyde may be added to either cleaning solution to aid in disinfection or removal of biological growths. Fluid Systems recommends a 0.5-to 1-percent solution that is 13.5 to 27.0 L/m$_3$ of water (3.6 to 7.2 gal/264.2 gal). (From Fluid Systems *Cleaning Instructions for ROGA Elements*, 1990)

8.1.7 *Hydranautics.* – Hydranautics has a set of recommended cleaning solutions for each membrane type. The general procedure is as follows:

- Mix a fresh batch of cleaning solution in the cleaning tank using clean permeate water.

- Circulate the cleaning solution through the pressure tubes for approximately 1 h the desired period of time, at a flow rate of 133 to 150 L/min (35 to 40 gal/min) per pressure tube for 8.0- and 8.5-in. pressure tubes or 34 to 38 L/min (9 to 10 gal/min) for 4.0-in. pressure tubes.

- After completion of cleaning, drain and flush the cleaning tank; then fill the cleaning tank with clean permeate water for rinsing.

- Rinse the pressure tubes by pumping clean product water from the cleaning tank (or equivalent source) through the tubes to the drain for several minutes.

- After the RO system is rinsed, operate it with the product dump valves open until the product water flows clean and is free of any foam or residues of cleaning agents (usually 15 to 30 min).

8.1.7.1 *Cellulose acetate blend RO membranes.* Cellulose acetate blend membranes will be severely damaged if exposed to high or low pH conditions for a prolonged period. If the RO system is to be out of service for more than 1 day, the pH of the rinse water should be adjusted to approximately 5.6.

Table 8.8. – Cellulose acetate blend RO membrane element foulant symptoms.

Foulant	General symptoms	Response
1. Calcium Precipitates (carbonates and phosphates, generally found at the concentrate end of the system)	A marked decrease in salt rejection and a moderate increase in ΔP between feed and concentrate. Also, a slight decrease in system production.	Chemically clean the system with Solution 1.
2. Hydrated Oxides (iron, nickel, copper, etc.)	A rapid decrease in salt rejection and a rapid increase in ΔP between feed and concentrate. Also, a rapid decrease in system production.	Chemically clean the system with Solution 1.
3. Mixed Colloids (iron, organics, and silicates)	A slight decrease in salt rejection and a gradual increase in ΔP between feed and concentrate. Also, a gradual decrease over several weeks in system production.	Chemically clean the system with Solution 2.
4. Calcium Sulfate (generally found at the concentrate end of the system)	A significant decrease in salt rejection and a slight to moderate increase in ΔP between feed and concentrate. Also, a slight decrease in system production.	Chemically clean the system with Solution 2.
5. Organic Deposits	Possible decrease in salt rejection and a gradual increase in ΔP between feed and concentrate. Also, a gradual decrease in system production.	Chemically clean the system with Solution 2. For heavy fouling use Solution 3.
6. Bacteria Fouling	Possible decrease in salt rejection and a marked increase in ΔP between feed and concentrate. Also, a marked decrease in system production.	Chemically clean the system with either of the solutions, depending on possible compounded fouling.

Table 8.9. – Summary of recommended cleaning solutions for Hydranautics cellulose acetate blend RO membranes.

Solution	Ingredient	Quantity per 100 gal (379 L)	pH Adjustment
1	Citric Acid	7.7 kg (17.0 lb)	Adjust to pH 4.0 with sodium hydroxide (NaOH)
	RO Permeate	379 L (100 gal)	
2	Sodium Laurel Sulfate or Triton X-100	1.0 kg (2.2 lb) 380 mL (0.1 gal)	Adjust to pH 10.0 with sulfuric acid (H_2SO_4)
	RO Permeate	379 L (100 gal)	

Note: Ensure that the pH in any cleaning solution does not fall below 2.0. Otherwise, damage to the RO membrane elements may occur, particularly at elevated temperatures. The maximum pH should be less than 10.0. Use ammonium hydroxide to raise the pH, or sulfuric or hydrochloric acid to lower it.

(From *Hydranautics Technical Service Bulletin TSB 100.00,* April 1992)

8.1.7.2 *Composite polyamide RO membranes.* Composite polyamide membranes are not chlorine tolerant. Cleaning solutions and rinse water must be chlorine-free. Cationic surfactants should not be used in cleaning solutions because irreversible fouling of the membranes may occur.

Table 8.10. – Composite polyamide RO membrane element foulant symptoms.

Foulant	General symptoms	Response
1. *Calcium precipitates* (carbonates and phosphates, generally found at the concentrate end of the system)	*A marked decrease in salt rejection and* a moderate increase in ΔP between feed and concentrate. Also, a slight decrease in system production.	*Chemically clean the system* with Solution 1.
2. Hydrated oxides (iron, nickel, copper, etc.)	A rapid decrease in salt rejection and a rapid increase in ΔP between feed and concentrate. Also, a rapid decrease in system production.	Chemically clean the system with Solution 1.
3. Mixed colloids (iron, organics, and silicates)	A slight decrease in salt rejection and a gradual increase in ΔP between feed and concentrate. Also, a gradual decrease over several weeks in system production.	Chemically clean the system with Solution 2.
4. Calcium sulfate (generally found at the concentrate end of the system)	A significant decrease in salt rejection and a slight to moderate increase in ΔP between feed and concentrate. Also, a slight decrease in system production.	Chemically clean the system with Solution 2.
5. Organic deposits	A marked decrease in salt rejection and a gradual increase in ΔP between feed and concentrate. Also, a gradual decrease ,in system production.	Chemically clean the system with Solution 2. For heavy fouling use Solution 3.
6. Bacteria fouling	A marked decrease in salt rejection and a marked increase in ΔP between feed and concentrate. Also, a marked decrease in system production.	Chemically clean the system with Solution 4.

Note: All problems require the cause of the fouling to be corrected. Contact Hydranautics for assistance.
(From *Hydranautics Technical Service Bulletin TSB 107.00,* April 1992)

Table 8.11. – Summary of recommended cleaning solutions for Hydranautics composite polyamide RO membranes.

Solution	Ingredient	Quantity per 379 L (100 gal)	pH adjustment
1	Citric acid	7.7 kg (17 lb)	Adjust to pH 4.0 with sodium hydroxide (NaOH)
	RO permeate (chlorine-free)	379 L (100 gal)	
2	Sodium tripolyphosphate	7.7 kg (17 lb)	Adjust to pH 10.0 with sulfuric acid (H_2SO_4
	Tetrasodium EDTA (Versene 220 or equal)	3.18 kg (7 lb)	
	RO permeate (chlorine-free)	379 L (100 gal)	
4	Sodium tripolyphosphate	7.7 kg (17 lb)	Adjust to pH 10.0 with sulfuric acid (H_2SO_4
	Sodium dodecylbenzene-sulfonate	0.91 kg (2.13 lb.)	
	RO permeate (chlorine-free)	379 L (100 gal)	

(From *Hydranautics Technical Service Bulletin TSB 107.00,* April 1992)

8.1.7.3 PVD (Polyvinyl derivative) RO membranes. Use of cationic or nonionic surfactants with Hydranautics PVD membranes may result in flux decline.

Table 8.12. – Polyvinyl derivative RO membrane element foulant symptoms.

Foulant	General symptoms	Response
1. Calcium precipitates (carbonates and phosphates, generally found at the concentrate end of the system)	A marked decrease in salt rejection and a moderate increase in ΔP between feed and concentrate. Also, a slight decrease in system production.	Chemically clean the system with Solution 1.
2. Hydrated oxides (iron, nickel, copper, etc.)	A rapid decrease in salt rejection and a rapid increase in ΔP between feed and concentrate. Also, a rapid decrease in system production.	Chemically clean the system with Solution 1.
3. Organic deposits	Possible decrease in salt rejection and a gradual increase in ΔP between feed and concentrate. Also, a gradual decrease in system production.	Chemically clean the system with Solution 2.
4. Bacteria fouling	Possible decrease in salt rejection and a marked increase in ΔP between feed and concentrate. Also, a marked decrease in system production.	Chemically clean the system with Solution 2.

Note: All problems require the cause of the fouling to be corrected. Contact Hydranautics for assistance.
(From *Hydranautics Technical Service Bulletin TSB 102.00,* April 1992)

Table 8.13. – Summary of recommended cleaning solutions for Hydranautics polyvinyl derivative RO membranes.

Solution	Ingredient	Quantity per 379 L (100 gal)	pH adjustment
1	Citric acid	7.7 kg (17.0 lb)	Adjust to pH 4.0 with sodium hydroxide (NaOH)
	RO permeate	379 L (100 gal)	
2	Sodium laurel sulfate or Triton X-100	1.0 kg (2.2 lb) 380 mL (0.1 gal)	Adjust to pH 10.0 with sulfuric acid (H_2SO_4)
	RO permeate	379 L (100 gal)	

Note: Ensure that the pH in any cleaning solution does not fall below 2.0. Otherwise, damage to the RO membrane elements may occur, particularly at elevated temperatures. The maximum pH should be less than 10.0. Use ammonium hydroxide to raise the pH, or sulfuric or hydrochloric acid to lower it.

(From *Hydranautics Technical Service Bulletin TSB 102.00,* April 1992)

8.1.8 *Ionpure Technologies Corporation.* – Ionpure, formerly Millipore's Water Products Division, provides regular and emergency membrane cleaning services as part of their maintenance agreement, so they do not include cleaning procedures in their catalog of products and specifications. They did send a paper by Janet L. White entitled "Water System Sanitization" (presented at the High Purity Water II Seminar, ISPE EXPO'92, 5/8/92). Table 8.14 lists sanitization procedures for membranes supplied by Ionpure.

8.1.9 *Koch Membrane Systems.* – Cleaning is a daily part of the operation cycle for Koch hollow fiber ultrafiltration membranes when they are used for RO pretreatment. They are back-flushed once each hour, and chemically cleaned once a week, depending on the feed stream quality. A detailed outline of cleaning and sanitization procedures and reagents is included in their specification booklet and also in tables 8.15 through 8.17.

Table 8.14. – RO and UF membrane sanitization chemicals.

Sanitization agent	Polyamide thin film composite RO	Polysulfone thin film composite RO	Cellulose acetate	Polysulfone UF
Peracetic acid (CH_3CO_3H)	400 mg/L typical[1], up to 2000 mg/L 2 h contact @ 400 mg/L	400 mg/L 2 h contact pH 4	400 mg/L 2 h contact pH 4.5	up to 2000 mg/L 2 h contact @ 400 mg/L
Hydrogen peroxide (H_2O_2)	up to 2500 mg/L[1] 2-12 h contact pH 3-4	not compatible	up to 2000 mg/L 2 h contact pH 4.5	3%[2] 1-2 h contact pH 3-4
Sodium hypochlorite[3] (NaOCl)	not compatible	up to 100 mg/L Cl_2 1-2 h contact pH 6.0-7.5	up to 10 mg/L Cl_2 1-2 h contact pH 6.0-7.0	up to 500 mg/L 1 h contact pH 6.0-7.5
Formaldehyde (HCHO)	0.5-3%[4] 0.5-2 h contact	1-4% 0.5-2 h contact	1-4% 0.5-2 h contact	1-4% 0.5-2 h contact

[1] Temperature must not exceed 25 °C or membrane degradation may occur. Also, iron or other heavy metals catalyze a membrane degradation reaction. Clean with hydrochloric acid first if present.

[2] Iron or other heavy metals catalyze a membrane degradation reaction. Clean with hydrochloric acid first if present.

[3] Do not use pH below 5.5 because chlorine gas will be liberated.

[4] Do not use formaldehyde with polyamide thin film composite membranes until cartridges have been operated for at least 6 h or severe flux loss may occur.

Table 8.15. – Sanitizing procedures for Koch hollow fiber membranes.

Sanitizing Agents	
HF 132-20-GM80	HF132-20-PM10
200 mg/L NaOCL or	200 mg/L NaOCL or
5% Hydrogen peroxide	5% Hydrogen peroxide or
---	95°C Water

RECOMMENDED PROCEDURE FOR SANITIZATION AT 95 °C OF HF132-20-PM10

1. Continue to permeate.
2. Reduce the feed pressure to 35 kPa.
3. Heat inlet water up to 95 °C at a rate of approximately 2 °C/min.
4. Continue to permeate water to drain for 30 min.
5. Reduce temperature to ambient at a rate of approximately 2 °C/min.

8.1.10 *Rhône Poulenc.* – Rhône Poulenc's Kerasep™ monolithic ceramic microfiltration membranes are the most durable of any membranes considered in this manual. They can withstand 5,000 kPa, a pH range from 0 to 14, and process temperatures up to 400 °C. They are not affected by oxidants or solvents and can be steam cleaned at 125 °C for 40 min. Most any cleaning solution recommended by the other membrane producers would work for these membranes as well. Their zirconia/titanium oxide ultrafiltration membranes, which operate at pressures up to 1,000 kPa, can also tolerate high temperatures and extreme pH and oxidizing conditions.

Table 8.16. – *Cleaning recommendations for Koch hollow fiber membranes.*

Cleaning Frequency			
Application	Fast/back flush	Chemical cleaning	Variability
Pretreatment	1 × h	1 × d	High
Post DI	1 × d	1 × week	Moderate
Post RO DI	1 × week	1 × y	Low
Cleaning Agents			
Fouling type		Reagents	
General cleaning, i.e. organics and bacteria		- 200 mg/L NaOCL	
Heavy organic fouling		- 0.5% NaOH	
Iron fouling		- 0.5% Citric acid + 0.2% Na EDTA - 0.5% Oxalic acid	
Silica fouling		- 0.5% NaOH at 27 °C	
Cleaning Cycle			
Mode		Duration (min.)	
Fast flush with chemical		1 - 5	
Soak		0 - 30	
Upper backflush		1 - 10	
Lower backflush		1 - 5	
Fast flush without chemical		1 - 5	
Pre-production rinse		As required	

Table 8.17. – *Chemical resistance of Koch hollow fiber membranes.*

Chemical Resistance		
Both the Romicon GM 80 and PM 10 hollow fiber membranes show exceptional resistance to many chemicals. For example, at 25 °C, these fibers have strong resistance to the following chemicals:		
10 % Acetic acid	25 % Ethanol	1 % Sulfuric acid
5 % Ammonium Hydroxide	1 % Hydrogen Peroxide	5 % Phosphoric acid
1 % Citric acid	5 % Hydrochloric acid	1 % Sodium Hydroxide
5 % Diethanolamine	1 % Nitric acid	.04 % Sodium Hydrochlorite
10 % EDTA-Na	1 % Oxalic acid	1 % Sodium Azide

8.2 Storage Recommendations

Whenever a membrane system is to be shut down for an extended time, certain storag procedures must be followed. If the time is a week or less, they are usually as simple a filling the system with permeate or treated feed water and then changing it when ever th water temperature exceeds a limit specified by the manufacturer. In warm climates, o where the system is exposed to sunlight, the water may have to be continuously recirculated

For extended storage periods, the cellulosic and synthetic elements should be filled with preservative to control biological growth and prevent freezing or overheating. Som membranes, such as AGT's ultrafilters and Desalination Systems membranes, may b reglycerized and stored dry. Ceramic and metallic membranes and some polysulfone tubula filters can be merely cleaned and stored dry. Table 8.18 outlines recommendations on storag times, temperatures, and preservatives from a selection of manufacturers.

Table 8.18. – Short- and long-term membrane storage procedures.

Company/ product	Limit for short-term storage	Flush water	Repeat cycle	Temp. limit	Long-term storage caveats	Water requirements	Preservative solution	Repeat cycle
Fluid Systems/ TFCL,TFCS	7 days	Oxidant-free permeate	2 days	45 °C	48 h in operation, Temp <45 °C, no direct sunlight	Oxidant-Free Permeate	0.5 to 1.0% formaldehyde or 0.5 to 1.0% sodium metabisulfite	30 days
Fluid Systems/ TFC	7 days	Oxidant-free permeate	2 days	45 °C	Temp <45 °C, no direct sunlight	Oxidant-Free Permeate	0.1 to 0.2% glutaraldehyde adjust pH to 5.0 to 8.5 w/NaOH, no ammonia compounds	As needed to maintain 0.05% glutaraldehyde
Fluid Systems/ ROGA	7 days	Water at pH 5.5 ± 0.5 with chlorine residual at 0.1 to 0.5 mg/L	2 days if >20 °C else 7 days	--	Temp <30 °C, no direct sunlight	--	0.1 to 0.2% glutaraldehyde adjust pH to 5.5 ± 0.5 w/NaOH, no ammonia compounds	As needed to maintain 0.05% glutaraldehyde
FILMTEC/ FT30,NF70,NF40	7 days	1% Sodium bisulfite solution w/chlorine residual <0.1 mg/L	7 days	45 °C	No caveats	Chlorine residual <0.1 mg/l	18% propylene glycol + 1.0% sodium metabisulfite	Monitor to ensure pH and temperature limits are not exceeded
CUNO Separations Sys./UF	2 days	Sanitize and flush with DI water	Repeat before start-up	50 °C	Sanitize first	Deionized	0.5% hydrogen peroxide or 20 mg/L sodium hypochlorite	Monitor to maintain concentrations
A/G Technology Ultrafilters	2 weeks	Thoroughly cleaned, and rinsed			Sanitize with 200 mg/L Sodium hypochlorite solution before returning to operation		5 to 10 mg/L chlorine, 0.1 N NaOH, up to 3% formalin, 30% ethanol in water, or up to 1% sodium azide	Monitor to ensure membranes remain wetted

Table 8.18. – Short- and long-term membrane storage procedures – Continued.

Company/ product	Limit for short-term storage	Flush water	Repeat cycle	Temp. limit	Long-term storage caveats	Water requirements	Preservative solution	Repeat cycle
Hydranautics - Cellulose Acetate	5 days	pH 5.5 ± 0.5 free chlorine residual of 0.1 to 0.5 mg/L at brine discharge	2 days		Flush at low pressure 1 h then high pressure for 5 to 10 min before returning to service.	pH 5.5 ± 0.5	Approved biocide	30 days at <27 °C, 5 days at >27 °C
Hydranautics - Polyamide	5 to 30 days	Purge gases from system with feed water, close system when filled	5 days		Clean and sanitize first	No chlorine	Approved biocide	30 days at <27 °C, 15 days at >27°C

9. WASTE DISPOSAL

Waste disposal from membrane systems is a major consideration. Typical waste products include the reject brine, pretreatment sludge, cleaning and storage solutions, even the used membranes themselves. Ways to minimize or avoid sludge production were discussed in the chapter on pretreatment. Cleaning and storage chemicals each have their own use and disposal problems, though some are more innocuous than others. Currently, used membranes are treated as solid waste, but possibilities for reconditioning do exist. The last problem must be addressed no matter what precautions are taken to minimize waste, and that is – what to do with the reject water.

The major difficulty is that the reject water is more concentrated than it was before membrane separation. The reject water may be in violation of effluent limits imposed by Federal and State Regulations. The legal procedure is to have the concentrate evaluated by the local health department for determination on whether it is "conventional," "nonconventional," or toxic waste. Conventional discharges contain BOD (Biological Oxygen Demand), coliform bacteria, suspended solids, pH, oil, and grease. Toxic discharges contain substances regulated as toxins under the Toxic Substances Control Act as "presenting an unreasonable risk of injury to health or the environment" (15 U.S.C. §2601(a)(3)). Nonconventional discharges are everything else, including some RO concentrate waters. In some States, Florida for instance, desalting process waste brine is classified as industrial waste and must therefore meet more stringent disposal requirements than municipal water treatment waste (Conlon, 1990).

All discharges into the navigable waters of the United States (all streams, lakes, rivers, and the ocean) are required to be permitted through the NPDES (National Pollution Discharge Elimination System) program. Toxic and nonconventional discharges are required to be treated with the "Best Available Technology" as determined by the EPA. A monitoring system is required to supply information for detailed reports to the EPA on the content and volume of the effluent. Further local restrictions may exist. The State Health Department is usually in charge of determining what permits are required.

The method generally accepted for determining the expected concentration of the brine assumes 100 percent rejection of contaminants. The concentration factor is calculated by the following equation.

$$CF = \frac{1}{1 - Y}$$

Y equals the percent recovery in decimal. By this equation, the concentration of the brine from a 75-percent recovery system purifying water with 3,000 mg/L TDS would be 12,000 mg/L, four times that of the feed water. At 90-percent recovery, the concentrate would have 30,000 mg/L. A 10 million m^3/d plant will produce 13.3 million m^3/d of concentrate at 75 percent and 11 million m^3/d at 90 percent. Where the concentrate can be disposed of depends on what specific contaminants are in the water and potential downstream impacts such as water treatment plants. Several general possibilities exist:

1. Dispose into a nearby brackish body of water or the ocean;
2. Concentrate in lined evaporation ponds or with mechanical brine concentrators;
3. Use disposal wells to inject into a brackish aquifer not being used for drinking water;
4. If TDS is not too high, use the concentrate for beneficial use such as irrigation water for salt tolerant plants;
5. Use brine in solar ponds for power generation.

The following sections examine these possibilities for coastal and inland plants in greater detail.

9.1 Coastal Desalination Plants

Most coastal desalting plants return waste brine to the sea. If the feed water was pumped from a brackish well, the TDS of the brine will be less than or close to the TDS of seawater. Coastal marine life is adapted to periodic changes in salinity, so theoretically, should experience little detrimental effect unless heavy metals are concentrated in the brine. However, if seawater is used for desalting, the salinity of the brine can be over 50 percent higher than ambient salinity. No existing evidence or studies indicate any ill effects caused by increases in salinity on coastal marine life. A call to the local seawater aquarium specialist revealed that marine organisms can tolerate gradual increases in salinity, but sudden changes are fatal. A specific gravity of 1.030 g/cm^3 is the upper limit of tolerance for most species outside of the Mediterranean. Mobile creatures can move away from the brine inlet, but sedentary species will die if the salinity suddenly increases too much. Assuming a linear relationship between specific gravity and concentration, an increase of 0.008 g/cm^3 (normal specific gravity = 1.022 for seawater) equates to an increase in salinity of only 36 percent (from a normal 35,000 to 47,727 mg/L). If good circulation is present at the brine outfall, the high salinity should dissipate rapidly, but discharging large volumes of high salinity water into enclosed bays, shellfish beds, or valuable fishing waters might pose some hazards.

An alternative to discharging seawater RO brine directly into the ocean is to combine it with sewage treatment plant effluent or storm water runoff. The TDS of sewage treatment plant effluent is much lower than that of seawater RO brine. If the brine is mixed with 1000 mg/L TDS water (probably high) at a ratio of 2:1 (brine:low TDS), the salinity can be reduced to that of the ambient seawater. Another solution may be to let the brine discharge through a seabed distribution system of pipelines, or one pipeline outfall with many outlets.

9.2 Inland Desalination Plants

The design of inland desalination plants has to be more sensitive to the disposal of concentrate. If it cannot be discharged locally into brackish surface water, possibilities exist for irrigation, deep well injection, or further concentration of the salts for beneficial use.

9.2.1 *Discharge into surface waters.* – Depending on its composition, inland desalting concentrates may be discharged into waterways or wetlands leading to the ocean. A NPDES permit is required for all point sources discharging into navigable waters. The discharge must meet the enforceable MCL (maximum contaminant levels) promulgated under the Safe Drinking Water Act 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 9.1 lists the SDWA regulated contaminants. The MCLG (Maximum Contaminant Level Goals) are recommended, but not enforceable, levels of contamination.

Local regulations must also be met. For instance, 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 against hydrogen sulfide, low dissolved oxygen levels, sulfide toxicity, low pH, corrosiveness and toxins (Malaxos and Morin, 1990; Colorado State Dept. of Health, personal communication, 1992).

The effect of discharge of RO or nanofiltration concentrates on aquatic life depends on the composition of the discharge and that of the receiving body of water. If the waterway is subject to periodic influxes of brackish water from the ocean, aquatic life should experience minimal impact there. In any case, impact studies may be required before a NPDES permit is issued.

Table 9.1. – Safe Drinking Water Act regulated contaminants: 1992.

Contaminants	MCLG (mg/L)	MCL (mg/L)	Contaminants	MCLG (mg/L)	MCL (mg/L)
Fluoride*	4.0	4.0			
Volatile Organics			**Added Phase II Contaminants**		
Trichloroethylene	zero	0.005	Ethylbenzene	0.7	0.7
Carbon Tetrachloride	zero	0.005	Styrene	0.1	0.1
1,1,1-Trichloroethane	0.2	0.2	Heptachlor	zero	0.0004
1,2-Dichloroethane	zero	0.005	Heptachlor epoxide	zero	0.0002
Vinyl Chloride	zero	0.002	Nitrite	1	1
Benzene	zero	0.005	Aldicarb sulfoxide	0.001	0.004
p-Dichlorobenzene	0.075	0.075	Aldicarb sulfone	0.001	0.002
1,1-Dichloroethylene	0.007	0.007	**Lead and Copper**		
Chloroform & Surface Water Treatment			Lead*	zero	TT+
Total Coliform*	zero	<5%	Copper	0.07	TT++
Turbidity*	N/A		**Phase V**		
Giardia lamblia	zero	TT	Dichloromethane	zero	0.005
Viruses	zero	TT	1,2,4-Trichlorobenzene	0.07	0.07
Standard plate count	N/A	TT	'Hexachlorobenzene	zero	0.001
Legionella	N/A	TT	Antimony	zero	0.006
Phase II			Nickel	0.1	0.1
Tetrachloroethylene	zero	0.005	Thallium	0.0005	0.002
Chlorobenzene	0.1	0.1	Beryllium	0.004	0.004
trans-1,2-Dichloroethylene	0.1	0.1	Cyanide	0.2	0.2
cis-1,2-Dichloroethylene	0.07	0.07	Endrin*	0.002	0.002
'o-Dichlorobenzene	0.6	0.6	Dalapon	0.2	0.2
Barium*	2	2	Diquat	0.1	0.1
Cadmium*	0.005	0.005	Endothall	0.1	0.1
Chromium* (total)	0.1	0.1	Glyphosphate	0.7	0.7
Mercury* (inorganic)	0.002	0.002	Adipates		
Nitrate*	10	10	(diethylphenyladipate)	0.5	0.5
Selenium*	0.05	0.05	2,3,7,8-TCDD (Dioxin)	zero	3x10^{-8}
Asbestos (fiber >10µm/L)	7MFL	7MFL	1,1,2-Trichloroethane	0.003	0.005
Lindane*	0.0002	0.0002	Vydate	0.2	0.2
Methoxychlor*	0.04	0.04	Simazine	0.004	0.004
Toxaphene*	zero	0.003	PAHs (benzo(a)pyrene)	zero	0.0002
2,4-D*	0.07	0.07	Phthalates		
2,4,5-TP*	0.05	0.05	[diethylhexylphthalate]	zero	0.004
Aldicarb	0.001	0.003	Picloram	0.5	0.5
Chlordane	zero	0.002	Dinoseb	0.007	0.007
Carbofuran	0.04	0.04	Hexachlorocyclopentadiene	0.05	0.05
Alachlor	zero	0.002	**Radionuclides (Proposed)**		
Epichlorohydrin	zero	TT	Radium 226*	zero	20pCi/L
Toluene	1	1	Radium 228*	zero	20pCi/L
PCBs	zero	0.0005	Beta particle and photon		
Atrazine	0.003	0.003	radioactivity	zero	4mrem/yr
Acrylamide	zero	TT	Uranium	zero	0.02
Dibromochloropropane (DBCP)	zero	0.0002	Gross alpha particle		
1,2-Dichloropropane	zero	0.005	activity	zero	15pCi/L
Pentachlorophenol	zero	0.001	Radon	zero	300pCi/L
Ethylene dibromide (EDB)		0.0000510	**Sulfate (Proposed)**		
Xylenes (total)	10		Sulfate	400/500	400/500
Removed from Initial Phase II List			**Arsenic (Interim)**		
Silver*			Arsenic*	0.05	0.05
Aluminum			**Disinfection By-Products**		
Molybdenum					
Vanadium			Total Trihalomethanes (Interim)	0.01	
Sodium					
Zinc					
Dibromomethane					

' Not on list of 83.

* Indicates original contaminants with interim standards that have or will be revised.

\+ Action level = 0.015 mg/L.

++ Action Level = 1.3 mg/L.

TT Treatment technique requirement.

Source: U.S.E.P.A.

9.2.2 *Irrigation.* – Irrigation has been considered for disposal of nanofiltration concentrates. Crops cannot generally tolerate high salinity irrigation water. If the salinity is low though, as with softening concentrates, and the water is mixed with freshwater, it may be usable for irrigation on some salt tolerant crops. Table 9.2 lists threshold levels of irrigation water salinity for several crops. 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 salinity in mg/L. These are levels above which yields begin to decrease linearly with increased salinity. Drainage must be considered as well. If drainage is inadequate, the threshold will be reached with a lower TDS irrigation water.

Because plants are capable of concentrating minerals from the soil, irrigation water should not contain excessive amounts of toxic metals. Table 9.3 lists guidelines for maximum trace element levels in irrigation water. Generally, if the concentrate meets these guidelines for TDS and trace element levels undiluted, it will be adequate for irrigation when mixed with freshwater.

Once the suitability of the water is established, the potential impacts of introducing the concentrate to the existing irrigation distribution and retention system must be evaluated. For instance, if the concentrate is discharged into an irrigation canal, will enough freshwater be there to dilute it sufficiently? Will aquatic organisms experience adverse impacts: will the canal experience adverse impacts? Impacts on ground water beyond the projected zone of discharge must be considered. Finally, potential impacts on the system that receives the excess blended water and runoff from the irrigation system must evaluated (Edwards and Bowdoin, 1990).

Table 9.2. – TDS threshold levels in applied irrigation water.

Crop	Maximum TDS	Crop	Maximum TDS
Lettuce	555	Alfalfa	853
Cotton Lint	3285	Grapes, table	640
Carrots	427	Cantaloupe	1422
Wheat	2560	Dates	1707
Oranges	725	Sugar Beets	2987
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, & Dorn, 1988, p.23)

Table 9.3. – Recommended maximum trace element levels in irrigation water.

Element	Suggested maximum irrigation water level (mg/L)
Aluminum (Al)	10.0
Arsenic	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 - 1.0
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 & Bowdoin, 1990 after Ayers & Westcot, 1976)

9.2.3 *Deep Well Injection.* – When surface discharge and irrigation are out of the question for concentrate disposal, deep well injection may be a solution. This option has certain geological requirements in addition to the regulatory requirements. The site must be over a confined aquifer that is unsuitable for drinking water, and transmissive enough to handle the amount of concentrate that will be generated (Muniz and Skehan, 1990).

The Clean Water Act does not regulate discharge into ground water because it only deals with navigable waters. The Safe Drinking Water Act, Resource Recovery and Conservation Act, and the Toxic Substances Control Act all have some effect on the legality of deep injection wells, but the states all have legislation to cover the specifics. Florida's Water Quality Assurance Act is one of the most comprehensive (Anderson et al., 1984). It classifies ground water into four categories: Class G-I is potable water with TDS less than 3,000 mg/L, class G-II is for unconfined aquifers containing from 3000 to 10,000 mg/L TDS, and Class G-IV is nonpotable water in confined aquifers with TDS $\geq$ 10,000 mg/L. Brackish RO and nanofiltration concentrates can be injected into Class G-III aquifers with $\geq$ 10,000 mg/L TDS or otherwise classified by the Environmental Regulation Commission as "having no potential as a source of drinking water," or Class G-IV aquifers. The concentrate must meet the primary drinking water standards set by the EPA under the SDWA, and be free from contamination by substances listed in the regulations, and deemed to be "carcinogenic, mutagenic, teratogenic or toxic or cause nuisance" (Dehan, 1990).

Because RO concentrate has been classified as industrial waste by the 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 9.1 illustrates the geologic layering and well construction requirements.

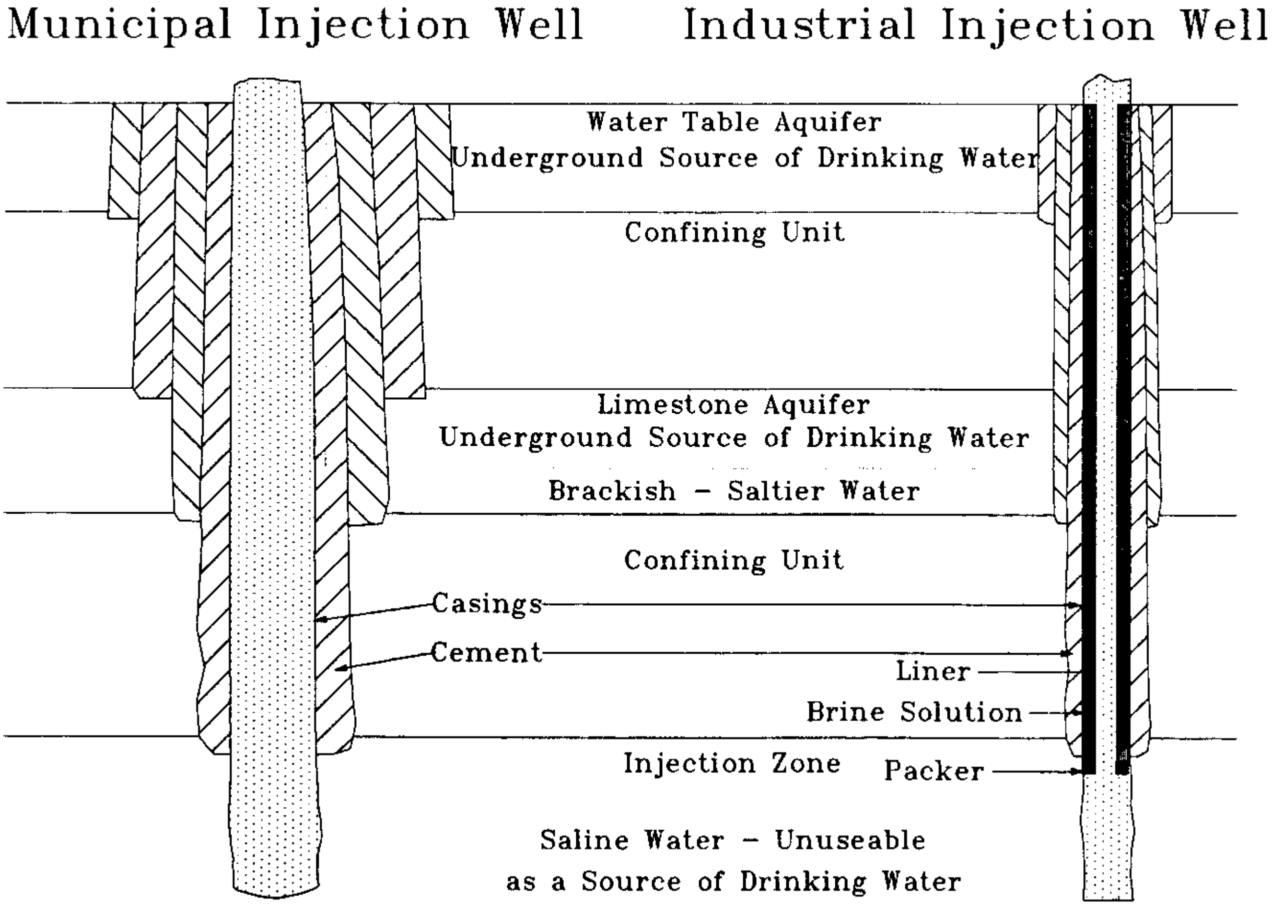


Figure 9.1. – Typical municipal and industrial injection well construction. (From *Florida Underground Injection Control Program Handbook*, FDER, 1983.)

9.2.4 *Solar Ponds*. – Solar ponds are heat traps that use high salinity water to trap heat from the sun that can then be used to generate electrical power or desalt water. A solar pond is from 3 to 7 m deep, constructed in three distinct layers as on figure 9.2. The surface has a thin layer of fresh or brackish 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 sodium chloride by weight, or other suitable salts, at the bottom. Normal convection currents cannot circulate water in this zone because of the density gradient caused by the increase in salinity with depth. The bottom of the pond is the thermal storage zone. This layer has a uniform salt 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 because it is less dense. The bottom of the pond is lined to prevent seepage into the ground water.

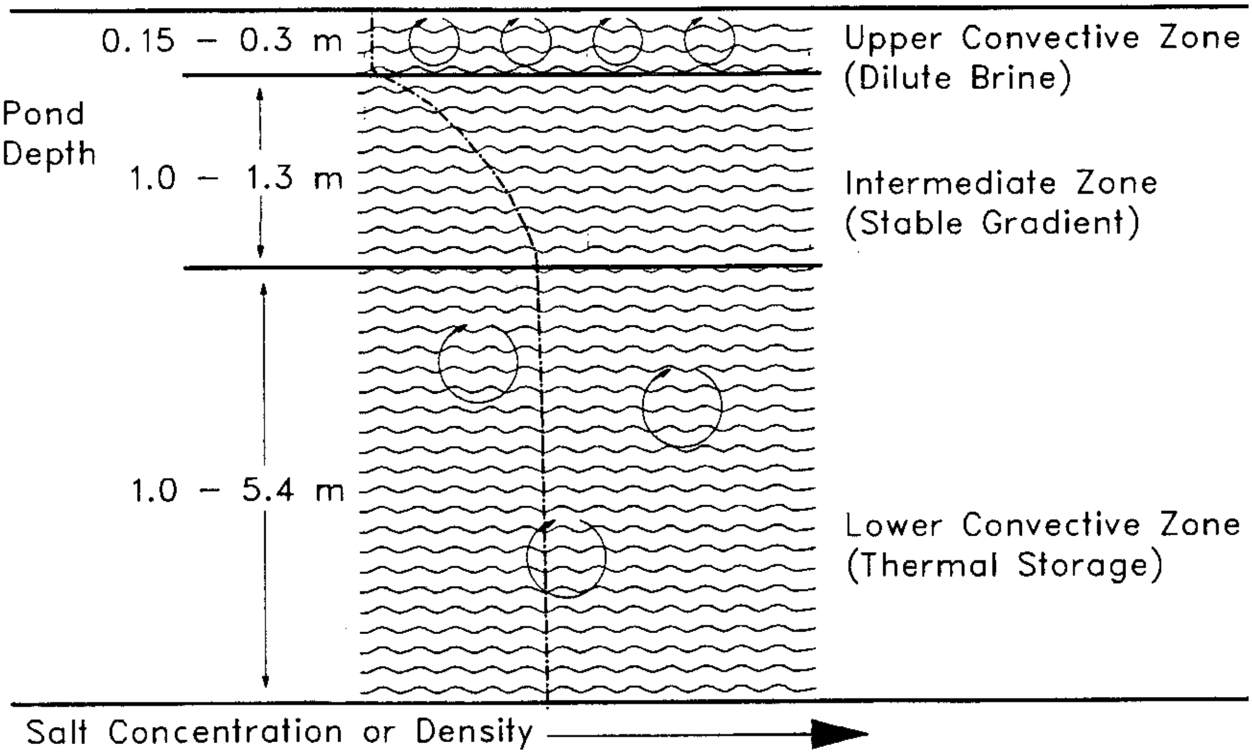


Figure 9.2. – Cross section of a solar pond.
(From Boegli et al., 1984)

The sun's energy penetrates the upper layers, warming the thermal storage zone just as i any lake, but the heat cannot circulate and dissipate into the atmosphere as it normall would. It becomes trapped in the dense lower layer of the pond. The temperature of th storage area becomes as high as 70 to 100 ° C (160 to 212 °F) within a few months after th salinity gradient is established. Heat in the storage zone may be used as process heat, o may be converted to electricity with thermal efficiency of up to 15 to 20 percent. Even durin the winter when the surface of the pond may be frozen, the storage zone will be hot enoug to generate electricity. (Boegli et al., 1983)

The first solar pond in the United States to generate electricity was the Bureau o Reclamation's El Paso Solar Pond. The El Paso Solar Pond project began as a University o Texas at El Paso research project sponsored by the Bureau of Reclamation and the Bruc Foods Corporation, and was later cost shared with the Texas Energy and Natural Resource Advisory Council and the El Paso Electric Company. After some experimentation wit process heat production, the solar pond was tied to the El Paso Electric Company power gri and began producing electricity on September 19, 1986. Figure 9.3 diagrams the process o electricity generation that is used at the El Paso Solar Pond. The hot brine can be circulate to the food processing plant to be used as preheat for boiler feedwater as needed, to a 100-kV organic Rankine cycle power conversion module, or to a multistage flash desalting system (Reid and Swift, 1987)

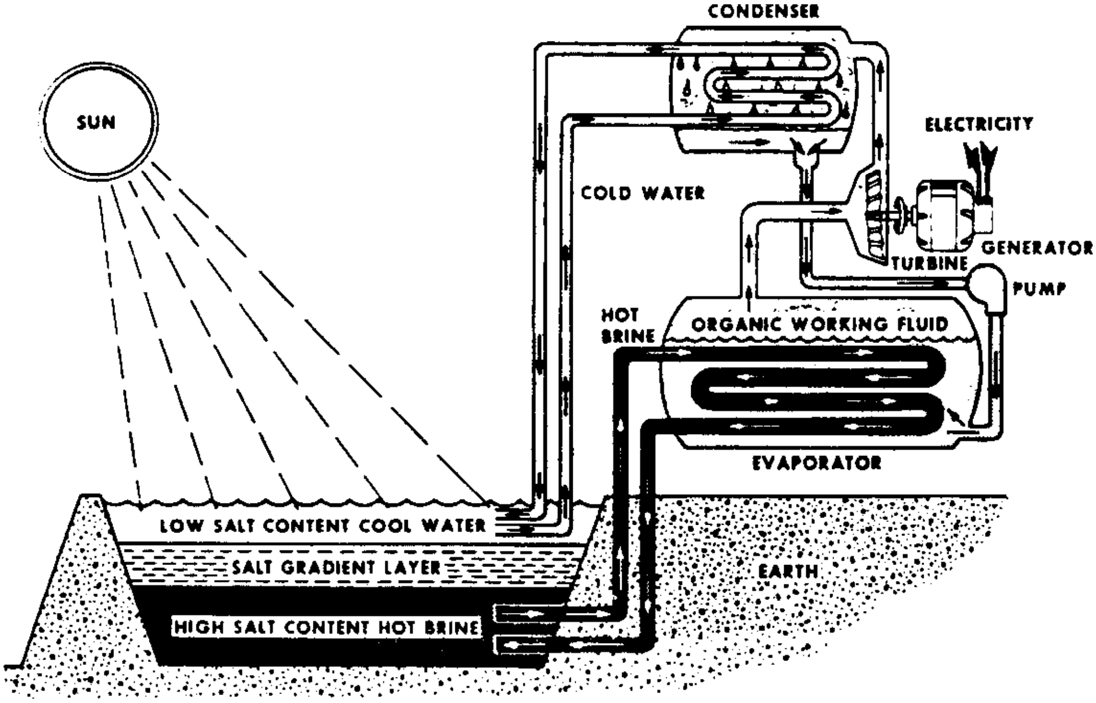


Figure 9.3. – Solar pond heat transfer for power generation. (From Boegli, et al., 1984)

To build the solar pond, the storage layer brine must be concentrated to 10 times the salinity of sea water, or 350,000 mg/L TDS. The normal RO recovery rate for sea water is 30 to 40 percent, which only produces a TDS of 50,000 to 67,000 mg/L. At another solar pond at the California Department of Water Resources, Los Banos, California, Demonstration Desalting Facility, brine was supplied by a three stage RO system designed to obtain 90 percent recovery of feed water with 8,930 mg/L TDS. Thermal evaporation was also used to increase recovery to 96 percent, resulting in a TDS of 223,250 mg/L. Further concentration of the brine could be carried out with heat energy from the solar pond, but to start, it was supplied by the local power utility (Smith, 1990). The Israelis have since developed an enhanced evaporation system that quickly concentrates the brine by spraying it down into evaporation ponds. This system is currently being used successfully at a saltworks (Hightower, personal communication, 1993).

The density gradient is important in maintaining a high temperature in the heat storage region. The gradient 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 brine at the right levels.

Solar ponds are a beneficial solution to the concentrate disposal problem, particularly at inland sites. The initial block of power produced can be used to run the pumps for the desalting system. Power production depends only on the desalting side for a source of brine, so as long as replacement brine 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 brine 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 the brine by brine concentration, deep well injection, or other means.

10. CURRENTLY AVAILABLE PRODUCTS

10.1 Explanation of Table Entries

10.1.1 *Manufacturers.* – The manufacturers included in this study were identified through their advertisements, the 1992 Thomas Registry, and communications with contacts in the membrane separations industry. The list of reverse osmosis membrane manufacturers is fairly complete for the United States. It is difficult to identify manufacturers in other countries, and more difficult to get information from them. Some manufacturers, such as Nitto-Denko and Toray Industries of Japan, market their municipal grade membranes through U.S. firms; other manufacturers did not respond to requests for information, but their addresses are included. Addresses are only given in the first category in which they appear. Toll free numbers should be used as indicated in the list.

10.1.2 *Model Number.* – Each manufacturer has a unique numbering system for their membranes. For most, however, the dimensions and composition of the membrane can be found in the model number. For example, Hydranautics model numbers have a four-digit prefix referring to length and width. This prefix is followed by a three letter code indicating the pressure tube used with it, and a four-digit code for composition and/or application. As an example, 8040-LSY-PVD1, is the model number for an 8- by 40-in. membrane module made of a polyvinyl alcohol derivative, using an 8-in. diameter, low pressure tube.

10.1.3 *Composition.* – The listed membrane composition is that given by the manufacturer in specification sheets. Some companies list their membrane's composition as just "Thin Film Composite." These membranes may be a combination of polysulfone and polyamide or cellulose acetate, but they are listed as TFC and grouped together in tables 10.10 and 10.11.

10.1.4 *Dimensions.* – If the diameter and length are whole integers, they are most likely nominal measurements. Some manufacturers give the actual dimensions, which are somewhat less than the nominal measurements. Dimensions supplied on the specification sheets are listed in the chart. The membrane manufacturer should be consulted before purchasing a membrane for use with another company's pressure tube.

Table 10.1. – Manufacturers of reverse osmosis membranes.

Desalination Systems, Inc. 1238A Simpson Way, Escondido, CA 92029	**Phone**: (619) 746-4995 **Toll Free**: 800-42-DESAL **TELEX**: 697890 **TELEFAX**: (619) 747-8253
The Dow Chemical Company 100 Larkin Center Midland, MI 48674	**Phone**: (517) 636-6880 **Toll Free**: 800-447-4369 **FAX**: (517) 638-9550
Du Pont Company "Permasep" Products Building 200, Glasgow Site, Wilmington, DE 19898	**Phone**: (302) 451-9681 **TELEX**: 6503433883 MCIUW **FAX**: (302) 451-9686
Fluid Systems - Allied Signal 10054 Old Grove Rd. San Diego, CA 92131	**Phone**: (619) 695-3840 **TELEX**: 188906 **Toll Free (for orders)**: 800-525-4369 **FAX**: (619) 695-2176
Hydranautics 8444 Miralani Drive San Diego, CA 92126	**Phone**: (619) 536-2500 **FAX**: (619) 536-2578 **TELEX**: 6839443
Ionpure Technologies Corp. 10 Technology Dr. Lowell, MA 01851	**Toll Free**: 800-783-PURE **FAX**: (508) 441-6025
Nitto-Denko America, Inc. 55 Nicholson Lane San Jose, CA 95134 (Also available through Hydranautics)	**Phone**: (408) 432-5400 **FAX**: (408) 432-5480 **TELEX**: 17-2540
Osmonics, Inc. 5951 Clearwater Dr. Minnetonka, MN 55343	**Phone**: (612) 933-2277 **FAX**: (612) 933-0141 **TELES**: 29-0847
Toray Industries, Inc. Membrane Products Dept. 2-2 Nihonbashi-Muromachi Chuo-Ku, Tokyo, 103, Japan	**Phone**: (03) 245-5607 **FAX**:(03) 245-5555 **TELEX**: J22623
Toyobo Co. Ltd. AC Operations Dept. 2-8 Dajima Hama 2-chrome Kita-ku Osaka, 530, Japan	**Phone**: (06) 348-3360 **FAX**:(06) 348-3332 **TELEX**: J63465 TOYOBO

Table 10.2. – Manufacturers of nanofiltration membranes.

Dow Chemical Co. – Filmtec

Fluid Systems

Hydranautics

Table 10.3. – Manufacturers of ultrafiltration membranes.

A/G Technology 34 Wexford Street Needham, MA 02194-2912	**Phone:** (617) 449-5774 **FAX:** (617) 449-5786
Cuno Separations Systems Division 50 Kerry Place Norwood, MA 02062	**Phone:** (617) 769-6112 **Toll Free:** 800-367-6805 **FAX:** (617) 769-3274
Desalination Systems	
Ionpure	
Koch Membrane Systems Inc. 850 Main Street Wilmington, MA 01887-3388	**Phone:** (508) 657-4250 **FAX:** (508) 657-5208 **TWX** 710 347 6537
Osmonics, Inc.	
Rhône-Poulenc CN 7500 Cranbury, NJ 80512-7500	**Phone:** (609) 860-3566 **FAX:** (609) 860-0129

Table 10.4. – Manufacturers of microfiltration membranes.
(Not complete)

A/G Technology Corp.
Ionpure
Osmonics

10.1.5 *Salt Rejection.* – Salt rejection is specified by manufacturers as chloride rejection. A average value is always given and most supply a minimum rejection. Where a minimum i not specified, the average is repeated as the minimum for flux computation. Hardnes rejection is given for nanofiltration membranes. RO membranes reject calcium an magnesium at 99.9 percent or better.

10.1.6 *Production.* – Production refers to permeate produced if operated for 24 hrs given i cubic meters per day. Manufacturers give the average production and the variability tha can be expected. Variability is usually 15 percent, but some are higher. The minimu production listed is the average production reduced by the specified variability.

10.1.7 *Membrane area.* – Membrane area is rarely on the specification sheet. Areas give came from either the specification sheets, literature supplied by the manufacturer, or fro sales representatives. Area is specified in square meters.

10.1.8 *Water Flux.* – Minimum and average flux are calculated from the membrane area an the minimum and average production. Flux is given as cubic meters per square meter pe day.

10.1.9 *Maximum Feed Flow.* – The maximum feed flow, in liters per minute, is the maximum recommended flow rate to each element. Running at higher than the recommended feed flow may void the warranty. The manufacturer should be able to suggest alternatives that would achieve the needed productivity level and still be within warranty obligations.

10.1.10 *Minimum concentrate flow.* – Minimum concentrate flow depends more on the silt density index and Langelier index than on membrane composition. Some manufacturers specify a minimum concentrate to permeate ratio for the last membrane of the last stage. For instance, if the ratio is 5:1, concentrate to permeate, the system may be operated at 6:1, but not 4:1. du Pont uses a maximum brine flow rate which is given in liters per minute. The standard Filmtec membrane warranty specifies that the "recovery ratio shall be consistent with concentration of sparingly soluble salts." This means that if the system is run at a recovery rate that results in scaling, the warranty is voided. Fluid Systems recommends that the salt concentration factor, β, at the membrane surface should not exceed 1.13. β is calculated from the water flux (F_w), the water transport coefficient for the membrane (A), applied pressure (P), and the osmotic pressure of the main stream (π):

$$\beta = -\frac{Fw - AP}{A\pi}$$

The dependence of the concentration factor on operating conditions and water composition is why manufacturers want to have a water analysis before they give membrane system quotes.

10.1.11 *Test conditions.* – The conditions under which the membranes were tested to obtain rejection and productivity are given here. Most manufacturers give the NaCl concentration, applied pressure, temperature in degrees centigrade, pH, and percent recovery. Water used in most cases is RO permeate. If test conditions were not reported in this way, a note explains what water was used.

10.1.12 *Pressure drop across element.* – The maximum allowable pressure drop across an element is listed in this column. Pressure drops greater than those given indicate that some type of fouling has occurred in the membrane. Smaller pressure drops indicate membrane damage. Most manufacturers prefer to use a 10-percent change in permeate flow as an indicator of fouling or damage.

10.1.13 *Recommended operating parameters.* – Recommended operating parameters include applied pressure, temperature, pH, maximum SDI (silt density index), NTU (nephelometric turbidity units, and maximum continuous concentration of oxidants. Applied pressure and temperature extremes are self explanatory. The pH limits given here are for continuous operation.

The SDI is a measurement of the level of colloids in the feed water. The maximum for spiral wound membranes is 5. Hollow fine fiber membranes require an SDI of 3 to 4 depending on the brand. Well water usually has an SDI of 1.0 and does not require pretreatment for colloids. Surface water SDI ratings range from 10 to 175 (du Pont, 1982).

Maximum continuous concentration of oxidants is the concentration while in operation. Higher levels may be allowed for short cleaning cycles. As stated earlier, only cellulosic membranes are oxidant tolerant. Others may be "resistant," which means that the membrane can survive small amounts of chlorine or ozone in the feed stream, but their useful life will be shortened.

Turbidity, in NTU, is a measurement of light scattering by colloids and other various suspended particles. Source water turbidity can range from 1 for very clear water to several thousands for pond water. The maximum NTU allowed with RO membranes is 1.0. Over that level, fouling of the membrane will probably be a problem.

10.1.14 *Cleaning Parameters.* – These columns give the minimum and maximum cleaning pH and oxidant concentration. Cleaning parameters are generally less restrictive than the operating parameters because they are for short periods of time and the cleaning solution is flushed out with RO permeate after the cycle is completed.

10.1.15 *Vital Statistics for Vessels.* – Elements per vessel, or pressure tube, and maximum recovery per vessel refer to statistics for the largest size vessel or pressure tube available from the manufacturer under standard operating conditions. Not all companies supply this information because they will build whatever size pressure tube is needed.

10.1.16 *Membrane Flux Coefficients.* – The membrane flux coefficients "*A*," for water transport in 10^{-9} m/s*kPa, and "*B*", for salt transport in 10^{-9} m/s, are calculated from the following:

- Temperature in degrees Kelvin

$$T = \text{Operating Temp. °C} + 273.15$$

- C_f, solute concentration of feed (moles/m^3):

$$C_f = \frac{\text{NaCl mg/L} * 1000 \text{ L/m}^3}{58{,}440 \text{ mg/mole}}$$

- C_p, solute concentration of product (moles/m^3):

$$C_p = C_f * \left(1 - \frac{\text{Ave. Cl}^- \text{ rejection}}{100}\right)$$

- C_r, solute concentration of reject (moles/m^3):

$$C_r = C_f - \frac{\%\ \text{recovery} * C_p}{100 - \%\ \text{recovery}}$$

- C_M, the mean concentration of the reject (moles/m^3):

$$C_M = \frac{C_f + C_r}{2}$$

- $\Delta\pi$, difference in osmotic pressure in kPa between mean concentration of the reject and product:

$$\Delta\pi = \phi i R T * (C_M - C_P)$$

where:
ϕ is the osmotic coefficient to 0.99 for brackish water and 0.94 for seawater,
i is the number of ions per molecule, 2 for NaCl,
R is the Gas Constant, 8.314 joules/mole °K,
T is the temperature in °K.

- ΔP, is the change in pressure of the reject in kPa.

$$\Delta P = \text{Applied Pressure} - \text{Pressure Drop}$$

with these values, A and B can be calculated from the following equations:

$$A = \frac{F_w}{(\Delta P - \Delta\pi)}$$

$$B = \frac{F_w * C_P}{(C_M - C_P)}$$

where:
F_w is water flux in m^3/m^2*s
A is in 10^{-9} m/s*kPa
B is in 10^{-6} m/s

10.2 Membrane Listing

Membranes specifications are presented in the following tables according to their application, size, and composition. Each table consists of three sections. Because of the large number of membranes available, brackish water RO membranes are in six sections – two sets of three.

Table 10.5. - Seawater reverse osmosis membranes by producer.
Table 10.6. - Brackish water reverse osmosis membranes by producer.
Table 10.7. - Nanofiltration membranes by producer.
Table 10.8. - Ultrafiltration membranes by producer.
Table 10.9. - Microfiltration membranes by producer.
Table 10.10. - 8-inch seawater reverse osmosis membranes by composition.
Table 10.11. - 8-inch brackish water reverse osmosis membranes by composition.
Table 10.12. - 8-inch nonofiltration membranes by composition.

The following notes are for reverse osmosis and nanofiltration membrane listings in tables 10.5 through 10.7 and 10.10 through 10.12.

Notes:

1 Optimum operating pH: 5-8. 1000 mg/L-h oxidant tolerance.
2 Chlorine sanitation, 30 mg/L for 30 min.
3 1000 mg/L-h oxidant tolerance.
4 San Diego tap water for test conditions.
5 Not available in U.S. or its possessions or territories.
6 Use softened or deionized feed water: <1 mg/L Ca^{+2}, or Mg^{+2}.
7 Feed water must be dechlorinated.
8 Maximum fouling index of 4.
9 2000 mg/L-h oxidant tolerance.

10.3 Comparisons

To condense the massive amount of information, the parameters for 8-in. diameter brackish and seawater membranes are summarized in table 10.13. Productivity measures are difficult to evaluate. Design productivity in cubic meters per day are given in all RO and NF specifications, but the conditions vary between companies so widely, that a comparison based on cubic meters per day would be meaningless. Figures 10.1, 10.2, and 10.3 are graphs comparing membrane water flux coefficients (the "*A*" Value) with average water flux (m^3/m^2*day) and productivity (L/m^3 module volume per day) for brackish and seawater RO, nanofiltration and ultrafiltration membranes. The data are separated by module type (hollow fine fiber and spiral) and membrane composition.

The "*A*" value is directly proportional to average flux and inversely proportional to net operating pressure. Although "*A*" is supposed to be a constant characteristic of the membrane, its initial value depends heavily on the initial testing conditions. Unfortunately, testing conditions for membranes are in no way consistent among manufacturers, so figure 10.1 is really no help in deciding which membranes have higher productivity.

Figure 10.2 is more valuable because it compares productivity per cubic centimeter of membrane module to the "*A*" value. This plot shows that the "*A*" value has little effect on overall module productivity. The hollow fine fiber membrane modules have so much surface area per unit of volume that their average flux per square meter approaches their productivity per cubic centimeter! Notice that these membranes have very low "*A*" values. The other membrane formations have higher "*A*" values, but none of them manage to produce more than 1.3 L/cm^3*day. The one brackish water TFC that did produce 1.4 L/cm^3*day was tested on tap water (500 mg/L). The others were tested with 1000 - 2000 mg/L NaCl.

Table 10.5.1. – Seawater reverse osmosis membranes by producer.

IFACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	DIVALENT REJECTION	CI REJECT		PROD. (m3/d)		FLUX (m3/m2*d)		MIN CONC:PERM	MAX FEED FLOW (L/min)
							MIN	AVE	AVE	MIN	AVE	MIN		
nt Permasep	B-10 6835T	Aramid HFF	22	81	450.6	99.6	98.7	99.2	26.5	22.5	0.059	0.050	15.9	300
	B-10 6880T	Aramid HFF	22	205	901.2	99.6	98.7	99.2	53.0	44.3	0.059	0.049	31.8	600
c	SW30HR-4040	Polyamide TFC	10	102	6.5		99.2	99.4	3.8	3.2	0.582	0.495	11:1	38
	SW30HR-8040	Polyamide TFC	20	102	30.7		99.2	99.4	15.1	12.9	0.494	0.420	11:1	151
	SW30-8040	Polyamide TFC	20	102	30.7		98.6	99.1	22.7	19.3	0.741	0.630	9:1	181
Systems	1021 SS	Polyether Urea TFC	10	102	7.4		99.2	99.4	3.8	3.2	0.509	0.433	13:1	43
	1501	Polyether Urea TFC	15	102	13.9		99.0	99.3	7.9	6.8	0.570	0.485	13:1	91
	2021 HF	Polyether Urea TFC	20	102	30.2		98.6	99.0	22.7	19.3	0.752	0.639	13:1	259
	2021 SS	Polyether Urea TFC	20	102	30.2		99.2	99.4	15.1	12.9	0.502	0.426	13:1	173
	2821 HP	Polyamide TFC	20	102	30.2		99.2	99.4	18.9	16.1	0.627	0.533	13:1	216
nautics	4040-HSA-SWC1	Polyamide TFC	10	102	6.5		99.2	99.5	4.5	3.9	0.699	0.594	9:1	36
	8040-HSY-SWC1	Polyamide TFC	20	102	29.3		99.2	99.5	18.9	16.1	0.647	0.550	9:1	151
o	HR5155	Cellulosic HFF	15	44			99.2	99.4	1.2	0.9			2.3:1	3
	HR5255	Cellulosic HFF	15	82			99.2	99.4	3.0	2.4			2.3:1	7
	HR5355	Cellulosic HFF	15	123			99.2	99.4	5.0	4.0			2.3:1	12
	HR8355	Cellulosic HFF	30	133			99.2	99.4	12.0	10.0			2.3:1	28
	HM8255	Cellulosic HFF	30	264			99.2	99.4	27.5	25.0			2.3:1	64
	HM9255	Cellulosic HFF	36	267			99.2	99.4	35.0	32.0			2.3:1	81

Table 10.5.2. – Seawater reverse osmosis membranes by producer.

		TESTING CONDITIONS					MAX	MAX PRESSURE DROP/	MAX	MAX	MAX	OPERATING pH		CLEANING pH		MAX CONT. OX.	# ELEMENTS/	MAX RECOVERY/	SHIPPED WEIGHT
FACTURER	MODEL#	mg/L NaCl	kPa	% RECOVERY	TEMP. °C	pH	kPa	ELEMENT	SDI	NTU	°C	MIN	MAX	MIN	MAX	CONC.	VESSEL	VESSEL	(kg)
nt Permasep	B-10 6835T	35000	6895	35	25		8274	60	4		40	4.0	9.0	2.3	11.9	0.0	1	50	59
	B-10 6880T	35000	6895	35	25		8274	60	4		40	4.0	9.0	2.3	11.9	0.0	2	50	113
c	SW30HR-4040	32000	5516	8	25	8.0	6895	138	5	1	45	2.0	11.0	1.0	12.0	0.1			
	SW30HR-8040	32000	5516	8	25	8.0	6895	138	5	1	45	2.0	11.0	1.0	12.0	0.1	6		
	SW30-8040	35000	5516	10	25	8.0	6895	138	5	1	45	2.0	11.0	1.0	12.0	0.1	6		
ystems	1021 SS	32800	5516	7	25	5.7	6895	69		1	45	5.0	10.0	4.5	11.0	0.0	6	42	5
	1501	32800	5516	7	25	5.7	6895	69		1	45	5.0	10.0	4.5	11.0	0.0	6	45	9
	2021 HF	32800	5516	7	25	5.7	6895	69		1	45	5.0	10.0	4.5	11.0	0.0	6	42	18
	2021 SS	32800	5516	7	25	5.7	6895	69		1	45	5.0	10.0	4.5	11.0	0.0	6	42	18
	2821 HP	32800	5516	7	25	7.5	6895	69		1	45	4.0	11.0	2.5	11.0	0.0	6	42	16
nautics	4040-HSA-SWC1	32000	5516	10	25	7.0	6895	69	4	1	45	3.0	10.0	4.0	10.0	0.1			8
	8040-HSY-SWC1	32000	5516	10	25	7.0	6895	69	4	1	45	3.0	10.0	4.0	10.0	0.1	7		18
o	HR5155	35000	5394	30	25		5884	7	4		40	3.0	8.0	3.0	8.0	1.0	1	30	13
	HR5255	35000	5394	30	25		5884	7	4		40	3.0	8.0	3.0	8.0	1.0	1	30	23
	HR5355	35000	5394	30	25		5884	14	4		40	3.0	8.0	3.0	8.0	1.0	1	30	32
	HR8355	35000	5394	30	25		5884	7	4		40	3.0	8.0	3.0	8.0	1.0	1	30	125
	HM8255	35000	5394	30	25		6374	19	4		40	3.0	8.0	3.0	8.0	1.0	2	30	205
	HM9255	35000	5394	30	25		6864	19	4		40	3.0	8.0	3.0	8.0	1.0	2	30	310

Table 10.5.3. – Seawater reverse osmosis membranes by producer.

		VITAL STATISTICS									L/DAY	MODEL # FOR OTHER SIZES			
ACTURER	MODEL#	TEMP°	Cf	Cp	Cr	CM	dPI	dP	A	B	PER CC	2.5" DIA.	4" DIA.	OTHER SIZES	NOTES:
Permasep	B-10 6835T	298	598.9	4.8	918.8	758.9	3701.1	6835.0	0.217	0.0004	0.86			5x23", 5x50"	
	B-10 6880T	298	598.9	4.8	918.8	758.9	3701.1	6835.0	0.217	0.0004	0.68		B-10 6410, 6440		
	SW30HR-4040	298	547.6	3.3	594.9	571.2	2787.6	5378.1	2.601	0.0034	0.46				
	SW30HR-8040	298	547.6	3.3	594.9	571.2	2646.8	5378.1	2.093	0.0029	0.46			153x444, 210x816, 210x1216mm	
	SW30-8040	298	598.9	5.4	664.9	631.9	2919.7	5378.1	3.486	0.0064	0.69	SW30-2540	SW30-4040		
stems	1021 SS	298	561.3	3.4	603.3	582.3	2697.8	5447.1	2.144	0.0030	0.46				
	1501	298	561.3	3.9	603.2	582.2	2695.1	5447.1	2.399	0.0039	0.43				
	2021 HF	298	561.3	5.6	603.1	582.2	2687.0	5447.1	3.154	0.0073	0.69				
	2021 SS	298	561.3	3.4	603.3	582.3	2697.8	5447.1	2.111	0.0029	0.46		1021 SS		
	2821 HP	298	561.3	3.4	603.3	582.3	2697.8	5447.1	2.639	0.0036	0.57	SW30HR-2540	SW30HR-4040		
autics	4040-HSA-SWC1	298	547.6	2.7	608.1	577.8	2680.2	5447.1	2.922	0.0033	0.57				
	8040-HSY-SWC1	298	547.6	2.7	608.1	577.8	2680.2	5447.1	2.706	0.0031	0.58				
	HR5155	298	598.9	3.6	854.0	726.5								153 X 444 mm	
	HR5255	298	598.9	3.6	854.0	726.5								153 X 825 mm	
	HR5355	298	598.9	3.6	854.0	726.5								153 X 1230 mm	
	HR8355	298	598.9	3.6	854.0	726.5								305 X 1330 mm	
	HM8255	298	598.9	3.6	854.0	726.5								298 X 2640 mm	
	HM9255	298	598.9	3.6	854.0	726.5								360 X 2665 mm	

Table 10.6.1. – Brackish water reverse osmosis membranes by producer.

							CI REJECT		PROD. (m3/d)		FLUX (m3/m2*d)			MAX FEED
FACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	DIVALENT REJECTION	MIN	AVE	AVE	MIN	AVE	MIN	MIN CONC:PERM	FLOW (L/min)
ation Systems	CD4040F	Cellulosic	10	102	8.4		96.5	97.5	6.1	5.1	0.7	0.6	5:1	48
	CD8040F	Cellulosic	20	102	31.6		96.5	97.5	22.7	19.3	0.7	0.6	5:1	181
	CE4040F	Cellulosic	10	102	8.4		95.0	96.0	7.9	6.8	1.0	0.8	5:1	63
	CE8040F	Cellulosic	20	102	31.6		95.0	96.0	30.3	25.7	1.0	0.8	5:1	242
	ND4040F	Thin Film Composite	10	102	8.4	98.5	97.0	98.0	6.8	5.8	0.8	0.7	5:1	54
	ND8040F	Thin Film Composite	20	102	31.6	98.5	97.0	98.0	25.0	21.2	0.8	0.7	5:1	200
	SE4040F	Thin Film Composite	10	102	8.4		97.5	98.5	7.6	6.4	0.9	0.8	5:1	60
	SE8040F	Thin Film Composite	20	102	31.6		97.5	98.5	29.1	24.8	0.9	0.8	5:1	233
	SG4040F	Thin Film Composite	10	102	8.4	99.5	97.0	98.0	7.6	6.4	0.9	0.8	5:1	60
	SG8040F	Thin Film Composite	20	102	31.6	99.5	97.0	98.0	29.1	24.8	0.9	0.8	5:1	233
	SH4040F	Thin Film Composite	10	102	8.4	99.5	95.0	96.0	7.6	6.4	0.9	0.8	5:1	60
	SH8040F	Thin Film Composite	20	102	31.6	99.5	95.0	99.0	29.1	24.8	0.9	0.8	5:1	233
t Permasep	B-9 0840	Aramid HFF	20	122	427.4	96.0	90.0	92.0	60.6	54.5	0.1	0.1	26.5	88
	B-9 0880	Aramid HFF	20	178	845.4	97.5	90.0	95.0	140.1	126.0	0.2	0.2	53.0	176
:	BW30-4040	Polyamide TFC	10	102	6.5		96.0	98.0	6.8	5.8	1.0	0.9	6:1	36
	BW30-8040	Polyamide TFC	20	102	30.7		96.0	98.0	28.4	24.1	0.9	0.8	6:1	151
ystems	4021 LP	Polyether Urea	10	102	7.4		96.0	97.5	6.8	5.8	0.9	0.8	9:1	54
	4221 HR	Cellulosic	10	102	7.4		97.0	98.0	6.1	5.1	0.8	0.7	9:1	48
	4221 SD	Cellulosic	10	102	7.4		94.0	95.5	7.6	6.4	1.0	0.9	9:1	60
	4231 HR	Cellulosic	10	152	11.1		97.0	98.0	10.0	8.5	0.9	0.8	5:1	50
	4231 SD	Cellulosic	10	152	11.1		94.0	95.5	12.3	10.5	1.1	0.9	5:1	61
	4821 LP	Polyamide TFC	10	102	7.4		96.5	98.5	6.8	5.8	0.9	0.8	9:1	54
	4826 LP	Polyamide TFC	10	102	7.4		96.0	98.0	7.6	6.4	1.0	0.9	9:1	60
	8021 LP	Polyether Urea	20	102	30.2		96.0	97.5	26.5	22.5	0.9	0.7	9:1	212
	8021 MP	Polyether Urea	20	102	30.2		98.0	99.0	30.7	26.1	1.0	0.9	9:1	245
	8221 HR	Cellulosic	20	102	30.2		97.0	98.0	23.8	20.3	0.8	0.7	9:1	190
	8221 SD	Cellulosic	20	102	30.2		94.0	95.5	30.3	25.7	1.0	0.9	9:1	242
	8231 HR	Cellulosic	20	152	48.8		97.0	98.0	39.7	33.8	0.8	0.7	5:1	198
	8231 SD	Cellulosic	20	152	48.8		94.0	95.5	49.2	41.8	1.0	0.9	5:1	246
	8821 LP	Polyamide TFC	20	102	30.2		96.5	98.5	28.4	24.1	0.9	0.8	9:1	227

Table 10.6.2. – Brackish water reverse osmosis membranes by producer.

		TESTING CONDITION						MAX PRESSURE				OPERATING pH		CLEANING pH		MAX	#	MAX	SHIPPED
ACTURER	MODEL#	mg/L NaCl	kPa	% RECOVERY	TEMP. °C	pH	MAX kPa	DROP/ ELEMENT	MAX SDI	MAX NTU	MAX °C	MIN	MAX	MIN	MAX	CONT. OX. CONC.	ELEMENTS VESSEL	REC./ VESSEL	WEIGHT (kg)
ation Systems	CD4040F	1000	2758	10	25		3103	69	5	1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	CD8040F	1000	2758	10	25		4137	69	5	1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	CE4040F	1000	2758	10	25		3103	69	5	1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	CE8040F	1000	2758	10	25		4137	69	5	1	35	5.0	6.5	3.0	8.0	1.0	6	53	5
	ND4040F	1000	2758	10	25		4137	69	5	1	50	4.0	11.0	2.0	11.5	Low	6	53	16
	ND8040F	1000	2758	10	25		4137	69	5	1	50	4.0	11.0	2.0	11.5	Low	6	53	16
	SE4040F	1000	2758	10	25		4137	69	5	1	50	4.0	11.0	2.0	11.5	Low	6	53	5
	SE8040F	1000	2758	10	25		4137	69	5	1	50	4.0	11.0	2.0	11.5	Low	6	53	16
	SG4040F	1000	1379	10	25	7.8	4137	69	5	1	50	4.0	11.0	2.0	11.5	0.1	6		16
	SG8040F	1000	1379	10	25	7.8	4137	69	5	1	50	4.0	11.0	2.0	11.5	0.1	6	53	5
	SH4040F	1000	1034	10	25	7.8	4137	69	5	1	50	4.0	11.0	2.0	11.5	0.1	6	53	16
	SH8040F	1000	1034	10	25	7.8	2758	60	5	1	50	4.0	11.0	2.0	11.5	0.1	6	53	16
t Permasep	B-9 0840	1500	2758	75	25		2758	60	3		40	4.0	11.0	2.3	11.9	0.0	1	50	59
	B-9 0880	1500	2758	75	25		4137	138	4		40	4.0	11.0	2.3	11.9	0.0	2	50	113
:	BW30-4040	2000	1551	15	25	8.0	4137	138	5	1	45	2.0	11.0	1.0	12.0	0.1	6	50	
	BW30-8040	2000	1544	15	25	8.0	4137	69	5	1	45	2.0	11.0	1.0	12.0	0.1	6		
ystems	4021 LP	2000	2896	10	25	5.7	4137	103		1	45	5.0	10.0	4.5	11.0	0.0	6	53	5
	4221 HR	2000	2896	10	25	5.7	4137	103		1	40					1.0			5
	4221 SD	2000	2896	10	25	5.7	4137	103		1	40	3.0	7.0	3.0	8.0	1.0	6	53	5
	4231 HR	2000	2896	16	25	5.7	4137	103		1	40	4.0	6.0	3.0	7.0	1.0	4	55	7
	4231 SD	*2000*	*2896*	*16*	*25*	*5.7*	*2413*	*69*		*1*	*40*	*3.0*	*7.0*	*3.0*	*8.0*	*1.0*	*4*	55	*7*
	4821 LP	2000	1551	10	25	7.5	2069	34		1	45	4.0	11.0	2.5	11.0	0.0	6	53	5
	4826 LP	2000	1551	10	25	7.5	2413	69		1	45	4.0	11.0	2.5	11.0	0.0	3	40	5
	8021 LP	2000	1551	10	25	5.7	4137	69		1	45	5.0	10.0	4.5	11.0	0.0	6	53	18
	8021 MP	2000	2896	10	25	5.7	4137	103		1	45	5.0	10.0	4.5	11.0	0.0	6	53	18
	8221 HR	2000	2896	10	25	5.7	4137	103		1	40	4.0	6.0	3.0	7.0	1.0	6	53	18
	8221 SD	2000	2896	10	25	5.7	4137	103		1	40	3.0	7.0	3.0	8.0	1.0	6	53	18
	8231 HR	2000	2896	16	25	5.7	4137	103		1	40	4.0	6.0	3.0	7.0	1.0	4	55	26
	8231 SD	2000	2896	16	25	5.7	4137	69		1	40	3.0	7.0	2.5	8.0	1.0	4	55	26
	8821 LP	2000	1551	10	25	7.5	1551	69		1	45	4.0	11.0	2.5	11.0	0.0	6	53	18

Table 10.6.3. – Brackish water reverse osmosis membranes by producer.

		VITAL STATISTICS									L/DAY PER	MODEL # FOR OTHER SIZES			
FACTURER	MODEL#	TEMP°	Cf	Cp	Cr	CM	dPI	dP	A	B	CC	2.5" DIA.	4" DIA.	OTHER SIZES	NOTES:
ation Systems	CD4040F	298	17.1	0.4	19.0	18.0	86.4	2689.1	3.221	0.018	0.78		CE4040F	25.0" x 3.88", 26.25"x 3.88"	2
	CD8040F	298	17.1	0.4	19.0	18.0	86.4	2689.1	3.198	0.017	0.71				
	CE4040F	298	17.1	0.7	18.9	18.0	85.1	2689.1	4.226	0.038	1.03				2
	CE8040F	293	17.1	0.7	18.9	18.0	83.7	2689.1	4.259	0.036	0.95	no	CD4040F	26.25"x 3.88", 25.0"x 3.88"	
	ND4040F	298	17.1	0.3	19.0	18.0	86.9	2689.1	3.625	0.016	0.88				
	ND8040F	298	17.1	0.3	19.0	18.0	86.9	2689.1	3.518	0.015	0.78				3
	SH4040F	298	17.1	0.3	19.0	18.0	87.3	2689.1	4.028	0.013	0.98		SH4040F	SH4026F, SH4025T	
	SH8040F	298	17.1	0.3	19.0	18.0	87.3	2689.1	4.105	0.013	0.91		ND4040F	26.25" x 3.88"	3
	SE4040F	298	17.1	0.3	19.0	18.0	86.9	1310.1	8.568	0.018	0.98				1
	SE8040F	298	17.1	0.3	19.0	18.0	86.9	1310.1	8.732	0.018	0.88		SE4040F	25.00" x 3.88", 26.25" x 3.88	1
	SG4040F	298	17.1	0.7	18.9	18.0	85.1	965.3	11.907	0.036	0.98		SG4040F	26.25" x 3.88", 25.0" x 3.88"	
	SG8040F	298	17.1	0.2	19.0	18.1	87.8	974.3	12.048	0.009	0.88				
Permasep	B-9 0840	298	25.7	2.1	96.5	61.1	289.8	2698.0	0.681	0.005	1.53		B-90410, 0420, 0440	10x55" 0040	
	B-9 0880	298	25.7	1.3	98.8	62.2	299.2	2620.1	0.826	0.003	2.43				
	BW30-4040	298	34.2	0.7	40.1	37.2	179.1	1413.5	9.825	0.020	0.83	BW30-2540	BW30-4040		
	BW30-8040	298	34.2	0.7	40.1	37.2	179.1	1475.5	8.268	0.017	0.86				
ystems	4021 LP	298	34.2	0.9	37.9	36.1	172.9	2792.5	4.051	0.022	0.83				
	4221 HR	298	34.2	0.7	37.9	36.1	173.8	2792.5	3.602	0.016	0.74				
	4221 SD	298	34.2	1.5	37.9	36.0	169.3	2792.5	4.495	0.045	0.92				
	4231 HR	298	34.2	0.7	40.6	37.4	180.3	2792.5	3.987	0.017	0.81		4231 HR		
	4231 SD	298	34.2	1.5	40.4	37.3	175.7	2827.0	4.818	0.047	1.00				
	4821 LP	298	34.2	0.5	38.0	36.1	174.6	1516.9	7.906	0.013	0.83				
	4826 LP	298	34.2	0.7	37.9	36.1	173.8	1482.4	9.009	0.020	0.92				
	8021 LP	298	34.2	0.9	37.9	36.1	172.9	1482.4	7.757	0.021	0.80		4021 LP		
	8021 MP	298	34.2	0.3	38.0	36.1	175.5	2792.5	4.492	0.010	0.93		4021 MP		
	8221 HR	298	34.2	0.7	37.9	36.1	173.8	2792.5	3.491	0.015	0.72		4221 HR		
	8221 SD	298	34.2	1.5	37.9	36.0	169.3	2792.5	4.426	0.045	0.92		4221 SD		
	8231 HR	298	34.2	0.7	40.6	37.4	180.3	2792.5	3.611	0.015	0.80		4231 HR		
	8231 SD	298	34.2	1.5	40.4	37.3	175.7	2827.0	4.405	0.043	1.00		4231 SD		
	8821 LP	298	34.2	0.5	38.0	36.1	174.6	1482.4	8.322	0.014	0.86		4821 LP		

Table 10.6.4. – Brackish water reverse osmosis membranes by producer.

ACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	DIVALENT REJECTION	Cl REJECT		PROD. (m3/d)		FLUX (m3/m2*d)		MIN CONC:PERM	MAX FEED FLOW (L/min)
							MIN	AVE	AVE	MIN	AVE	MIN		
utics	4040-LSA-CPA2	Polyamide TFC	10	102	7.0		99.0	99.0	7.6	6.4	1.087	0.924	5:1	40
	4040-LSA-NCM1	Polyamide TFC	10	102	7.0		98.0	99.0	7.2	6.1	1.032	0.877	5:1	38
	4040-LST-CPA2	Polyamide TFC	10	102	7.0		99.0	99.0	7.6	6.4	1.087	0.924	5:1	40
	4040-LST-NCM1	Polyamide TFC	10	102	7.0		98.0	98.0	7.2	6.1	1.032	0.877	5:1	38
	4040-MSA-CAB1	Cellulosic	10	102	7.0		95.0	92.5	7.2	6.1	1.032	0.877	5:1	57
	4040-MSA-CAB2	Cellulosic	10	102	7.0		98.0	97.0	5.7	4.8	0.815	0.693	5:1	45
	4040-MSA-CAB3	Cellulosic	10	102	7.0		99.0	98.5	3.8	3.2	0.543	0.462	5:1	30
	8040-MSY-CAB1	Cellulosic	20	102	31.6		92.5	95.0	32.2	27.3	1.019	0.866	5:1	257
	8040-MSY-CAB2	Cellulosic	20	102	31.6		97.0	98.0	25.7	21.9	0.815	0.693	5:1	206
	8040-MSY-CAB3	Cellulosic	20	102	31.6		98.5	99.0	17.0	14.5	0.539	0.458	5:1	136
	8040-LSY-CPA2	Polyamide TFC	20	102	30.2		99.0	99.0	34.1	29.0	1.128	0.959	5:1	181
	8040-LSY-NCM1	Polyamide TFC	20	102	33.9		98.0	99.0	34.1	29.0	1.005	0.854	5:1	181
	8540-LSY-NCM1	Polyamide TFC	22	102	39.0		98.0	99.0	39.7	33.8	1.019	0.866	5:1	212
	CDRO M60 S1	Cellulosic	10	64	5.6	96.0	93.0	93.0	1.9	1.6	0.338	0.287	6:1	56
	CDRO 120 S0	Cellulosic	15	64	11.1	96.0	93.0	93.0	6.2	5.3	0.556	0.472	6:1	136
	CDRC M60 S1	Polyamide TFC	9	64	5.6		95.0	98.0	6.8	5.8	1.214	1.032		56
	SM1B A12 V7	Polyamide TFC	10	102	8.4	98.0	96.0	96.0	6.8	5.8	0.810	0.688		56
	CDRS M60 S1	Polysulfone TFC	10	64	5.6		95.0	95.0	5.3	4.5	0.943	0.801		56
ics	815HR (PA)	Polyamide TFC	21	102	32.0		96.0	98.0	28.4	24.1	0.137	0.116	5:1	265
	811HR (PA)	Polyamide TFC	21	102	35.0		96.0	98.0	29.1	27.1	0.831	0.774	5:1	265
	411HR (PA)	Polyamide TFC	10	102	7.0		96.0	98.0	6.8	5.8	0.971	0.829	5:1	76
	815HR	Cellulosic	21	102	32.0		96.0	97.5	24.3	20.6	0.759	0.644	5:1	302
	815SR	Cellulosic	21	102	32.0		92.5	95.0	30.7	27.6	0.959	0.863	5:1	302
	811HR	Cellulosic	21	102	35.0		96.0	97.5	25.0	21.3	0.714	0.609	5:1	302
	811SR	Cellulosic	21	102	35.0		92.5	95.0	31.6	28.4	0.903	0.811	5:1	302
	411HR	Cellulosic	10	102	7.0		96.0	97.5	6.1	5.1	0.871	0.729	5:1	76
	411SR	Cellulosic	10	102	7.0		92.5	95.0	7.6	6.8	1.086	0.971	5:1	76
	416HR	Cellulosic	10	102	7.0		96.0	97.5	6.2	5.3	0.886	0.757	5:1	114
	416SR	Cellulosic	10	102	7.0		92.5	95.0	8.0	6.8	1.143	0.971	5:1	114
	HA5110	Cellulosic HFF	14	42			92.0	94.0	2.5	2.0			2.3:1	6
	HA5230	Cellulosic HFF	15	84			92.0	94.0	15.0	11.0			0.3:1	14
	HA5330	Cellulosic HFF	15	124			92.0	94.0	24.0	20.0			0.3:1	22
	HA8130	Cellulosic HFF	29	132			92.0	94.0	60.0	54.0			0.3:1	55

Table 10.6.5. – Brackish water reverse osmosis membranes by producer.

		TESTING CONDITIONS										OPERATING pH		CLEANING pH					
FACTURER	MODEL#	mg/L NaCl	kPa	% RECOVERY	TEMP. °C	pH	MAX kPa	PRESSURE DROP/ ELEMENT	MAX SDI	MAX NTU	MAX °C	MIN	MAX	MIN	MAX	MAX CONT. OX. CONC.	# ELEMENTS/ VESSEL	MAX RECOVERY/ VESSEL	SHIPPED WEIGHT (kg)
nautics	4040-LSA-CPA2	1500	1551	15	25	7.0	2758	69	4	1	45	3.0	10.0	4.0	10.0	0.1			8
	4040-LSA-NCM1	1500	1551	15	25	6.7	2758	69	4	1	45	3.0	10.0	2.0	11.0	0.0			8
	4040-LST-CPA2	1500	1551	15	25	7.0	2758	69	4	1	45	3.0	10.0	4.0	10.0	0.1			8
	4040-LST-NCM1	1500	1551	15	25	6.7	2758	69	4	1	45	3.0	10.0	3.0	10.0	0.0			8
	4040-MSA-CAB1	2000	2896	10	25	6.0	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0			8
	4040-MSA-CAB2	2000	2896	10	25	6.0	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0			8
	4040-MSA-CAB3	2000	2896	10	25	6.0	4137	97	4	1	40	4.0	6.0	4.0	7.5	1.0			8
	8040-MSY-CAB1	2000	2896	10	25	6.0	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0			18
	8040-MSY-CAB2	2000	2896	10	25	6.0	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0	6	74	18
	8040-MSY-CAB3	2000	2896	10	25	6.0	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0			8
	8040-LSY-CPA2	1500	1551	15	25	7.0	2758	69	4	1	45	3.0	10.0	4.0	10.0	0.1			8
	8040-LSY-NCM1	1500	1551	15	25	7.0	2758	69	4	1	45	3.0	10.0	2.0	11.0	0.0			8
	8540-LSY-NCM1	1500	1551	15	25	7.0	2758	69	4	1	45	3.0	10.0	2.0	11.0	0.0			24
's	CDRO M60 S1	250	1379	8	25	7.0	2758	103	5		35	4.0	7.5	3.0	7.5	1.5			
	CDRO 120 S0	500	2758	20	25	7.0	2758	103	5		35	4.0	7.5	3.0	7.5	1.5			
	CDRC M60 S1	250	1379	15	25	7.0	2758	69	5		45	2.0	11.0	1.0	12.0	0.1			
	SM1B A12 V7	2000	1551	17	25	7.0	4137	69	5		45	2.0	11.0	1.0	12.0	0.1			
	CDRS M60 S1	250	1379	20	25	7.0	2758	69	6		35	4.0	11.0	2.0	12.0	5.0			
nics	815HR (PA)	2000	1550	10	25	8.0	3448	103	5	1	40	3.0	11.0	2.0	12.0	1.0	6		20
	811HR (PA)	2000	1550	10	25	8.0	3448	103	5	1	40	3.0	11.0	2.0	12.0	1.0	6		20
	411HR (PA)	2000	1550	10	25	8.0	3448	138	5	1	40	3.0	11.0	2.0	12.0	1.0	6		5
	815HR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	815SR	2000	2894	10	25	5-6	4134	103	6	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	811HR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	811SR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	411HR	2000	2894	10	25	5-6	4134	138	5	1	40	5.5	6.5	3.0	7.0	1.0	6		5
	411SR	2000	2894	10	25	5-6	4134	138	5	1	40	5.5	6.5	3.0	7.0	1.0	6		5
	416HR	2000	2894	10	25	5-6	4134	138	5	1	40	5.5	6.5	3.0	7.0	1.0	6		5
	416SR	2000	2894	10	25	5-6	4134	138	5	1	40	5.5	6.5	3.0	7.0	1.0	6		5
o	HA5110	500	981	30	25		1471		4		35	3.0	8.0	3.0	8.0	1.0	1	30	11
	HA5230	1500	2942	75	25		3922		4		35	3.0	8.0	3.0	8.0	1.0	1	75	21
	HA5330	1500	2942	75	25		3922		4		35	3.0	8.0	3.0	8.0	1.0	1	75	31
	HA8130	1500	2942	75	25		3922		4		35	3.0	8.0	3.0	8.0	1.0	1	75	100

Table 10.6.6. – Brackish water reverse osmosis membranes by producer.

		VITAL STATISTICS									L/DAY	MODEL # FOR OTHER SIZES			
ACTURER	MODEL#	TEMP°	Cf	Cp	Cr	CM	dPi	dP	A	B	PER CC	2.5" DIA.	4" DIA.	OTHER SIZES	NOTES:
utios	4040-LSA-NCM1	298	25.7	0.3	30.2	27.9	135.7	1482.4	9.339	0.010	0.94				5
	4040-LSA-CPA2	298	25.7	0.3	30.2	27.9	135.7	1482.4	8.872	0.010	0.87				
	8540-LSY-NCM1	298	25.7	0.3	30.2	27.9	135.7	1482.4	9.339	0.010	0.94				5
	8040-LSY-NCM1	298	25.7	0.5	30.1	27.9	134.4	1482.4	8.863	0.019	0.90				
	8040-MSY-CAB2	298	34.2	2.6	37.7	36.0	164.0	2792.5	4.545	0.079	0.90				
	4040-MSA-CAB1	298	34.2	1.0	37.9	36.1	172.0	2792.5	3.599	0.024	0.71				
	4040-MSA-CAB2	298	34.2	0.5	38.0	36.1	174.6	2799.4	2.396	0.008	0.47				
	8040-MSY-CAB1	298	34.2	1.7	37.8	36.0	168.4	2792.5	4.493	0.051	0.99		4040-MSY-CAB1	8060-MSY-CAB1	
	8040-MSY-CAB3	298	34.2	0.7	37.9	36.1	173.8	2792.5	3.602	0.016	0.79		4040-MSY-CAB2	8060-MSY-CAB2	
	4040-MSA-CAB3	298	34.2	0.3	38.0	36.1	175.5	2792.5	2.385	0.005	0.52		4040-MSY-CAB3	8060-MSY-CAB3	
	4040-LST-CPA2	298	25.7	0.3	30.2	27.9	135.7	1482.4	9.698	0.010	1.05				5
	8040-LSY-CPA2	298	25.7	0.3	30.2	27.9	135.7	1482.4	8.635	0.009	1.03		4040-LST-NCM1		
	4040-LST-NCM1	298	25.7	0.3	30.2	27.9	135.7	1482.4	8.755	0.009	1.07		4040-LSA-NCM1		
	CDRO M60 S1	298	4.3	0.3	4.6	4.5	20.4	1275.6	3.119	0.024	0.40				
	CDRO 120 S0	298	8.6	0.6	10.5	9.6	43.9	2654.6	2.463	0.037	0.57				
	SM1B A12 V7	298	4.3	0.1	5.0	4.6	22.4	1310.1	10.914	0.023	1.62		SM1B A12 V7	3.25"-3.4" X 25"	
	CDRC M60 S1	298	34.2	1.4	41.0	37.6	177.8	1482.4	7.181	0.031	0.82	no		8"X40" w/ ZWRO4000 systems	7
	CDRS M60 S1	298	4.3	0.2	5.3	4.8	22.4	1310.1	8.473	0.044	1.13			5.9" X 25" CDRS 120 S0	6
cs	815HR (PA)														
	811HR (PA)													2"X26", 2"X39"	
	411HR (PA)												411HR(PA)	2.5"X14", 2.5"X21", 2.5"X40"	
	815HR														
	815SR														
	811HR														
	811SR														
	411HR														
	411SR														
	416HR														
	416SR														
	HA5230													150 X 840 mm	
	HA5110													140 X 420 mm	
	HA5330													150 X 1240 mm	
	HA8130													295 X 1320 mm	

Table 10.7.1. – Nanofiltration membranes by producer.

							CI REJECT		PROD. (m3/d)		FLUX (m3/m2*d)			MAX FEED
UFACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	DIVALENT REJECTION	MIN	AVE	AVE	MIN	AVE	MIN	MIN CONC:PERM	FLOW (L/min)
ination Systems	CG4040F	Cellulosic	10	102	8.4		80.0	84.0	7.6	6.4	0.9	0.8	5:1	60
	CG8040F	Cellulosic	20	102	31.6		80.0	84.0	27.6	23.5	0.9	0.7	5:1	221
	DK4040F	TFC	10	102	8.4	98.0	50.0	50.0	7.6	7.6	0.9	0.9		
	DK8040F	TFC	20	102	31.6	98.0	50.0	50.0	30.3	30.3	1.0	1.0		
ec	NF40-8040	Polyamide TFC	20	102	28.2	95.0	40.0	45.0	26.5	21.2	0.9	0.8	6.6:1	141
	NF70-4040	Polyamide TFC	10	102	6.4	93.0	20.0	30.0	6.4	5.1	1.0	0.8	6.6:1	34
	NF70-8040	Polyamide TFC	20	102	29.3	75.0	40.0	45.0	26.5	21.2	0.9	0.7	6.6:1	141
Systems	4921	Polyamide TFC	10	102	7.4	95.0	80.0	85.0	4.5	3.9	0.6	0.5	6:1	36
	8921	Polyamide TFC	20	102	30.2	95.0	80.0	85.0	18.2	15.4	0.6	0.5	6:1	145
nautics	4040-LSA-PVD1	Polyvinyl Alcohol	10	102	7.0		80.0	80.0	7.9	6.8	1.1	1.0	5:1	25
	4040-LST-PVD1	Polyvinyl Alcohol	10	102	7.0		80.0	80.0	7.9	6.8	1.1	1.0	5:1	25
	8040-LSY-PVD1	Polyvinyl Alcohol	20	102	33.9		80.0	80.0	41.6	35.4	1.2	1.0	5:1	133
nics	815PR	Cellulosic	20	102	30.0	90.0	90.0	90.0	39.5	35.4	1.3	1.2	5:1	302
	811PR	Cellulosic												
	411PR	Cellulosic	10	102	7.0	90.0	90.0	90.0	9.9	8.9	1.4	1.3	5:1	76
	416PR	Cellulosic	10	102	7.0	90.0	90.0	90.0	6.8	6.1	1.0	0.9	5:1	76
	815NF100	Cellulosic	20	102	30.0	85.0	80.0	80.0	19.7	16.7	0.7	0.6	5:1	302
	411NF100	Cellulosic	10	102	7.0	85.0	80.0	80.0	4.9	4.2	0.7	0.6	5:1	76
	815NF200	Cellulosic	20	102	30.0	80.0	50.0	50.0	29.5	25.1	1.0	0.8	5:1	302
	411NF200	Cellulosic	10	102	7.0	80.0	50.0	50.0	7.4	6.3	1.1	0.9	5:1	76
	815NF300(PA)	Polyamide TFC	20	102	30.0	70.0	45.0	45.0	29.1	26.2	1.0	0.9	5:1	302
	411NF300(PA)	Polyamide TFC	10	102	7.0	70.0	45.0	45.0	7.3	6.4	1.0	0.9	5:1	76

Table 10.7.2. – Nanofiltration membranes by producer.

		TESTING CONDITIONS										OPERATING pH		CLEANING pH					
ACTURER	MODEL#	mg/L NaCl	kPa	% RECOVERY	TEMP. °C	pH	MAX kPa	PRESSURE DROP/ ELEMENT	MAX SDI	MAX NTU	MAX °C	MIN	MAX	MIN	MAX	MAX CONT. OX. CONC.	# ELEMENTS/ VESSEL	MAX RECOVERY/ VESSEL	SHIPPED WEIGHT (kg)
ation Systems	CG4040F	1000	1379	10	25		3103	69		1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	CG8040F	1000	1379	10	25		3103	69		1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	DK4040F	1000	690	10	25		4137	69			50	4.0	11.0	2.0	11.5	2000.0	6		15
	DK8040F	1000	690	10	25		4137	69			50	4.0	11.0	2.0	11.5	2000.0	6		15
	NF40-8040	2000	1551	15	25	8.0	4137	138	5	1	45	2.0	11.0	1.0	12.0	0.5			
	NF70-4040	2000	483	15	25	8.0	1724	138	5	1	35	3.0	9.0	1.0	9.0	0.1			
	NF70-8040	2000	483	15	25	8.0	1724	138	5	1	35	3.0	9.0	1.0	9.0	0.1			
	4921	*500*	*552*	*10*	*25*	*7.5*	*2413*	*69*	*5*	*1*	*45*	*4.0*	*11.0*	*2.5*	*11.0*	*0.0*	*6*	*53*	*5*
stems	8921	500	552	10	25	7.5	2413	69	5	1	45	4.0	11.0	2.5	11.0	0.0	6	53	18
autics	4040-LSA-PVD1	590	1034	25	25	8.0	2758	69			40	2.0	8.0	4.0	10.0	1.0			8
	4040-LST-PVD1	590	1034	25	25	8.0	2758	69			40	2.0	8.0	4.0	10.0	1.0			8
	8040-LSY-PVD1	590	1034	25	25	8.0	2758	69			40	2.0	8.0	2.0	10.0	1.0			21
ics	815PR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	811PR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	411PR	2000	2894	10	25	5-6	4134	138	5	1	40	5.5	6.5	3.0	7.0	1.0	6		5
	416PR	2000	1723	10	25	5-6	4134	138	5	1	40	5.5	6.5	3.0	7.0	1.0	6		5
	815NF100																		
	411NF100																		
	815NF200																		
	411NF200																		
	815NF300(PA)																		
	411NF300(PA)																		

Table 10.7.3. – Nanofiltration membranes by producer.

UFACTURER	MODEL#	VITAL STATISTICS									L/DAY PER CC	MODEL # FOR OTHER SIZES			NOTES:
		TEMP°	Cf	Cp	Cr	CM	dPI	dP	A	B		2.5" DIA.	4" DIA.	OTHER SIZES	
ination Systems	CG4040F	298	17.1	2.7	18.7	17.9	74.5	1310.1	8.482	0.163	0.98				2
	CG8040F	298	17.1	2.7	18.7	17.9	74.5	1310.1	8.195	0.158	0.86		CG4040F	26.25" x 3.88", 25.0" x 3.88"	2
	DK4040F	298	17.1	6.6	18.1	17.6	44.3	620.6	18.188	0.858	0.92		DK4040F		9
	DK8040F	298	17.1	8.6	18.1	17.6	44.3	620.6	19.258	0.908	0.92		DK4040F		9
	NF40-8040	298	34.2	18.8	36.9	35.6	82.3	1413.5	8.158	1.054	0.80	NF40-2540	NF40-4040		
ec	NF70-4040	298	34.2	24.0	36.0	35.1	54.8	344.8	40.079	2.153	0.78	NF70-2540		2.5X14", 2.5X21"	
	NF70-8040	298	34.2	18.8	36.9	35.6	82.3	344.8	39.926	1.017	0.80	NF70-2540	NF70-4040	2.5X14", 2.5X21"	
	4921	298	8.6	1.3	9.4	9.0	37.7	482.7	15.898	0.102	0.55				
Systems	8921	298	8.6	1.3	9.4	9.0	37.7	482.7	15.653	0.101	0.55		4921		
enautics	4040-LSA-PVD1	298	10.1	2.0	12.8	11.4	46.2	965.3	14.368	0.244	0.99				4
	4040-LST-PVD1	298	10.1	2.0	12.8	11.4	46.2	965.3	14.368	0.244	0.99				4
	8040-LSY-PVD1	298	10.1	2.0	12.8	11.4	46.2	965.3	15.465	0.263	1.28				4
onics	815PR														
	811PR														
	411PR														
	416PR														
	815NF100														
	411NF100														
	815NF200														
	411NF200														
	815NF300(PA)														
	411NF300(PA)														

Table 10.8.1. – Ultrafiltration membranes by producer.

										TEST CONDITIONS						
FACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	MWCO	CONFIG.	AVERAGE PROD. (m3/Day)	MAX FEED FLOW (L/min)	TEMP. °C	kPa	pH	% RECOVERY	TDS (mg/L)	MAX (kPa)	PRES. DROP/ ELEMENT
chnology Corp.	UFP	Polysulfone Bundles	10	109	3.7	10000	HF	9.1		25	207	7		Clean	241	
	KC-840-010 SL	Polyethersulfone	20	102	25.5	10000	SW	109.0	96.7	25	345	7	90	Clean	689	138
	KC-440-010 SL	Polyethersulfone	10	102	5.6	10000	SW	43.6	38.7	25	345	7	90	Clean	689	138
ation Systems	G-5	Thin Film Composite	10	102	8.4	1000	SW									
	G-10	Thin Film Composite	10	102	8.4	1500	SW			25			70-90	255		
	G-20	Thin Film Composite	10	102	8.4	2500	SW			25			70-90	255		
	G-50	Thin Film Composite	10	102	8.4	1100	SW			25			70-90	255		
s	CDUF H25 01	Thin Film Composite	10	64	4.6	10000	SW	26.5		25	345	7		Clean	689	
	CDUF 025 01	Thin Film Composite	10	64	4.6	10000	SW	10.9		25	345	7		Clean	655	12
ics	815PT1	Polysulfone	20	102	30	1000	SW	49.8	302	25	689	5-6	10	23.5		138
	815PT2	Polysulfone	20	102	30	20000	SW	49.8	302	25	345	5-6	10	20.0		138
	815PT3	Polysulfone	20	102	30	50000	SW	49.8	302	25	172	5-6	10	73.7		138
	815PT4	Polysulfone	20	102	30	100000	SW	49.8	302	25	172	5-6	10	47.7		138
	411PT1	Polysulfone	10	102	7	1000	SW	14.3	76	25	689	5-6	10	38.2		
	411PT2	Polysulfone	10	102	7	20000	SW	14.3	76	25	345	5-6	10			
	411PT3	Polysulfone	10	102	7	50000	SW	14.3	76	25	172	5-6	10			
	411PT4	Polysulfone	10	102	7	100000	SW	14.3	76	25	172	5-6	10			
	815PT1	Polyvinyldine Flouride	20	102	30	1000	SW	49.8	302	25	689	5-6	10			138
	815PT2	Polyvinyldine Flouride	20	102	30	20000	SW	49.8	302	25	345	5-6	10			138
	815PT3	Polyvinyldine Flouride	20	102	30	50000	SW	49.8	302	25	172	5-6	10			138
	411PT1	Polyvinyldine Flouride	10	102	7	1000	SW	14.3	76	25	689	5-6	10			138
	411PT2	Polyvinyldine Flouride	10	102	7	20000	SW	14.3	76	25	345	5-6	10			
	411PT3	Polyvinyldine Flouride	10	102	7	50000	SW	14.3	76	25	172	5-6	10			
	815PT1	Cellulosic	20	102	30	1000	SW	31.4	302	25	689	5-6	10			138
	815PT2	Cellulosic	20	102	30	20000	SW	31.4	302	25	345	5-6	10			138
	815PT3	Cellulosic	20	102	30	50000	SW	31.4	302	25	172	5-6	10			138
	411PT1	Cellulosic	10	102	7	1000	SW	7.5	76	25	689	5-6	10			207
	411PT2	Cellulosic	10	102	7	20000	SW	7.5	76	25	345	5-6	10			207
	411PT3	Cellulosic	10	102	7	50000	SW	7.5	76	25	172	5-6	10			207
	817UP1	Polysulfone	20	102	30	1000	SW	54.7	454	25	689	5-6	10			68.9
	817UP2	Polysulfone	20	102	30	20000	SW	54.7	454	25	345	5-6	10			68.9
	817UP3	Polysulfone	20	102	30	50000	SW	54.7	454	25	172	5-6	10			68.9
	416UP1	Polysulfone	10	102	7	1000	SW	17.3	113	25	689	5-6	10			68.9
	416UP2	Polysulfone	10	102	7	20000	SW	17.3	113	25	345	5-6	10			68.9
	416UP3	Polysulfone	10	102	7	30000	SW	17.3	113	25	172	5-6	10			68.9
	817UP1	Polyvinyldine Flouride	20	102	30	1000	SW	54.7	454	25	689	5-6	10			68.9
	817UP2	Polyvinyldine Flouride	20	102	30	20000	SW	54.7	454	25	345	5-6	10			68.9
	817UP3	Polyvinyldine Flouride	20	102	30	50000	SW	54.7	454	25	172	5-6	10			68.9
	416UP1	Polyvinyldine Flouride	10	102	7	1000	SW	17.3	113	25	689	5-6	10			68.9
	416UP2	Polyvinyldine Flouride	10	102	7	20000	SW	17.3	113	25	345	5-6	10			68.9
	416UP3	Polyvinyldine Flouride	10	102	7	50000	SW	17.3	113	25	172	5-6	10			68.9
-Poulenc	Carbosep U252	Zirconia/metalic Ox	10	122	5.7	10000	TUBE	ERR							1000	
	HF 132-20-GM80	Acrylic	13	109	12.3	80000	HFF	ERR	21.2	25	689		90	RO-DI	689	172
	HF-132-20-PM10	Polysulfone	13	102	12.3	10000	HFF	ERR	16.0	25	689		95	RO-DI	689	241

Table 10.8.2. – Ultrafiltration membranes by producer.

UFACTURER	MODEL#	ELEMENTS/ VESSEL	MAX RECOVERY	AVE. FLUX m3/m2*day	MAX °C	OPERATING pH MIN	OPERATING pH MAX	CLEANING pH MIN	CLEANING pH MAX	NOTES:	PRETREATMENT
Technology Corp.	UFP			2.44	80	2.0	13.0	2.0	13.0	3	REJECTION RATE
)	KC-840-010 SL			4.27	50	2.0	10.0	1.5	13.0	1,4	1) 25-50μm filter
	KC-440-010 SL			7.82	50	2.0	10.0	1.5	10.0	4	2) SDI<1, Prefiltration:10μm
ination Systems	G-5	4	70-90							3,8	
	G-10	4	70-90							3,5	4) 99.99%:pyrogens, colloids & bacteria
	G-20	4	70-90							3,6	5) 95% Cobalamine MW 1355, 72% Streptomycyn
	G-50	4	70-90							3,7	6) 96.4% Silica, 70% THMFP
re	CDUF H25 01			5.70	35	2.0	12.0	1.0	13.0	2,3,4	3) Dechlorination
	CDUF 025 01			2.35	43	1.0	13.0	1.0	13.0	4	
nics	815PT1	3	70 - 90	1.7							
	815PT2	3	70 - 90	1.7							
	815PT3	3	70 - 90	1.7							
	815PT4	3	70 - 90	1.7							
	411PT1	3	70 - 90	2.0							
	411PT2	3	70 - 90	2.0							
	411PT3	3	70 - 90	2.0							
	411PT4	3	70 - 90	2.0							
	815PT1	3	70 - 90	1.7							
	815PT2	3	70 - 90	1.7							
	815PT3	3	70 - 90	1.7							
	411PT1	3	70 - 90	2.0							
	411PT2	3	70 - 90	2.0							
	411PT3	3	70 - 90	2.0							
	815PT1	3	70 - 90	1.0							
	815PT2	3	70 - 90	1.0							
	815PT3	3	70 - 90	1.0							
	411PT1	3	70 - 90	1.1							
	411PT2	3	70 - 90	1.1							
	411PT3	3	70 - 90	1.1							
	817UP1	3 - 4	95 - 100	1.8							
	817UP2	3 - 4	95 - 100	1.8							
	817UP3	3 - 4	95 - 100	1.8							
	416UP1	3 - 4	95 - 100	2.5							
	416UP2	3 - 4	95 - 100	2.5							
	416UP3	3 - 4	95 - 100	2.5							
	817UP1	3 - 4	95 - 100	1.8							
	817UP2	3 - 4	95 - 100	1.8							
	817UP3	3 - 4	95 - 100	1.8							
	416UP1	3 - 4	95 - 100	2.5							
	416UP2	3 - 4	95 - 100	2.5							
	416UP3	3 - 4	95 - 100	2.5							
-Poulenc	Carbosep U252				300	0.0	14.0	0.0	14.0		7) 88% THMFP, 94% Silica
	HF 132-20-GM80				45	1.0	13.0	1.0	13.0		8) 99% Cobalamine MW 1355
	HF-132-20-PM10	1	90		75	1.0	13.0	1.0	13.0		

Table 10.9.1. – Microfiltration membranes by producer.

UFACTURER	MODEL#	TYPE/MODE	COMPOSITION	RETENTION RATING (MICRONS)	INSIDE DIA.(cm)	OUTSIDE DIA.(cm)	RETENTION RATINGS AVAILABLE (MICRONS)
nics	Flotrex-PN	PCF/CF	Polypropylene	0.2	2.5	7	0.2-30
	Flotrex-PN	PCF/CF	Polypropylene	0.45	2.5	7	0.2-30
	Flotrex-PN	PCF/CF	Polypropylene	1	2.5	7	0.2-30
	Flotrex-GF	PCF/CF	Glass Microfiber	0.45	2.5	7	0.45-40
	Flotrex-GF	PCF/CF	Glass Microfiber	1	2.5	7	0.45-40
	Memtrex-PM	PCF/CF	Polypropylene	0.1	2.5	7	0.1, 0.2
	Memtrex-PM	PCF/CF	Polypropylene	0.2	2.5	7	0.1, 0.2
	Memtrex-NM	PCF/CF	Nylon	0.2	2.5	7	0.2, 0.45
	Memtrex-NM	PCF/CF	Nylon	0.45	2.5	7	0.2, 0.45
	Memtrex-PC	PCF/CF	Polycarbonate	0.05	2.5	7	0.05-0.45
	Memtrex-PC	PCF/CF	Polycarbonate	0.1	2.5	7	0.05-0.45
	Memtrex-PC	PCF/CF	Polycarbonate	0.2	2.5	7	0.05-0.45
	Memtrex-PC	PCF/CF	Polycarbonate	0.45	2.5	7	0.05-0.45
	Memtrex-PS	PCF/CF	Polysulfone	0.2	2.5	7	0.2-0.65
	Memtrex-PS	PCF/CF	Polysulfone	0.45	2.5	7	0.2-0.65
	Memtrex-PS	PCF/CF	Polysulfone	0.65	2.5	7	0.2-0.65
	Memtrex-FE	PCF/CF	Teflon	0.1	2.5	7	0.1-0.65
	Memtrex-FE	PCF/CF	Teflon	0.2	2.5	7	0.1-0.65
	Memtrex-FE	PCF/CF	Teflon	0.45	2.5	7	0.1-0.65
	Memtrex-FE	PCF/CF	Teflon	0.65	2.5	7	0.1-0.65
	Hytrex II GX	DF/DE	Polypropylene	1	1	6.35	1-75
	Hytrex II RX	DF/DE	Polypropylene	1	3.5	7	1-75
	Selex	DF/DE	Polypropylene	1	2.5	6.35	1-30
ure	Rogard II	DF/DE	Polypropylene	1	2.5	6.35	5-75
	Polygard-CR	DF/DE	Polypropylene	0.5	2.5	7	1-99
	Ionpure DI	DF/DE	Polyvinyl	0.1		7	0.1, 0.2
	Ionpure PV	DF/DE	Polyvinyl	0.2		7	0.2
Technology	MF Cartridges	HFB/CF	Polysulfone	0.1	0.1	10	0.1-0.65
ne Poulenc	Kerasep K09-W	TM/CF	Ceramic	0.2	2.5*	20*	0.2, 0.4, 1

Table 10.9.2. – Microfiltration membranes by producer.

NUFACTURER	MODEL#	AREA (sq.m/25cm)	LENGTHS (cm)	MAX kPa	MAX kPa Reverse	FLOW FACTOR (L/min*kPa)	MAX OPERATING TEMP. °C	MAXIMUM TEMP. °C
onics	Flotrex-PN	0.48	25, 50, 75, 100	414	275	5.2	60	82 at 69 kPa
	Flotrex-PN	0.57	25, 50, 75, 100	414	275	10.0	60	82 at 69 kPa
	Flotrex-PN	0.62	25, 50, 75, 100	414	275	12.9	60	82 at 69 kPa
	Flotrex-GF	0.52	25, 50, 75, 100	552	414	3.0	88	82 at 69 kPa
	Flotrex-GF	0.45	25, 50, 75, 100	552	414	4.2	88	82 at 69 kPa
	Memtrex-PM	0.69	25, 50, 75, 100	414	275	0.3	50	82 at 69 kPa
	Memtrex-PM	0.69	25, 50, 75, 100	414	275	1.6	50	82 at 69 kPa
	Memtrex-NM	0.65	25, 50, 75, 100	414	207	0.6	49	82 at 69 kPa
	Memtrex-NM	0.65	25, 50, 75, 100	414	207	1.1	49	82 at 69 kPa
	Memtrex-PC	1.60	25, 50, 75, 100	690	69	0.1	79	88 at 69 kPa
	Memtrex-PC	1.60	25, 50, 75, 100	690	69	0.4	79	88 at 69 kPa
	Memtrex-PC	1.60	25, 50, 75, 100	690	69	0.5	79	88 at 69 kPa
	Memtrex-PC	1.60	25, 50, 75, 100	690	69	0.6	79	88 at 69 kPa
	Memtrex-PS	0.46	25, 50, 75, 100	414	345	0.6	82	82 at 69 kPa
	Memtrex-PS	0.46	25, 50, 75, 100	414	345	2.7	82	82 at 69 kPa
	Memtrex-PS	0.46	25, 50, 75, 100	414	345	4.6	82	82 at 69 kPa
	Memtrex-FE	0.52	25, 50, 75, 100	414	275	0.3	82	82 at 69 kPa
	Memtrex-FE	0.52	25, 50, 75, 100	414	275	0.6	82	82 at 69 kPa
	Memtrex-FE	0.52	25, 50, 75, 100	414	275	0.9	82	82 at 69 kPa
	Memtrex-FE	0.52	25, 50, 75, 100	414	275	1.1	82	82 at 69 kPa
	Hytrex II GX	N/A	3 - 127	414	207	0.6	60	82 at 69 kPa
	Hytrex II RX	N/A	3 - 127	414	207	0.6	60	82 at 69 kPa
	Selex	N/A	3 - 102	414	207	0.2	60	82 at 69 kPa
ure	Rogard II	N/A	25, 50, 75, 100	482		6.6	90	126 for 20 min
	Polygard-CR	N/A	25, 50, 75	482		4.4	80	20 min at 120, 100 kPa
	Ionpure DI	0.65	25, 50, 75, 100	345		0.6	90	Autoclavable
	Ionpure PV	0.62	25, 50, 75	550		0.9	90	Autoclavable
Technology	MF Cartridges	0.70	14, 25, 43	207		1.1	80	Autoclavable
ne Poulenc	Kerasep K09-W	1.00	855	5000	5000	0.04	400	Autoclavable

Table 10.10.1. – 8-inch seawater reverse osmosis membranes by composition.

FACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	DIVALENT REJECTION	Cl REJECT		PROD. (cu.m/d)		FLUX (m3/m2*d)		MIN CONC:PERM	MAX FEED FLOW (L/min)
							MIN.	AVE.	AVE.	MIN.	AVE.	MIN.		
t Permasep	B-106880T	Aramid HFF	20	206	901.2	99.6	98.7	99.2	53.0	44.3	0.059	0.049		420
	B-10 6835T	Aramid HFF	20	150	450.6	99.6	98.7	99.2	26.5	22.5	0.059	0.050	4.2	60
:	SW30-8040	Polyamide TFC	20	102	30.7		98.6	99.1	22.7	19.3	0.741	0.630	9:1	181
	SW30HR-8040	Polyamide TFC	20	102	30.7		99.2	99.4	15.1	12.9	0.494	0.420	11:1	151
ystems	2021 SS	Polyether Urea TFC	20	102	30.2		99.2	99.4	15.1	12.9	0.502	0.426	13:1	173
	2021 HF	Polyether Urea TFC	20	102	30.2		98.6	99.0	22.7	19.3	0.752	0.639	13:1	259
	2821 HP	Polyamide TFC	20	102	30.2		99.2	99.4	18.9	16.1	0.627	0.533	13:1	216
autics	8040-HSY-SWC1	Polyamide TFC	20	102	29.3		99.2	99.5	18.9	16.1	0.647	0.550	9:1	151
o	HM8255	Cellulosic HFF	30	264			99.2	99.4	27.5	25.0			2.3:1	64
	HM9255	Cellulosic HFF	36	267			99.2	99.4	35.0	32.0			2.3:1	81
	HR5255	Cellulosic HFF	15	82			99.2	99.4	3.0	2.4			2.3:1	7
	HR5355	Cellulosic HFF	15	123			99.2	99.4	5.0	4.0			2.3:1	12
	HR8355	Cellulosic HFF	30	133			99.2	99.4	12.0	10.0			2.3:1	28

Table 10.10.2. – 8-inch seawater reverse osmosis membranes by composition.

		TESTING CONDITIONS											OPERATING pH		CLEANING pH		MAX	#	MAX	SHIPPED
UFACTURER	MODEL#	ppm NaCl	kPa	% RECOVERY	TEMP. °C	pH	MAX kPa	PRESSURE DROP/ ELEMENT	MAX SDI	MAX NTU	MAX °C	MIN	MAX	MIN	MAX	CONT. OX. CONC.	ELEMENTS/ VESSEL	REC./ VESSEL	WEIGHT (kg)
nt Permasep	B-106880T	35000	6895	35	25		8274	41	4		40	4.0	9.0	2.3	11.9	0.0	2	50	113
	B-10 6835T	35000	6895	35	25		8274	41	4		40	4.0	9.0	2.3	11.9	0.0	1	50	59
ic	SW30-8040	35000	5516	10	25	8	6895	138	5	1	45	2.0	11.0	1.0	12.0	0.1	6		
	SW30HR-8040	32000	5516	8	25	8	6895	138	5	1	45	2.0	11.0	1.0	12.0	0.1	6		
Systems	2021 SS	32800	5516	7	25	5.7	6896	69		1	45	5.0	10.0	4.5	11.0	0.0	6	42	18
	2021 HF	32800	5516	7	25	5.7	6895	69		1	45	5.0	10.0	4.5	11.0	0.0	6	42	16
	2821 HP	32800	5516	7	25	7.5	6895	69		1	45	4.0	11.0	2.5	11.0	0.0	6	42	18
nautics	8040-HSY-SWC1	32000	5516	10	25	7	6895	69	4	1	45	3.0	10.0	4.0	10.0	0.1	7		16
o	HM8255	35000	5393	30	25		6374		4		40	3.0	8.0	3.0	8.0	1.0	2	30	205
	HM9255	35000	5393	30	25		6864		4		40	3.0	8.0	3.0	8.0	1.0	2	30	310
	HR5255	35000	5393	30	25		5884		4		40	3.0	8.0	3.0	8.0	1.0	1	30	23
	HR5355	35000	5393	30	25		5884		4		40	3.0	8.0	3.0	8.0	1.0	1	30	32
	HR8355	35000	5393	30	25		5884		4		40	3.0	8.0	3.0	8.0	1.0	1	30	125

Table 10.10.3. – 8-inch seawater reverse osmosis membranes by composition.

		VITAL STATISTICS									L/DAY	MODEL # FOR OTHER SIZES			
FACTURER	MODEL#	TEMP°	Cf	Cp	Cr	CM	dPI	dP	A	B	PER CC	2.5" DIA.	4" DIA.	OTHER SIZES	NOTES:
t Permasep	B-10 6880T	298	598.9	4.8	918.8	758.9	3514.2	6853.6	0.204	0.000	0.79		B-10 6410, 6440		
	B-10 6835T	298	598.9	4.8	918.8	758.9	3514.2	6853.6	0.204	0.000	0.55			5x23", 5x50"	
	SW30-8040	298	598.9	5.4	664.9	631.9	2919.7	5378.1	3.488	0.006	0.69	SW30-2540	SW30-4040		
	SW30HR-8040	298	547.6	3.3	594.9	571.2	2646.8	5378.1	2.093	0.003	0.46	SW30HR-2540	SW30HR-4040		
ystems	2021 SS	298	561.3	3.4	603.3	582.3	2697.8	5447.1	2.111	0.003	0.46		1021 SS		
	2021 HF	298	561.3	5.6	603.1	582.2	2687.0	5447.1	3.154	0.007	0.69				
	2821 HP	298	561.3	3.4	603.3	582.3	2697.8	5447.1	2.639	0.004	0.57				
autics	8040-HSY-SWC1	298	547.6	2.7	608.1	577.8	2680.2	5447.1	2.706	0.003	0.58				5
	HM8255	298	598.9	3.6	854.0	726.5								298 X 2640 mm	
	HM9255	298	598.9	3.6	854.0	726.5								360 X 2665 mm	
	HR5255	298	598.9	3.6	854.0	726.5								153 X 825 mm	
	HR5355	298	598.9	3.6	854.0	726.5								153 X 1230 mm	
	HR8355	298	598.9	3.6	854.0	726.5								305 X 1330 mm	

Table 10.11.1. – 8-inch brackish water reverse osmosis membranes by composition.

UFACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	DIVALENT REJECTION	CI REJECT		PROD. (m3/d)		FLUX (m3/m2•d)		MIN CONC:PERM	MAX FEED FLOW (L/min)
							MIN	AVE	AVE	MIN	AVE	MIN		
nation Systems	CD8040F	Cellulosic	20	102	31.6		96.5	97.5	22.7	19.3	0.7	0.6	5:1	181
	CE8040F	Cellulosic	20	102	31.6		95.0	96.0	30.3	25.7	1.0	0.8	5:1	242
	ND8040F	Thin Film Composite	20	102	31.6	98.5	97.0	98.0	25.0	21.2	0.8	0.7	5:1	200
	SE8040F	Thin Film Composite	20	102	31.6		97.5	98.5	29.1	24.8	0.9	0.8	5:1	233
	SG8040F	Thin Film Composite	20	102	31.6	99.5	97.0	98.0	29.1	24.8	0.9	0.8	5:1	233
	SH8040F	Thin Film Composite	20	102	31.6	99.5	95.0	99.0	29.1	24.8	0.9	0.8	5:1	233
nt Permasep	B-9 0880	Aramid HFF	20	178	845.4	97.5	90.0	95.0	140.1	126.0	0.2	0.2	53	176
	B-9 0840	Aramid HFF	20	122	427.4	96.0	90.0	92.0	60.6	54.5	0.1	0.1	26.5	88
c	BW30-8040	Polyamide TFC	20	102	30.7		96.0	98.0	28.4	24.1	0.9	0.8	6:1	151
Systems	8021 MP	Polyether Urea	20	102	30.2		98.0	99.0	30.7	26.1	1.0	0.9	9:1	245
	8021 LP	Polyether Urea	20	102	30.2		96.0	97.5	26.5	22.5	0.9	0.7	9:1	212
	8221 SD	Cellulosic	20	102	30.2		94.0	95.5	30.3	25.7	1.0	0.9	9:1	242
	8221 HR	Cellulosic	20	102	30.2		97.0	98.0	23.8	20.3	0.8	0.7	9:1	190
	8231 SD	Cellulosic	20	152	48.8		94.0	95.5	49.2	41.8	1.0	0.9	5:1	246
	8231 HR	Cellulosic	20	152	48.8		97.0	98.0	39.7	33.8	0.8	0.7	5:1	198
	8821 LP	Polyamide TFC	20	102	30.2		96.5	98.5	28.4	24.1	0.9	0.8	9:1	227
nautica	8540-LSY-NCM1	Polyamide TFC	22	102	39.0		98.0	99.0	39.7	33.8	1.0	0.9	5:1	212
	8040-LSY-NCM1	Polyamide TFC	20	102	33.9		98.0	99.0	34.1	29.0	1.0	0.9	5:1	181
	8040-MSY-CAB2	Cellulosic	20	102	31.6		97.0	98.0	25.7	21.9	0.8	0.7	5:1	206
	8040-MSY-CAB3	Cellulosic	20	102	31.6		98.5	99.0	17.0	14.5	0.5	0.5	5:1	136
	8040-MSY-CAB1	Cellulosic	20	102	31.6		92.5	95.0	32.2	27.3	1.0	0.9	5:1	257
	8040-LSY-CPA2	Polyamide TFC	20	102	30.2		99.0	99.0	34.1	29.0	1.1	1.0	5:1	181
nics	815HR (PA)	Polyamide TFC	21	102	32.0		96.0	98.0	28.4	24.1	0.9	0.8	5:1	265
	811HR (PA)	Polyamide TFC	21	102	35.0		96.0	98.0	29.1	27.1	0.8	0.8	5:1	265
	815HR	Cellulosic	21	102	32.0		96.0	97.5	24.3	20.6	0.8	0.6	5:1	302
	815SR	Cellulosic	21	102	32.0		92.5	95.0	30.7	27.6	1.0	0.9	5:1	302
	811HR	Cellulosic	21	102	35.0		96.0	97.5	25.0	21.3	0.7	0.6	5:1	302
	811SR	Cellulosic	21	102	35.0		92.5	95.0	31.6	28.4	0.9	0.8	5:1	302

Table 10.11.2. – 8-inch brackish water reverse osmosis membranes by composition.

10.11.2: 8" DIA. BRACKISH WATER MEMBRANES

		TESTING CONDITION						MAX				OPERATING pH		CLEANING pH					
FACTURER	MODEL#	*mg/L* NaCl	kPa	*%* RECOVERY	*TEMP.* °C	pH	*MAX* kPa	PRESSURE *DROP/* ELEMENT	*MAX* SDI	*MAX* NTU	*MAX* °C	MIN	MAX	MIN	MAX	MAX CONT. *OX.* CONC.	# *ELEMENTS* VESSEL	MAX *REC./* VESSEL	SHIPPED *WEIGHT* (kg)
ation Systems	CD8040F	1000	2758	10	25		3103	69	5	1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	CE8040F	1000	2758	10	25		3103	69	5	1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	ND8040F	1000	2758	10	25		4137	69	5	1	50	4.0	11.0	2.0	11.5	Low	6	53	16
	SE8040F	1000	2758	10	25		4137	69	5	1	50	4.0	11.0	2.0	11.5	Low	6	53	16
	SG8040F	1000	1379	10	25	8	4137	69	5	1	50	4.0	11.0	2.0	11.5	0.1	6	53	16
	SH8040F	1000	1034	10	25	8	4137	69	5	1	50	4.0	11.0	2.0	11.5	0.1	6	53	16
t Permasep	B-9 0880	1500	2758	75	25		2758	60	4		40	4.0	11.0	2.3	11.9	0.0	2	50	113
	B-9 0840	1500	2758	75	25		2758	60	3		40	4.0	11.0	2.3	11.9	0.0	1	50	59
	BW30-8040	2000	1544	15	25	8	4137	138	5	1	45	2.0	11.0	1.0	12.0	0.1	6		0
ystems	8021 MP	2000	2896	10	25	6	4137	69		1	45	5.0	10.0	4.5	11.0	0.0	6	53	18
	8021 LP	2000	1551	10	25	6	2413	69		1	45	5.0	10.0	4.5	11.0	0.0	6	53	18
	8221 SD	2000	2896	10	25	6	4137	103		1	40	3.0	7.0	3.0	8.0	1.0	6	53	18
	8221 HR	2000	2896	10	25	6	4137	103		1	40	4.0	6.0	3.0	7.0	1.0	6	53	18
	8231 SD	2000	2896	16	25	6	4137	103		1	40	3.0	7.0	2.5	8.0	1.0	4	55	26
	8231 HR	2000	2896	16	25	6	4137	103		1	40	4.0	6.0	3.0	7.0	1.0	4	55	26
	8821 LP	2000	1551	10	25	8	4137	69		1	45	4.0	11.0	2.5	11.0	0.0	6	53	18
autics	8540-LSY-NCM1	1500	1551	15	25	7	2758	69	4	1	45	3.0	10.0	2.0	11.0	0.0			24
	8040-LSY-NCM1	1500	1551	15	25	7	2758	69	4	1	45	3.0	10.0	2.0	11.0	0.0	6	74	18
	8040-MSY-CAB2	2000	2896	10	25	6	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0			8
	8040-MSY-CAB3	2000	2896	10	25	6	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0			8
	8040-MSY-CAB1	2000	2896	10	25	6	4137	103	4	1	40	4.0	6.0	4.0	7.5	1.0			8
	8040-LSY-CPA2	1500	1551	15	25	7	2758	69	4	1	45	3.0	10.0	4.0	10.0	0.1			18
ics	815HR (PA)	2000	1550	10	25	8	3448	103	5	1	40	3.0	11.0	2.0	12.0	1.0	6		20
	811HR (PA)	2000	1550	10	25	8	3448	103	5	1	40	3.0	11.0	2.0	12.0	1.0	6		20
	815HR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	815SR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	811HR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	811SR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20

Table 10.11.3. – 8-inch brackish water reverse osmosis membranes by composition.

		VITAL STATISTICS									L/DAY PER CC	MODEL # FOR OTHER SIZES			
UFACTURER	MODEL#	TEMP°	Cf	Cp	Cr	CM	dPI	dP	A	B		2.5" DIA.	4" DIA.	OTHER SIZES	NOTES:
ination Systems	CD8040F	293	17.1	0.4	19.0	18.0	85.0	2689.1	3.196	0.017	0.71	no	CD4040F	26.25x3.88", 25.0x3.88"	
	CE8040F	298	17.1	0.7	18.9	18.0	85.1	2689.1	4.262	0.038	0.95		CE4040F	25.0x3.88", 26.25x3.88"	2
	ND8040F	298	17.1	0.3	19.0	18.0	86.9	2689.1	3.518	0.015	0.78		ND4040F	26.25" x 3.88"	3
	SE8040F	298	17.1	0.3	19.0	18.0	87.3	2689.1	4.105	0.013	0.91		SE4040F	25.00x3.88", 26.25x3.88"	1
	SG8040F	298	17.1	0.3	19.0	18.0	86.9	1310.1	8.732	0.018	0.88		SG4040F	26.25x3.88", 25.0x3.88"	
	SH8040F	298	17.1	0.2	19.0	18.1	87.8	965.3	12.171	0.009	0.88		SH4040F	SH4026F, SH4025T	
ont Permasep	B-9 0880	298	25.7	1.3	98.8	62.2	299.2	2698.0	0.799	0.003	2.43				
	B-9 0840	298	25.7	2.1	96.5	61.1	289.8	2698.0	0.681	0.005	1.53		B-9 0410, 0420, 0440	10x55" 0040	
o	BW30-8040	298	34.2	0.7	40.1	37.2	179.1	1406.6	8.732	0.017	0.86	BW30-2540	BW30-4040		
Systems	8021 MP	298	34.2	0.3	38.0	36.1	175.5	2827.0	4.433	0.010	0.93		4021 MP		
	8021 LP	298	34.2	0.9	37.9	36.1	172.9	1482.4	7.757	0.021	0.80		4021 LP		
	8221 SD	298	34.2	1.5	37.9	36.0	169.3	2792.5	4.426	0.045	0.92		4221 SD		
	8221 HR	298	34.2	0.7	37.9	36.1	173.8	2792.5	3.491	0.015	0.72		4221 HR		
	8231 SD	298	34.2	1.5	40.4	37.3	175.7	2792.5	4.463	0.043	1.00		4231 SD		
	8231 HR	298	34.2	0.7	40.6	37.4	180.3	2792.5	3.611	0.015	0.80		4231 HR		
	8821 LP	298	34.2	0.5	38.0	36.1	174.6	1482.4	8.322	0.014	0.86		4821 LP		
neutics	8540-LSY-NCM1	298	25.7	0.3	30.2	27.9	135.7	1482.4	8.755	0.009	1.07		4040-LSA-NCM1		
	8040-LSY-NCM1	298	25.7	0.3	30.2	27.9	135.7	1482.4	8.635	0.009	1.03		4040-LST-NCM1		
	8040-MSY-CAB2	298	34.2	0.7	37.9	36.1	173.8	2792.5	3.602	0.016	0.79		4040-MSY-CAB2	8060-MSY-CAB2	
	8040-MSY-CAB3	298	34.2	0.3	38.0	36.1	175.5	2792.5	2.385	0.005	0.52		4040-MSY-CAB3	8060-MSY-CAB3	
	8040-MSY-CAB1	298	34.2	1.7	37.8	36.0	168.4	2792.5	4.493	0.051	0.99		4040-MSY-CAB1	8060-MSY-CAB1	
	8040-LSY-CPA2	298	25.7	0.3	30.2	27.9	135.7	1482.4	9.698	0.010	1.05				5
nics	815HR (PA)														
	811HR (PA)													2"X26", 2"X39"	
	815HR														
	815SR														
	811HR														
	811SR														

Table 10.12.1. – 8-inch nanofiltration membranes by composition.

JFACTURER	MODEL#	COMPOSITION	DIAMETER (cm)	LENGTH (cm)	AREA (sq. m)	DIVALENT REJECTION	Cl REJECT %		PROD. (m3/d)		FLUX (m3/m2*d)		MIN CONC:PERM	MAX FEED FLOW (L/min)
							MIN	AVE	AVE	MIN	AVE	MIN		
ination Systems	CG8040F	Cellulosic	20	102	31.6	80.0	84.0	84.0	27.6	23.5	0.9	0.7	5:1	221
	DK8040F	TFC	20	102	31.6	50.0	50.0	50.0	30.3	30.3	1.0	1.0		
ic	NF40-8040	Polyamide TFC	20	102	28.2	40.0	45.0	45.0	26.5	21.2	0.9	0.8	6.6:1	141
	NF70-8040	Polyamide TFC	20	102	29.3	40.0	45.0	45.0	26.5	21.2	0.9	0.7	6.6:1	141
Systems	8921	Polyamide TFC	20	102	30.2	80.0	85.0	85.0	18.2	15.4	0.6	0.5	6:1	145
nautics	8040-LSY-PVD1	Polyvinyl Alcohol	20	102	33.9	80.0	80.0	80.0	41.6	35.4	1.2	1.0	5:1	133
nics	815PR	Cellulosic	20	102	30.0	90.0	90.0	90.0	39.5	35.4	1.3	1.2	5:1	302
	811PR	Cellulosic												
	815NF100	Cellulosic	20	102	30.0	85.0	80.0	80.0	19.7	16.7	0.7	0.6	5:1	302
	815NF200	Cellulosic	20	102	30.0	80.0	50.0	50.0	29.5	25.1	1.0	0.8	5:1	302
	815NF300(PA)	Polyamide TFC	20	102	30.0	70.0	45.0	45.0	29.1	26.2	1.0	0.9	5:1	302

Table 10.12.2. – 8-inch nanofiltration membranes by composition.

		TESTING CONDITION										OPERATING pH		CLEANING pH					
UFACTURER	MODEL#	mg/L NaCl	kPa	% RECOVERY	TEMP. °C	pH	MAX kPa	PRESSURE DROP/ ELEMENT	MAX SDI	MAX NTU	MAX °C	MIN	MAX	MIN	MAX	MAX CONT. OX. CONC.	# ELEMENTS VESSEL	MAX REC./ VESSEL	SHIPPED WEIGHT (kg)
lination Systems	CG8040F	1000	1379	10	25		3103	70		1	35	5.0	6.5	3.0	8.0	1.0	6	53	15
	DK8040F	1000	690	10	25		4137	70			50	4.0	11.0	2.0	11.5	2000.0	6		15
tec	NF40-8040	2000	1551	15	25	8	4137	140	5	1	45	2.0	11.0	1.0	12.0	0.5			
	NF70-8040	2000	483	15	25	8	1724	140	5	1	35	3.0	9.0	1.0	9.0	0.1			
Systems	8921	500	552	10	25	7.5	2413	70	5	1	45	4.0	11.0	2.5	11.0	0.0	6	53	726
anautics	8040-LSY-PVD1	590	1034	25	25	8	2758	70			40	2.0	8.0	2.0	10.0	1.0			21
onics	815PR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	811PR	2000	2894	10	25	5-6	4134	103	5	1	40	5.5	6.5	3.0	7.0	1.0	6		20
	815NF100																		
	815NF200																		
	815NF300(PA)																		

Table 10.12.3. – 8-inch nanofiltration membranes by composition.

		VITAL STATISTICS									L/DAY PER CC	MODEL # FOR OTHER SIZES			
FACTURER	MODEL#	TEMP°	Cf	Cp	Cr	CM	dPI	dP	A	B		2.5" DIA.	4" DIA.	OTHER SIZES	NOTES:
ation Systems	CG8040F	298	17.1	2.7	18.7	17.9	74.5	1309.0	8.202	0.158	0.86		CG4040F	26.25x3.88", 25.0x3.88"	2
	DK8040F	298	17.1	8.6	18.1	17.6	44.3	619.5	19.293	0.908	0.92		DK4040F		9
:	NF40-8040	298	34.2	18.8	36.9	35.6	82.3	1411.4	8.170	1.054	0.80	NF40-2540	NF40-4040		
	NF70-8040	298	34.2	18.8	36.9	35.6	82.3	342.7	40.248	1.017	0.80	NF70-2540	NF70-4040	2.5X14", 2.5X21"	
ystems	8921	298	8.6	1.3	9.4	9.0	37.7	481.6	15.690	0.101	0.55		4921		
autics	8040-LSY-PVD1	298	10.1	2.0	12.8	11.4	46.2	964.3	15.482	0.263	1.26				4
ics	815PR	298	34.2	3.4	37.6	35.9	159.5	2791.0	5.791	0.139	1.23				
	811PR														
	815NF100														
	815NF200														
	815NF300(PA)														

Table 10.13. – Comparison of operating parameters for 8-in-diameter membranes.
Brackish Water Membranes

Composition	pH range		Maximum temp.°C	Maximum continuous oxidant conc. (p/m)	Manufacturer
	Operational	Cleaning			
Cellulosic	5-6.5	3-8	35	1.0	Desal
	4-6	4-7.5	40	1.0	Hydranautics
	4-6	3-7	40	1.0	Fluid Sys. HR
	3-7	3-8	40	1.0	Fluid Sys. SD
	3-8	3-8	35	1.0	Toyobo
Polyamide TFC	3-11	2-12	40	0.1	Osmonics
	3-10	4-10	45	0.1	Hydranautics
	4-11	2.5-11	45	0	Fluid Sys.
	3-10	2-11	45	0	Nitto Denko
Polyether Urea	5-10	4.5-11	45	0	Fluid Sys.
Polysulfone	2-11	1-12	45	0.1	Filmtec
Polyvinyl Alcohol	2-8	2-10	40	0.1	Hydranautics
Generic TFC	4-11	2-11.5	50	0.1	Desal.

Seawater Membranes

Composition	pH range		Maximum temp. °C	Maximum continuous oxidant conc. (p/m)	Manufacturer
	Operational	Cleaning			
Aramid HFF	4-9	2.5-12	40	0	DuPont
Cellulosic HFF	3-8	3-8	40	1.0	Toyobo
Polyamide TFC	4-11	2.5-11	45	0 - 0.1*	Hyd., F.S.
Polyether Urea	5-10	4.5-11	45	0	Fluid Sys.
Polysulfone	2-11	1-12	45	0.1	Filmtec

* Hydranautics recommends free chlorine less than 0.1 p/m. Fluid Systems membranes require 0 p/m free chlorine.

Figure 10.1. - Comparison of Productivity Measures: "*A*" vs. Average Flux

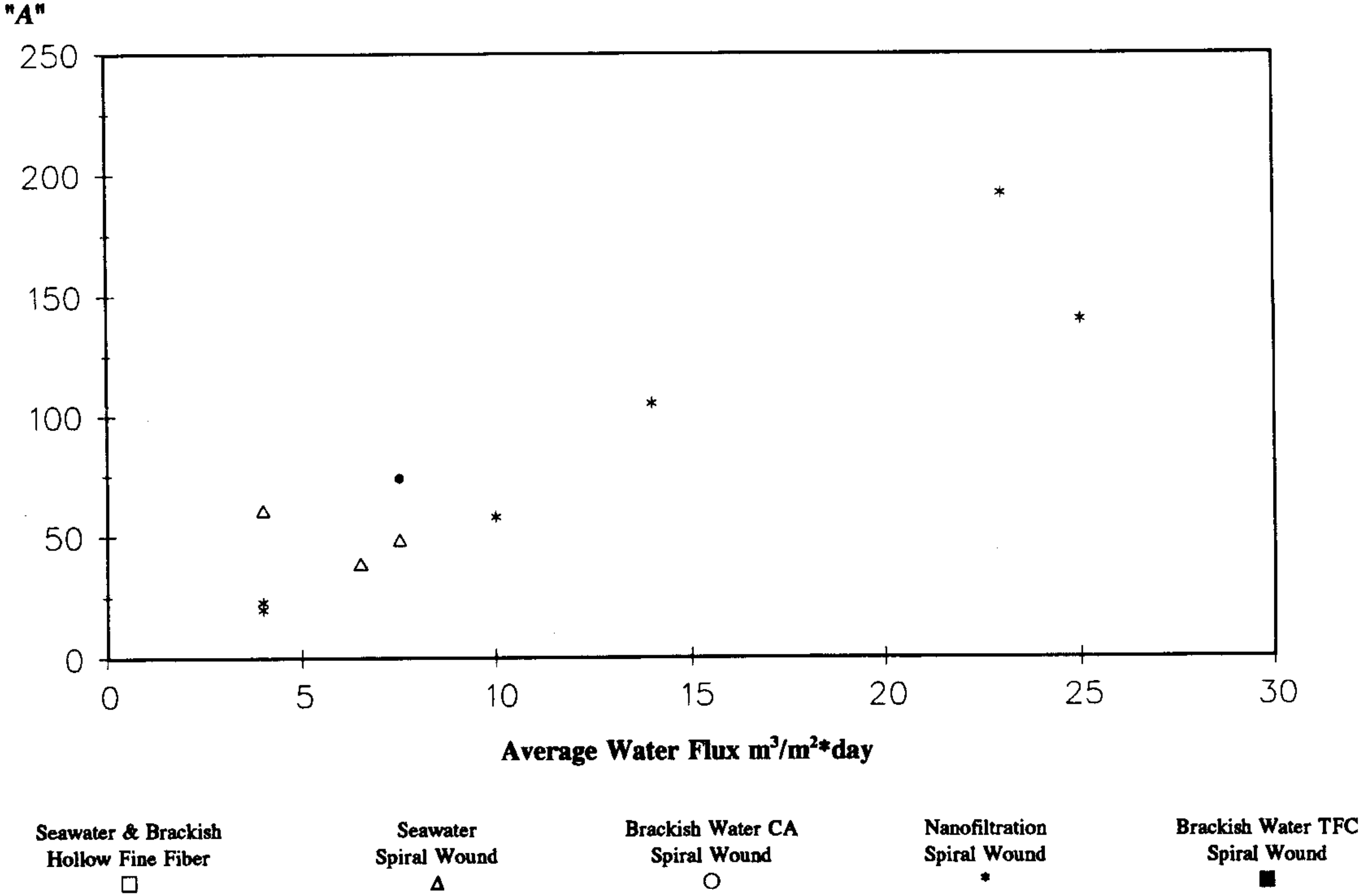


Figure 10.2. - Comparison of Productivity Measures: Module Productivity vs. Average Flux

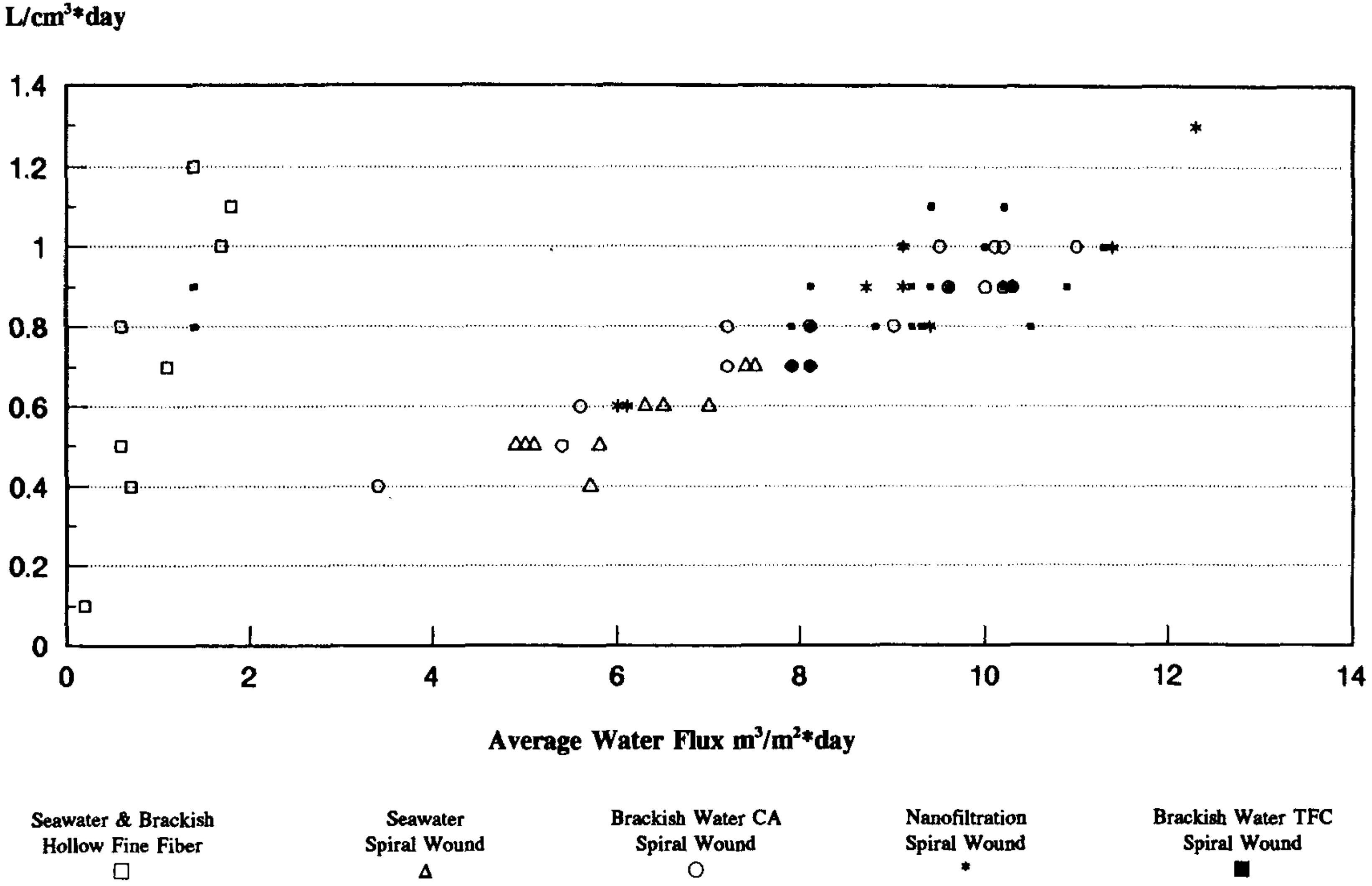
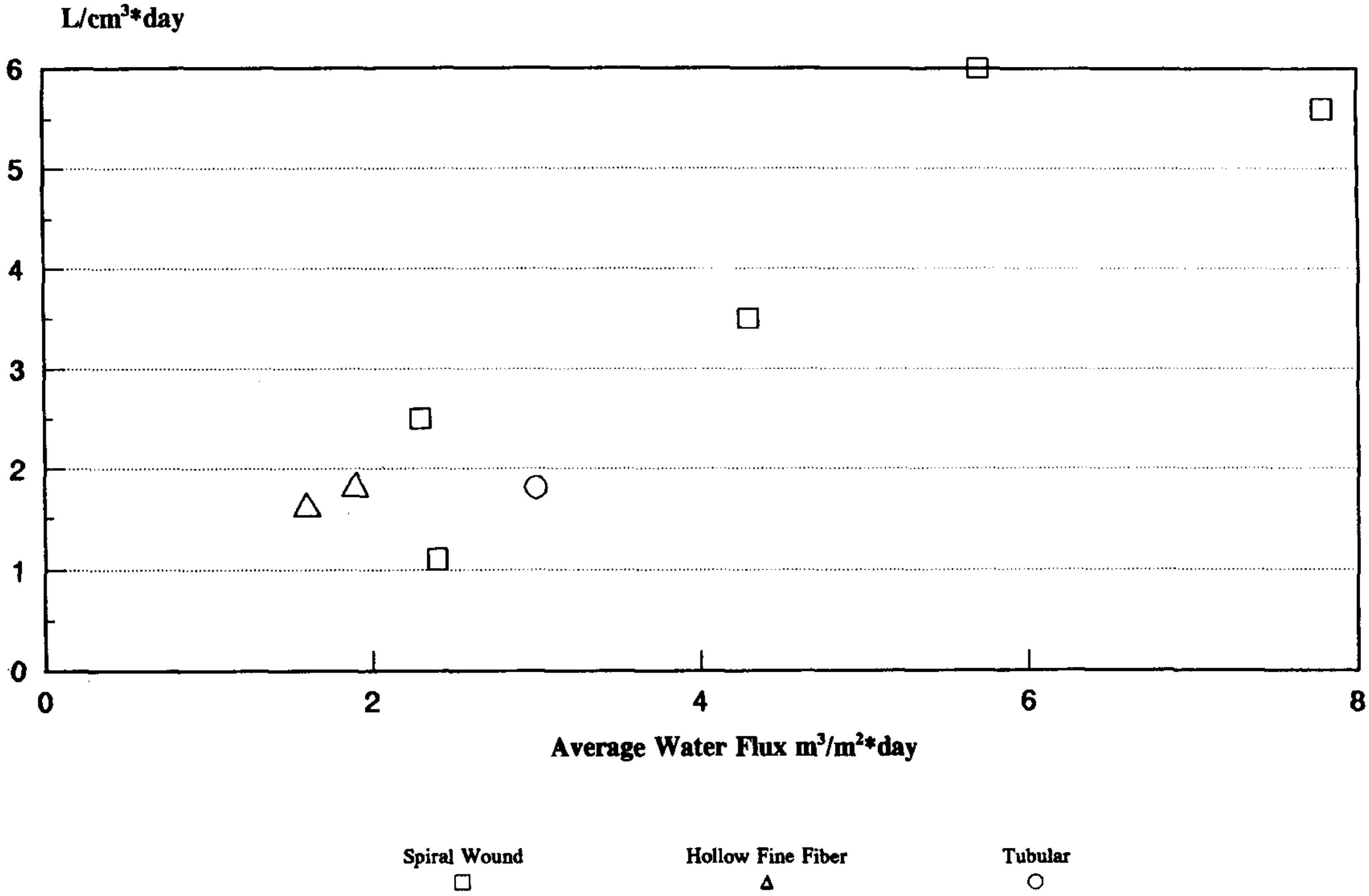


Figure 10.3. - Comparison of Productivity Measures: Ultrafiltration Membranes



Because ultrafiltration productivity depends so much on percent recovery, recirculation rate, and feedwater composition, comparing their flux rates or claimed productivity is difficult. Given specifications are usually based on "clean" water tests. No information is given on exact feed water composition, recirculation rate, pressure differential, or percent recovery. The best one can do under the circumstances is to compare water flux per square meter of membrane with water flux per cubic centimeter of element volume as is done on figure 10.3. Here, SW = Spiral Wound, HFF = Hollow Fine Fiber, HF = Hollow Fiber, and Tube = Tubular.

10.4 On Specifications

One of the most difficult tasks in completing this project was determining what should go into the comparison chart. The first catalog to come in belonged to Fluid Systems and everything in their specifications went into the table. As other catalogs and brochures came in, it became apparent that not all specifications were of equal worth in the eyes of the manufacturers. When the ultrafiltration brochures arrived, a new list was needed because so many informational holes could not be filled.

The market for membrane processes may still be too limited for standard test conditions to have evolved. Or membrane salespeople may want to design the whole treatment system. In the case of ultrafiltration, so many different applications exist that salespeople emphasize the diversity of applications for the process rather than describing the effectiveness for any particular application. Membrane processes are becoming common enough now that some effort should be made to create a list of what needs to be known about a membrane for a consumer to make an informed assessment.

A good starting point would be to establish a standard membrane testing procedure for each of the four membrane types. RO membrane producers have a collection of testing parameters that they report: temperature, pressure, and feed concentration. However, these parameters are not sufficient to assess the quality of the membrane.

The testing parameters that should be supplied are:

1. Composition of the feed water, including concentrations for substances above and below the rejection range of the membrane;
2. Composition of the concentrate, with concentrations of substances in No. 1;
3. Number of elements in the test module;
4. Feed flow rate;
5. Applied pressure;
6. Operating temperature;
7. Operating pH;
8. Recirculation rate for cross-flow microfiltration and ultrafiltration;

9. Pressure drop across the test system; and
10. Percent recovery.

The composition of the feed water is another item that needs standardization. An established standard exists for seawater, but not for brackish water, ground water, or hard water. If membrane producers would test their products with the same standard water, brackish water membranes with a standard brackish water, seawater membranes with a standard seawater, and nanofiltration membranes with a standard hard water etc., the separation process would be illuminated and the data produced would be of great value to the industry as a whole. Such standards would also help a great deal in making an informed assessment of membrane effectiveness.

Pressure from major desalting concerns, such as the NWSIA (National Water Supply Improvement Association) or the Interagency Consortium for Desalination and Membrane Separation Research, could stimulate the industry to accept standards. The U.S. Government is probably one of the biggest consumers of RO membranes. They should be able to specify a particular water for testing membranes purchased under contract. Then, the companies could use those results in their specifications for the rest of the market. This scenario brings up an obvious problem though - what constituents should be in a standard brackish water, standard ground water, and standard surface water?

11. PROCESS DESIGN

11.1 Introduction

The purpose of this section is to illustrate a simple modeling approach for estimating the overall performance of a pressure-driven, liquid filtration, membrane module. 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). 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 (reverse osmosis), NF (nanofiltration), UF (ultrafiltration) and MF (microfiltration) 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, these notes will begin with the specific flux models for the different types of filtration. Following that section, 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 to illustrate how one can proceed.

11.2 Solvent Flux Models

Figure 11.1 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.

Schematic of pressure-driven membrane process

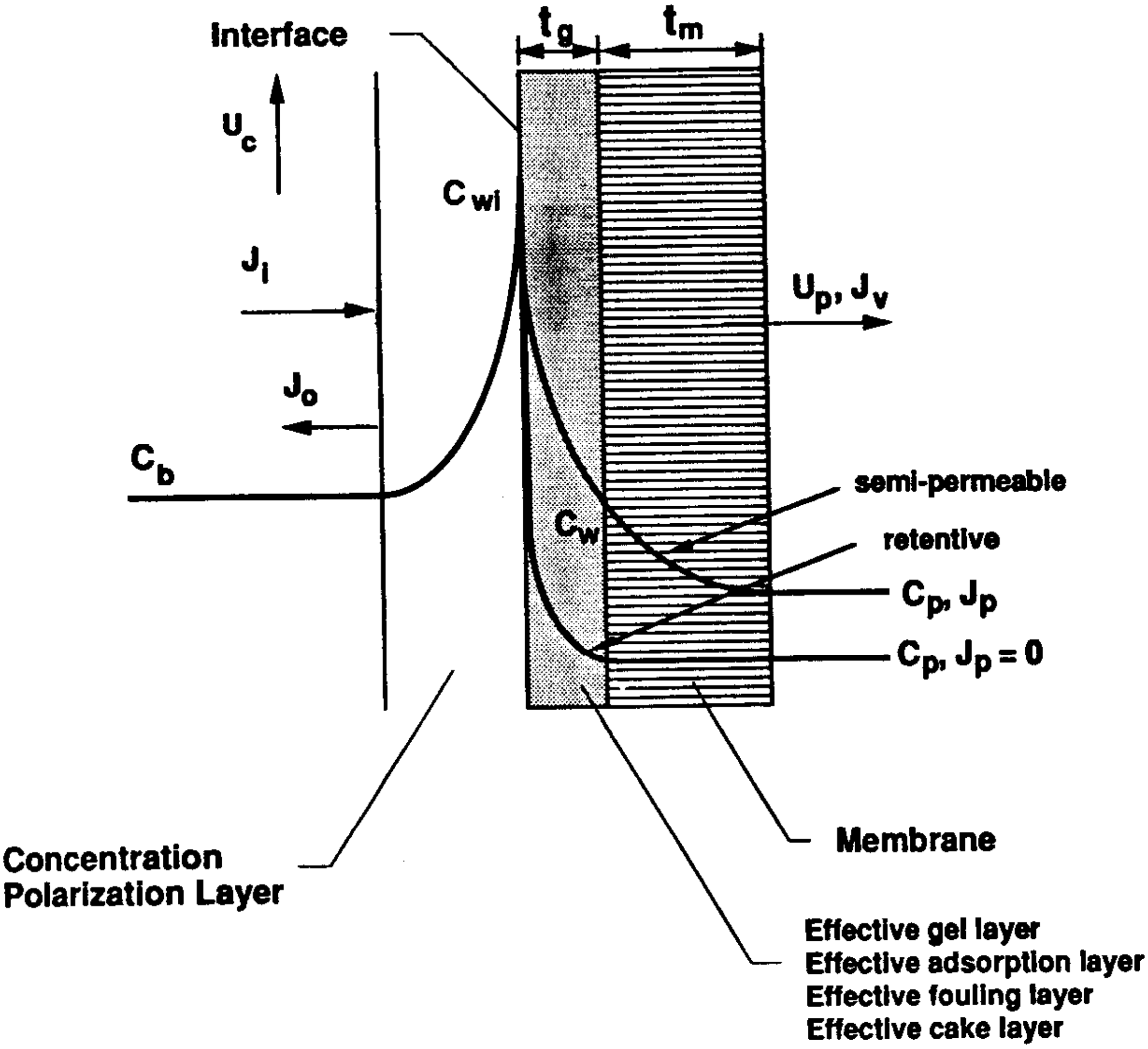


Figure 11.1 – Schematic of pressure-driven membrane processes.

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.

11.2.1 *Reverse Osmosis.* – Permeation of water through the membrane, J_v, in reverse osmosis is given by the following equation:

$$J_v = \frac{P_v}{t_m} (\Delta P - \Delta\pi)$$

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 concentrations refer to the solute (salt) concentration.

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 based on C_w and C_p (at a specific point)

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 ($\Delta P - \Delta\pi$).

11.2.2 *Nanofiltration, ultrafiltration, and microfiltration.* – The most general form for water permeation in nanofiltration, ultrafiltration, and microfiltration is:

$$J_v = \frac{\Delta P - \Delta\pi}{\left[\frac{t_m}{\hat{P}_v} + \frac{t_g}{\hat{P}_g}\right] \cdot \mu}$$

where:
$\hat{P}_v$ = specific water permeability of clean membrane
t_m = membrane thickness
$\hat{P}_g$ = specific permeability of the 'g' layer
t_g = thickness of:
- effective adsorption layer, and/or
- effective gel layer, and/or
- effective fouling layer,and/or
- effective cake layer

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:

$$J_v = \frac{\Delta P - \Delta\pi}{\mu(R_m - R_g)}$$

is used,

where:

R_m = clean membrane resistance

R_g = additional resistance from gels, cakes and adsorption (fouling)

For nanofiltration and ultrafiltration, both $\Delta\pi$ and R_g can be significant. For microfiltration, $\Delta\pi$ 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.

11.3 Estimation of Permeability Parameters

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

- For a group of pores containing laminar flow:

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

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):

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

where:

ε = void fraction

V_p = volume of particles

S_p = surface area of particles

11.4 Estimation of Osmotic Pressure

Osmotic pressure is a thermodynamic property and can therefore 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:

$$\pi = cRT$$

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:

$$\pi = ac \qquad \text{where } a \text{ is a constant}$$

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:

$$R^\circ = (C_w - C_p)/C_w$$

results in:

$$\Delta\pi = a(C_w - C_p) = aR^\circ C_w$$

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

$$\pi \, v_{solvent} = RT \ln \left[\frac{P^v_{solvent}}{P^v_{solution}} \right]$$

where:

$v_{solvent}$ = partial molar volume of the solvent
R = gas constant
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):

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

where:

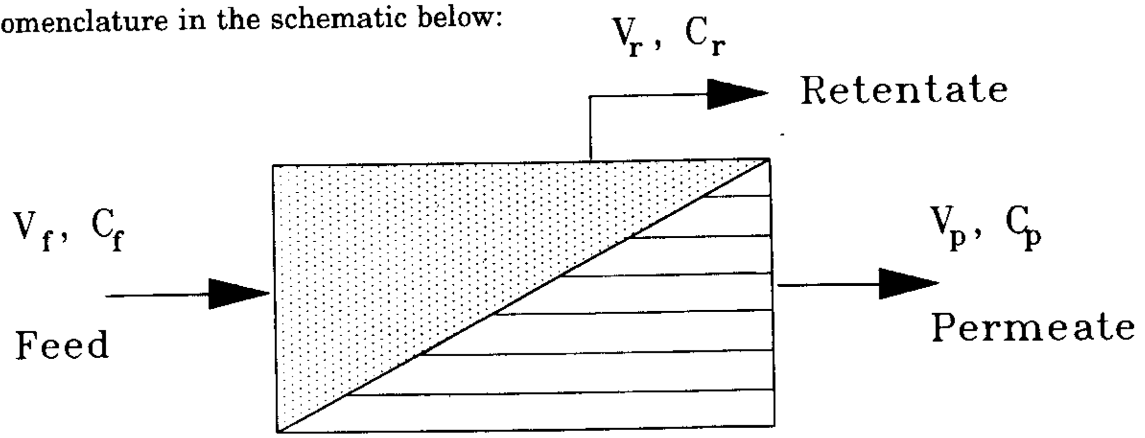
T_f^v = freezing point of pure solvent
T_f = freezing point of solution with concentration c
ΔH_f = latent heat of freezing of pure solvent
T = actual temperature for estimating π, °K
R_{press} = gas constant in pressure units
R_{energy} = gas constant in energy units

Useful values:

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

11.5 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. Using the nomenclature in the schematic below:



C = Molar Concentration of the "key" Component
V = Volumetric Flow Rate

(The "key" component is the solute whose rejection

by the membrane is under study.)

Figure 11.2. – Variables for module mass balance.

External mass balance:

$$V_f = V_r + V_p$$

define Θ = stage cut or recovery factor

External key component balance:

$$V_f c_f = V_r c_r + V_p c_p$$

The overall volumetric flux through the membrane is J_v, therefore $V_p = J_v{*}A$, where A = 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:

$$C_b = \frac{C_f - \Theta C_p}{1 - \Theta}$$

The definition of the intrinsic rejection of a membrane, $R°$, $R° = 1 - C_p/C_w$

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

$$C_p = \frac{C_b(1 - R°)\exp(J_v/k)}{R° + (1 - R°)\exp(J_v/k)}$$

And from the material balance $C_r = C_b$ is defined by:

$$c_r = c_b = \frac{c_f}{(1 - \Theta) + \dfrac{\Theta(1 - R°)\exp(J_v/k)}{R° + (1 - R°)\exp(J_v/k)}}$$

With k = the boundary layer mass transfer coefficient. The variable 'k' is obtained via correlations that usually assume that $J_v << U_c$, 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

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

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

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:

$$k = \bar{U}_c \frac{0.0791}{2} Re^{-1/4} Sc^{-2/3}$$

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

$$k = 0.023 \frac{D}{d} Re^{0.83} Sc^{1/3}$$

For turbulent flow in batch stirred vessels:

$$k = 0.0443 \frac{D}{d} \left(\frac{\rho}{\mu D}\right)^{1/3} \left(\frac{\omega \mu d^2}{\rho}\right)^{3/4}$$

d = vessel diameter
For laminar flow in round tubes:

$$k = 1.295 \left(\frac{2 \cdot \bar{U}_c D^2}{d\ L}\right)^{1/3}$$

L = distance along tube length

For laminar flow between parallel plates spaced at $2h$:

$$k = 1.177 \left(\frac{\bar{U}_c D^2}{hL} \right)^{1/3}$$

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:

$$C_r = \frac{C_f}{(1 - \Theta) + \dfrac{\Theta\,(1 - R^\circ)\exp(J_v/k)}{R^\circ + (1 - R^\circ)\exp(J_v/k)}}$$

[2] The wall concentration, C_w:

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

[3] The permeate concentration, C_p:

$$c_p = \frac{c_b\,(1 - R^\circ)\exp(J_v/k)}{R^\circ + (1 - R^\circ)\exp(J_v/k)}$$

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

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

[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 iterative solution of RO with concentration polarization, we define J_v as follows:

$$J_v = \left(\frac{P_v}{tm}\right) \cdot \left(\Delta P - aR^\circ \left[\frac{C_f \exp\,(J_v/k)}{(1 - \Theta)\,R^\circ + (1 - R^\circ)\exp\,(J_v/k)}\right]\right)$$

The objective function for an iterative solution is:

$$F(J_v) = 0 = \left(\frac{P_v}{tm}\right) \cdot \left(\Delta P - aR^\circ \left[\frac{C_f \exp\,(J_v/k)}{(1 - \Theta)\,R^\circ + (1 - R^\circ)\exp\,(J_v/k)}\right]\right)$$

If J_v/k <<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:

$$a'Jv^2 + b'Jv + c' = 0$$

where:

$$a' = (1 - R^\circ)/k$$

$$b' = 1 - \Theta R^\circ - \left[\frac{P_v}{tm}\,\frac{1 - R^\circ}{k}\,\Delta P\right] + \left[\frac{P_v}{tm}\,\frac{c_f a}{k}\,R^\circ\right]$$

$$c' = \frac{P_v}{tm}\,c_f a R^\circ - \left[(1 - \Theta R^\circ)\,\frac{P_v}{tm}\,\Delta P\right]$$

The general form for the roots of a quadratic is:

$$J_v = \frac{-b' \pm \sqrt{b'^2 - 4a'c'}}{2a'}$$

Once J_v is known, then C_p and C_r can be calculated.

11.6 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.

$$C_f = 0.0375 \text{ mol/L}$$

(1) Develop linear relationship for osmotic pressure, using literature data.

1 (kPa)	C(mol/L)
100	0.0375
300	0.125
700	0.3

$\pi = 2286 * C$ $\qquad$ $a = 2.286$ (kPa m^3/mol)

(2) Intrinsic rejection of membrane defined to be $R° = 0.987$

(3) Available transmembrane pressure, $\Delta P = 22905$ kPa (225 lb/in^2).

(4) Spiral wound module with:

L = 1.016 m (40 in)
h = 2.5×10^{-3} m
A = 39.02 m^2 (420 ft^2) per module

(5) Pure water permeability was determined to be 1500 L/m^2·day (1.736×10^{-5} m^3/m^2·sec) at 30,000 kPa. Therefore, estimate $P_v/t_m = (1.736 \times 10^{-5})/(3 \times 10^4) = 5.787 \times 10^{-10}$ m^3/m^2·s·kPa.

(6) Estimate mass transfer coefficient 'k'

- need estimate of U_c
- pump characteristic curve will give 189 L/min @ 1655 kPa (this rating is within specifications of the membrane module)
- need to approximate cross sectional area for flow in the spiral wound module: $A_x = (A * h)/2L = (39.02*0.0025)/(2*1.016) = 0.048$ m^2
- Average $U_c = \bar{U}_c = V_f/A_x$
- $V_f = 189$ L/min $= 3.155 \times 10^{-3}$ m^3/s
- U_c:

$$\frac{3.155 \times 10^{-3}}{4.8 \times 10^{-2}} = 0.0657 \text{ m/s } (6.6 \text{ cm/s})$$

- Approximation for Re:

$$Re \equiv \frac{h\bar{U}_c\rho}{\mu} = \frac{(2.5 \times 10^{-3})(6.57 \times 10^{-2})(10^{6})}{1}$$

Re ≅ 164, probably laminar flow

- use 'k' correlation for laminar flow in a flat channel
- $D_{NaCl} = 1.2 \times 10^{-9}$ m^2/s
- Calculate k:

$$k \cong 1.177 \left(\frac{6.57 \times 10^{-2} * (1.2 \times 10^{-9})^2}{2.5 \times 10^{-3} * (1.016/2)} \right)^{1/3}$$

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

- $k \cong 4.96 \times 10^{-6}$ m/s

(7) Check pure water value of J_v (1.736×10^{-5} m/s) versus k (4.96×10^{-6} m/s) versus U_c (6.57×10^{-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 the strictest sense.

(8) Set up to do trial and error iterative solution. See table 11.1 for an example of the spreadsheet output.

(9) This model shows:

- The value of J_v that drives the objective function to 0 is ~ 1.26295×10^{-5} m/s. This value is a relatively large fraction of the pure water flux.

- If we assume that $\exp(J_v/k) \cong 1 + J_v/k$, then $J_v \cong 1.305 \times 10^{-5}$ m/s.

- The recovery is ~ 15.62 percent.

- The apparent rejection of salt, defined as $R = (1 - C_p/C_f) * 100$ is 83.4 percent. This value seems low, which results from the conservative nature of this module model and the value of the intrinsic rejection, $R°$, that was used.

Table 11.1. – Sample spreadsheet for process design computations.

Intrinsic Rejection, R°	0.987	0.987	0.987	0.987	0.987	0.987
k (m/s)	4.96E-06	4.96E-06	4.96E-06	4.96E-06	4.96E-06	4.96E-06
Cut	0.156193783	0.156195019	0.156196256	0.156197493	0.15619873	0.156199966
Pv/tm [m3/(m2 s kPa)]	5.79E-10	5.79E-10	5.79E-10	5.79E-10	5.79E-10	5.79E-10
delP (kPa)	22905	22905	22905	22905	22905	22905
Cf (mol/l)	0.0375	0.0375	0.0375	0.0375	0.0375	0.0375
a [(kPa l)/mol]	2286	2286	2286	2286	2286	2286
Feed (m3/s)	3.16E-03	3.16E-03	3.16E-03	3.16E-03	3.16E-03	3.16E-03
Permeate (m3/s)	4.93E-04	4.93E-04	4.93E-04	4.93E-04	4.93E-04	4.93E-04
Area (m2)	39.02	39.02	39.02	39.02	39.02	39.02
Cw (mol/l)	0.479072641	0.479081281	0.479089921	0.479098562	0.479107202	0.479115843
Cp (mol/l)	0.006227944	0.006228057	0.006228169	0.006228281	0.006228394	0.006228506
Cr (mol/l)	0.043288652	0.043288686	0.043288719	0.043288753	0.043288786	0.04328882
EXP(Jsolv/k)	12.75865261	12.75890984	12.75916708	12.75942432	12.75968157	12.75993883
1+Jsolv/k	3.546209677	3.546229839	3.54625	3.546270161	3.546290323	3.546310484
	Iteration	Iteration	Iteration	Iteration	Iteration	Iteration
F(Jv) = 0	3.93E-10	2.82E-10	1.71E-10	5.95E-11	-5.18E-11	-1.63E-10
Guess Jv [m3/(m2 s)]	1.2629200E-05	1.2629300E-05	1.2629400E-05	1.2629500E-05	1.2629600E-05	1.2629700E-05
	1.26292					
increment	0.00001					
Polynomial Solution						
a'	2620.967742	2620.967742	2620.967742	2620.967742	2620.967742	2620.967742
b'	0.820967288	0.820966067	0.820964846	0.820963626	0.820962405	0.820961184
c'	-1.11627E-05	-1.11627E-05	-1.11627E-05	-1.11627E-05	-1.11626E-05	-1.11626E-05
Jsolv(+)	1.30531E-05	1.30531E-05	1.30531E-05	1.30531E-05	1.30531E-05	1.30531E-05
Jsolv(-)	-0.000326284	-0.000326283	-0.000326283	-0.000326282	-0.000326282	-0.000326281

12. CONCLUSIONS

- Membrane processes can replace many traditional water treatment methods. Fo instance, the use of nanofiltration instead of lime softening could reduce the vast amoun of sludge produced every day by municipal water treatment facilities.

- Pretreatment for RO and NF feed water is important in maintaining membran effectiveness. Traditional pretreatment methods are highly specific for the contaminan they are meant to remove though, causing many processes to be combined to treat wate from some sources. Traditional methods also tend to require a large amount of spac Membrane separations, on the other hand, are compact and specific only for size an charge. Combinations of micro-, ultra-, and/or nano- filtration could simplify wate pretreatment where multiple processes would normally be required ahead of RO and als where a space shortage exists.

- A standard membrane testing procedure and a standard test water composition is neede for each of the four membrane types.

- Membrane cleaning procedures are membrane and foulant specific, and symptoms whic determine the procedure needed are vaguely defined by manufacturers. Most companie do have technical hotlines or even offer regular cleaning as part of the maintenanc agreement. An added benefit to these services would consist of some type of monitorin device in the membrane element that would physically indicate the type of foulin present, or a non-destructive method of determining membrane condition.

- Concentrate disposal is probably the biggest obstacle to expansion of membrane processe at inland water treatment facilities. To facilitate this expansion, research is needed i the areas of concentrate reuse, methods of extracting more water from the concentrat the effects of concentrate on the flora and fauna of the receiving body of water, and o ground water systems. Especially helpful today would be direct discharge of seawate desalting concentrate streams back to the ocean.

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03427388 E.I. Monthly No: EIM9205-021974
Title: **"Desalination and industrial waste water treatment with the ROCHEM Disc Tube Module DT."**
Author: Peters, Thomas A.
Conference Title: Proceedings of the 12th International Symposium on Desalination and Water Re-use
Conference Location: Malta Conference Date: 1991 Apr 15-18
E.I. Conference No.: 16028
Source: *Desalination* v 83 n 1-3 Sep 1991. p 159-172
Publication Year: 1991
CODEN: DSLNAH ISSN: 0011-9164
Language: English

03413486 E.I. Monthly No: EIM9204-016517

Title: "**Desalination of brackish groundwater for a Prairie community using electrodialysis reversal.**"

Author: Harries, Robert C.; Elyanow, David; Heshka, Dale N.; Fischer, Kelly L.

Corporate Source: CH2M HILL Engineering Ltd, Waterloo, Ont, Can

Conference Title: Proceedings of the 12th International Symposium on Desalination and Water Re-use

Conference Location: Malta Conference Date: 1991 Apr 15-18

E.I. Conference No.: 15703

Source: *Desalination* v 84 n 1-3 Oct 1991. p 109-121

Publication Year: 1991

CODEN: DSLNAH ISSN: 0011-9164

Language: English

03427385 E.I. Monthly No: EIM9205-021971

Title: "**Proposal for the completely closed system in the Columbus Space Station.**"

Author: Madsen, R. F.; Thomassen, J. R.; Vial, D.; Binot, R. A.

Corporate Source: DANISCO A/S, Nakskov, Den

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Author: Marinas, B. J.

Corporate Source: Purdue Univ, West Lafayette, IN, USA

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Author: Mujeriego, R. (Ed.); Asano, T. (Ed.)

Corporate Source: Univ Politecnica de Cataluna, Barcelona, Spain

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Corporate Source: Life Systems, Inc, Cleveland, OH, USA

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Author: Sekoulov, I.; Figueroa, A.; Oles, J.

Corporate Source: Technical Univ Hamburg-Harburg, Hamburg, Ger

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E.I. Conference No.: 15243

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CODEN: WSTED4 ISSN: 0273-1223 ISBN: 0-08-040774-9 Language: English

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Author: Balaban, Miriam (Ed.)

Conference Title: Proceedings of the Twelfth International Symposium on Desalination and Water Re-use

Conference Location: Malta, Malta Conference Date: 1991 Apr 15-18

E.I. Conference No.: 15040

Source: *Institution of Chemical Engineers Symposium Series* v 2 n 125 1991. Publ by Inst of Chemical Engineers, Davis Building, Rugby, Engl. 402p

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CODEN: ICESDB ISSN: 0307-0492

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Author: Allott, Keith
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Author: Madsen, R. F.; Thomassen, J. R.; Jorgensen, L. B.; Bersillon, J. -L.; Vial, D.; Binot, R. A.
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