ABSTRACT

REVERSE OSMOSIS AS A PRETREATMENT FOR ION EXCHANGE AT PSE&G'S HUDSON GENERATING STATION

by
Steven Leon

Public Service Electric and Gas Company's Hudson Generating Station has historically had problems providing sufficient high quality water for its two once through, supercritical design boilers. The station requires over 60 million gallons annually to compensate for system losses.

The stations' ion exchange demineralizers proved to be costly to operate and overall inefficient in performance. Drawbacks include: short service cycles, periodic contamination of storage tanks due to premature breakthrough, prohibitively high chemical regenerant costs, excessive labor requirements, frequent resin replacement, and overall unreliable plant operations.

A reverse osmosis unit was installed as a pretreatment to the demineralizers to offset these shortcomings. This did not eliminate the demineralizers, but vastly reduced the influent loading and extended the service cycle more than twenty fold.

Significant cost savings have been realized, water quality was greatly improved, and plant reliability is secured into the next century.

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by
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REVERSE OSMOSIS AS A PRETREATMENT FOR ION EXCHANGE EQUIPMENT AT PSE&G'S HUDSON GENERATING STATION

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This thesis is dedicated to Ann and Dominick Leon.

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CHAPTER 1

INTRODUCTION

1.1 Plant Design

The Hudson Generating Station is the largest fossil fuel-fired electric generating station in PSE&G's system. The station is unique in that both of the steam units are once-through, supercritical design, meaning there are no drums in the boiler to collect and distribute steam, as are found in most generating units. The once-through design allows all the water in the boiler to steam simultaneously, prior to entering the turbine; the supercritical design involves the system pressure and temperature, which is in excess of 3500 psig and 1,050 degrees Fahrenheit. Supercritical, temperatures and pressures greater than 704 degrees Fahrenheit and of 3204 psig, denotes the point above which water can no longer exist in the liquid form regardless of applied pressure. The high amount of energy held in this steam makes these units much more efficient than the lower pressure units in the system. At these high temperatures and pressures the thermal expansion in the turbines, which drive the electric generators, is greatly enhanced, thereby, producing more work with less input. Unfortunately, the advantages gained from a supercritical design is a disadvantage from a water quality aspect, since the higher temperatures reduce mineral solubility making them more susceptible to precipitating on boiler tubes and turbine blades.

The No.1 unit, built in 1964 and rated for 400 megawatts, is either gas or oil fired depending on the cost, and generates 2.8 million pounds per hour (MPH) or 5600 gallons per minute (gpm) of steam. Unit No.2, built in 1968, is a coal-fired unit rated at 600 megawatts. This unit generates 4.0 MPH or 8000 gpm of steam. Figure 1 shows the flow pattern for the units. Each unit has an identical flow pattern. The condensate from the surface condensers is pumped by the primary condensate pumps at approximately 200 psig to the condensate polishers. These polishers contain mixed cation/anion resins that provide a water purity conductivity of less than 0.1 micromhos, or generally termed just micromhos. From here, the condensate is pumped by the secondary condensate pumps at approximately 800 psig through the low pressure heaters. The temperature at the outlet of these heaters is approximately 250 degrees Fahrenheit.

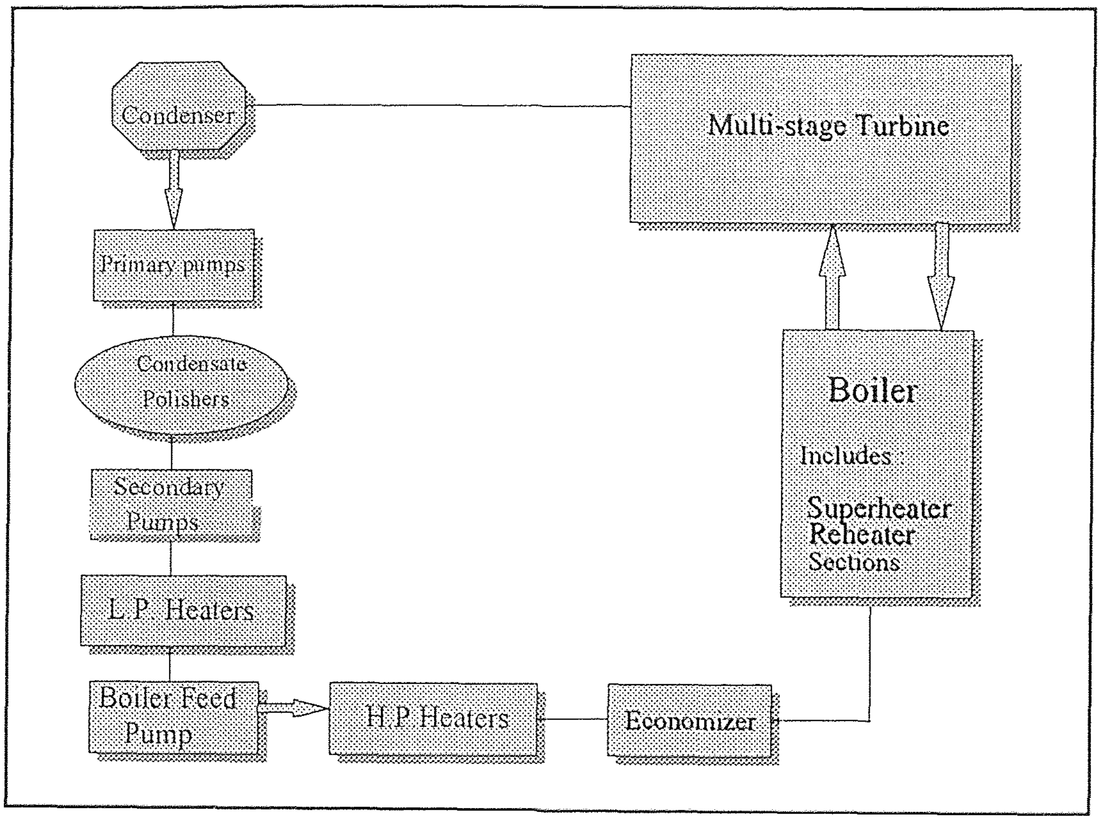

Figure 1 Hudson Generating Station condensate and feed water flow path

The effluent of the heaters is pressurized in the main boiler feed pumps to approximately 4700 psig. The condensate is pumped through high pressure steam heaters into the economizer where the temperature increases to 420 degrees F. The flow continues into the boiler and exits the boiler as steam at approximately 750 degrees F. Steam is sent to the superheater section where it is dried and heated to approximately 1050 degrees F. At this point, the steam enters the Very High Pressure (VHP) section of the multi-stage turbine at a pressure of 3500 psig. The turbine spins at a constant speed of 3600 rpm. For efficiency, the steam returns to the boiler's first reheater, which is another superheater section of the boiler that reheats the steam. The steam is sent back to the High Pressure (HP) section of the turbine at a pressure of 1500 psig and 1050 degrees F. After exiting the HP section, the steam enters the second reheater section, where once again it's heated to 1050 degrees F. The steam is then sent to the Intermediate Pressure (IP) section of the turbine, entering the turbine at 600 psig. From the IP section, the steam is passed through the Low Pressure (LP) sections of the turbine and then enters the surface condensers. These condensers serve the dual purpose of condensing the steam back to condensate, and creating a vacuum, which deters excessive back pressure on the turbine blades, making the turbine more efficient. The cycle then continues.

The original design specification indicated a 0.5% makeup to compensate for steam losses, which invariably occur in a generating unit. At 0.5%, Unit No.1 would require approximately 30 gpm and Unit No.2 approximately 40 gpm make-up, providing a total of 70 gpm or less of make-up quality water. Unfortunately, these design parameters were optimistic, the unit typically runs with a loss of 125 gpm; many times approaching

200-300 gpm make-up. Since these units are comparatively efficient and critical to the PSE&G system, it is not economical, during peak electric periods, to remove them from service, and therefore, they are kept in service even though it strains the capacity of the demineralization make-up plant.

1.2 Water Quality Criteria

The water quality criteria, based on design pressure and turbine manufacturer warranty guidelines, are 10 ppb dissolved or reactive silica, and 0.3 micromhos total conductivity. For steam generators, conductivity is the preferred method for detection of primarily sodium and chloride ions, but also measures the total inorganic composition of the water. These two parameters are used in lieu of monitoring all the constituents of the water such as calcium, magnesium, alkalinity, etc. It is known that if the ion exchange units can remove silica and total water conductivity to design residuals, then due to the selectivity of the ion exchange resins, (which have a low preference for silica and sodium), the other constituents must be lower than these concentrations. Silica is the most detrimental of all the mineral constituents and is responsible for the majority of steam turbine blade deposits. The formation of adherent deposits distorts the original shape of turbine nozzles and blades. The deposits, frequently rough and uneven at the surface, increase the resistance to the flow of steam. Distortion of the steam passages alters the steam velocity, pressure drops, and reduces the efficiency with which the energy is recovered from the steam. Where the conditions are severe, the deposits may develop unevenly and cause vibrational problems. As deposits accumulate on turbine blades, efficiency rapidly diminishes. Continued operation in excess of the manufacturer's guidelines is not advisable. Further, the amount of steam needed per kilowatt hour of power generated will increase to uneconomical levels. Silica or conductivity in excess of design limits will be a deciding factor for the station's chief engineer as to whether he should trip (remove from service) the unit. Removing the unit from service, even during non-peaking times, would result in a loss of millions of dollars to PSE&G since a less-efficient gas or oil fired unit would be put in service to supply the lost megawatt capacity from the Hudson station; a situation that PSE&G, an associate member of PJM (Pennsylvania, New Jersey, Maryland), a power sharing network, finds uneconomical to the rate-payer.

The condensate polishers, which are in-line to the flow pattern should not be mistaken for the demineralization make-up plant. They are both ion exchange units, but are separate entities. The condensate polishers remove ionic impurities and suspended solids from the circulating water by utilizing deep bed ion exchange vessels. Each of Hudson's steam units has four of these vessels; a minimum of three vessels must remain in service to accommodate maximum unit load. The fourth polisher vessel is available for service when one of the three active vessels is removed from service for chemical regeneration. The water quality of the polisher effluent approaches the ultrapure level of 0.055 micromhos at 25 degrees C. This level is consistently realized with polisher effluent conductivity levels usually in the 0.06-0.07 micromhos range, and silica levels between 3-7 ppb. The demineralizer plant supplies make-up water to compensate for system losses and automatically feeds this water to the surface condensers. As such, the water quality from the demineralizer plant does not have to attain the same high purity water as the effluent from the condensate polishers since this water enters upstream of the polishers; consequently, any impurities are removed by the polishers. However, the water being supplied to the condensers does have certain limitations. The condensate polishers cannot kinetically withstand a high ionic loading due to the extremely high circulation rate through the vessels. Therefore, limitations of the demineralizer plant effluent are typically 15 ppb silica and 1.0 micromhos conductivity.

1.3 Demineralization Make-up Plant

The demineralization plant, constructed in 1964, employs ion exchange technology to produce the high purity water required for the steam units. The plant produces approximately 60-65 million gallons annually of high quality make-up water for water loss replacement, auxiliary equipment cooling, and for ion exchange regeneration water.

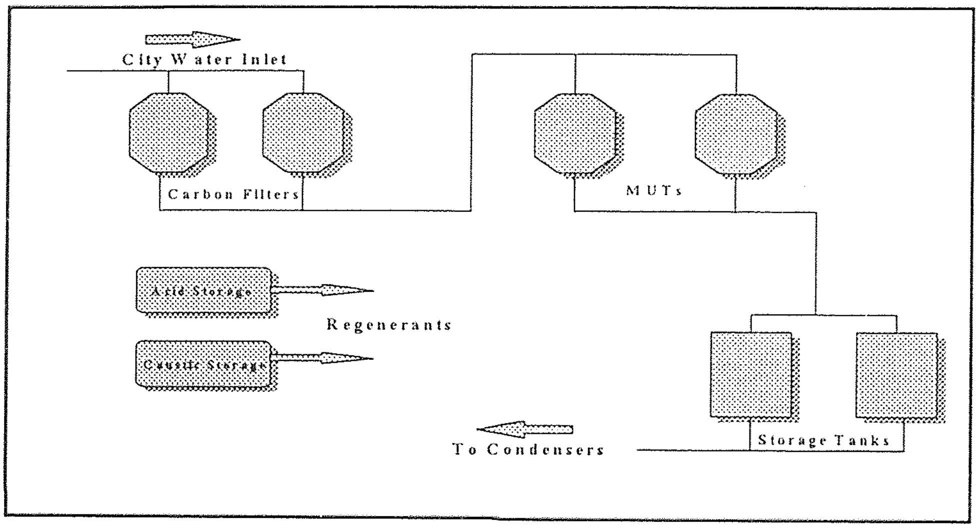

Figure 2 The Demineralization Plant

Figure 2 is a schematic of the demineralization plant. Initially, the water receives treatment from two activated carbon filters whose primary purpose is to remove chlorine that will oxidize and degrade ion exchange resins. The additional benefits of activated carbon, is that it removes, to a lesser extent, organic matter and suspended solids from the influent. The activated carbon undergoes replacement when total chlorine residuals exceed 0.1 mg/l. The effluent of the carbon filters enters either of two mixed bed ion exchange vessels, more commonly referred to as MUTs (Make Up Tanks). Each vessel contains 225 cubic feet of anion resins and 140 cubic feet of cation resins. Cation resins are maintained in the hydrogen form, while anion resins are maintained in the hydroxide form. One MUT is in service while the other is on standby. Typical water quality of the effluent is less than 15 ppb silica and 1.0 micromhos conductivity. This is a very efficient removal rate considering that the influent to the MUTs is 250 micromhos conductivity and approximately 7000 ppb dissolved silica. The MUT supplies water to two 200,000 gallon storage tanks which, in turn, supply water to the unit condensers. When a MUT exhausts, that is, silica in excess of 15 ppb, or conductivity in excess of 1.0 micromhos, it is removed from service for regeneration. Regeneration takes approximately 6 hours. The mixed bed resin is backwashed to separate the resin, based on their density difference. The lighter anion resins transport to the top of the resin bed; and the heavier, more dense cation resins settle to the bottom. The anion resin is regenerated with a 6% solution of sodium hydroxide and the cation resin is regenerated with 4% sulfuric acid. Each regeneration requires 300 gallons of sodium hydroxide (50%), 150 gallons of sulfuric acid (92-96%), and approximately 60,000 gallons of demineralized water.

The major drawback of the plant is that the present method of producing water is also one of the most inefficient; treating city water with mixed bed ion exchange resins. Ideally, it would have been advantageous to have separate anion and cation vessels subsequent to the MUTs. However, this was not anticipated at the Hudson Station since originally, the design thought was that once-through units will have a very low make-up demand, and therefore, would not require a large make-up plant. This has not proven to be the case and the plant has operated for years in a distressed manner. At least one MUT is regenerated daily when the units are in service. This has resulted in an excessive operating and maintenance budget of more than $500,000 for ion exchange resin replacement, chemical regenerants, labor requirements, and miscellaneous equipment purchases. A convenient way to evaluate a demineralizer's efficiency is to compare the amount of solids removed to the amount of acid and caustic used per regeneration. If the amount of solids removed are 30-40% of the amount of regenerant used, the exchanger is performing efficiently. This is shown in equation 1.

$$\frac{(\text{mg/l influent anion/cation})\ (\text{million gallons/cycle})\ (8.34)\ (100)}{(\text{gallons regenerant})\ (\text{concentration})\ (\text{pounds/gallon})} = \text{Efficiency} \tag{1}$$

This formula provided an efficiency rating of 17.4% cations and 19.6% anions; indicating that the ion exchangers were well below performance expectations. These efficiency ratings were based on: 150 ppm Total Dissolved Solids, 0.3 million gallons of throughput, 150 gallons of 94% sulfuric acid, and 300 gallons of 50% sodium hydroxide per regeneration.

In addition to the plant being inefficient, it has been unable to consistently supply the needs of the units. This has resulted in the costly rental of mobile demineralizer systems, which significantly increased the budget. In addition to the excessive cost to operate the plant, the demineralizers show high amounts of silica leakage in excess of 15 ppb during peak flow periods which has placed a burden on the condensate polishers and shortened their service cycle. Another shortcoming is the large amount of regenerant chemicals required by the plant. Weekly deliveries of acid and caustic, besides their expense, increased the potential of an accident or spill during unloading. It is preferable to reduce the amount of chemical deliveries. Lastly, PSE&G is downsizing its workforce and the inefficiency of the demineralizer plant is far too labor-intensive, especially in overtime situations. With these factors in mind, a more efficient and less costly method for producing make-up water was of critical importance to the operation of Hudson Generating Station. As the technical supervisor, responsible for the demineralization and chemistry plant, I was assigned to find an alternate water treatment technology to correct these shortcomings, and whose longevity will likely suffice for the remaining life of the station; the year 2010.

1.4 Alternate Technology Investigations

Our criteria for selecting alternate technologies was to find one that guaranteed acceptable water quality, was labor-efficient, and did not require chemical regeneration. Two technologies exist that economically meet these criteria; reverse osmosis and electrodialysis. Neither of these technologies require chemical regeneration and both have a certain usefulness in their application. However, which system best suited Hudson's needs?

Electrodialysis works on the principle of chemical electroneutrality. The unit consists of an anode and a cathode that attracts its oppositely charged ion, such that calcium (a cation) attracts to the anode while chloride (an anion) would be attracted to the cathode. By forcing feed water through sealed passages, it is possible to create a concentrated waste stream and a relatively pure product stream. This type of system was considered because it is durable and not susceptible to fouling. Its major drawback for our utilization was its inability to remove silica. Since silica is only weakly ionized, it can not be removed, because only strongly ionized materials are removed by this process. The manufacturer claimed that electrodialysis would remove other anions in the water leaving extra capacity in the ion exchange vessels for the removal of the silica, thereby, making it a benefit for our use. We visited other utilities using electrodialysis, and found their claim to be generally true. However, it did saturate their ion exchange resins with excessive amounts of silica. Since silica is difficult to remove during regeneration, subsequent service cycle capacity diminished. We did not feel this was a suitable alternative.

Another option was reverse osmosis. This technology has been employed for years for desalinization, and for water production in the pharmaceutical and electronic industries. Within the last ten years, reverse osmosis has made significant inroads in the electric generation field. Reverse osmosis met our requirements in that it does not require chemical regeneration; it's a pressure-driven membrane technology, has a very high removal efficiency for silica, organics, and other mineral constituents, and is not labor intensive. Since the technology is new to the power industry, competition is strong and vendors are eager to supply their units at a reasonable cost. We visited a number of sites utilizing reverse osmosis technology to see their operation, collect data, and to discuss operation, maintenance, and drawbacks of the unit. Generally, reverse osmosis received good marks from other utilities.

Based on our preliminary calculations and determinations, a reverse osmosis system would remove at a minimum 95% of the raw water solids. This would increase the MUT service cycle from the present 300,000 gallons to approximately 6 million gallons. In addition, by installing the proper instrumentation and control equipment with the new system, it would be possible to automate this system, thereby, reducing the labor requirements to operate that plant.

After ascertaining that reverse osmosis is the technology of choice, two undecided factors remained. A major factor was the decision to buy or lease the technology. Both options are readily available. By purchasing the system, our payback would have been quicker; however, leasing the equipment left the burden of membrane maintenance and replacement with the vendor. A major cost for maintaining a reverse osmosis system is membrane replacement, which would cost approximately $100,000.00. Under optimum conditions, membrane life is approximately 4-5 years, and even with the best operation and maintenance, this option would be potentially too costly. Our decision was to lease the system with the station having the ultimate control of membrane selection, and system criteria, such as pretreatment, automation, and water quality objectives. We found the vendors to be agreeable and the station developed and implemented specifications for the system bid package.

Our second consideration was whether to integrate the system with our existing plant, or to decommission our demineralization plant and rely solely on the reverse osmosis system. A single reverse osmosis unit could not supply the water quality required for our make-up plant. Two units are required; one as a roughing system, the other as a polisher. This option was disqualified because of the cost; using two reverse osmosis systems in line is expensive. In addition, the demineralization plant regeneration system is required to regenerate the condensate polishers, therefore the demineralization plant could not be fully decommissioned. The final decision was to integrate the reverse osmosis system with the existing demineralization plant; utilize the reverse osmosis system as a roughing demineralizer, and use the existing plant as a polishing unit to achieve acceptable water quality criteria. Figure 3, was a useful guide to help make our decision. It shows that at our influent conductivity, 250 micromhos, the most economical approach was a reverse osmosis (RO) system followed by ion exchange (IX). The difference between RO/RO and RO/IX is approximately $0.80/1000 gallons, which is a significant savings considering that the plant produces 60-65 million gallons annually. If it had been possible to decommission the demineralizer plant, serious consideration would have been given to an RO/RO installation.

Bid specifications were developed at the station and provided to five reverse osmosis vendors in late November 1993. The specifications required the vendors to provide a cost per thousand gallons of product water. This cost was to include the reverse osmosis units, pretreatment equipment, membrane replacement and cleaning , filter media replacement, and technical support. A five year lease for a 300 gpm reverse osmosis system was awarded to the Polymetrics Corporation, of South Windsor, CT., in December 1993.

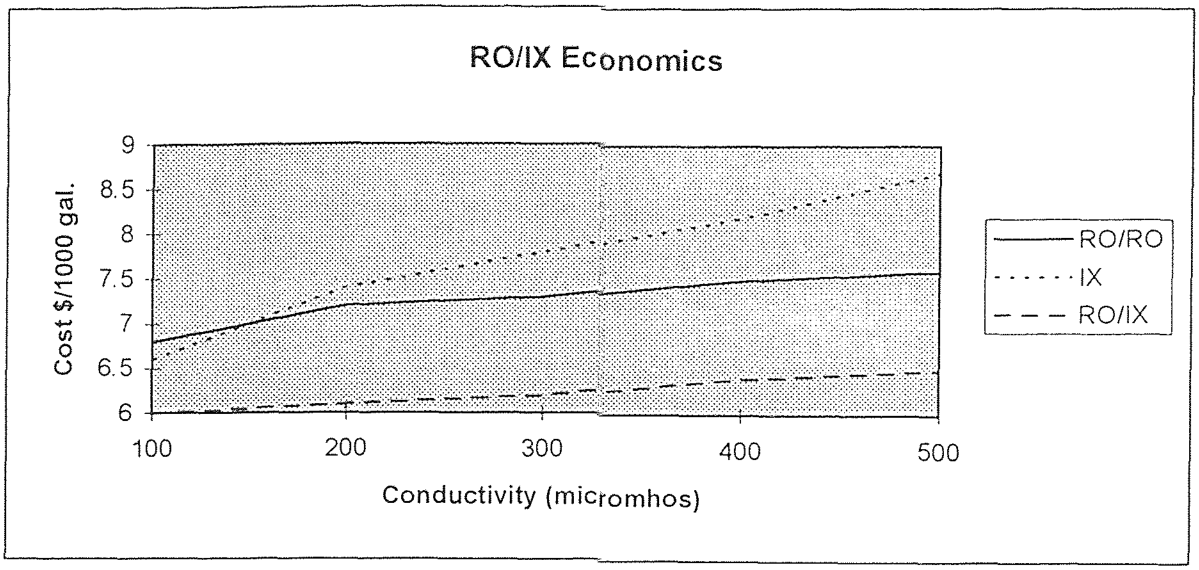

Figure 3 Reverse osmosis and ion exchange economics

CHAPTER 2

REVERSE OSMOSIS SYSTEM DESIGN

2.1 Basic Concepts

Reverse osmosis is a water filtration technology which utilizes a semi-permeable membrane. Semi-permeable membranes allow water to pass through (permeate) readily, but are fairly impermeable to other constituents of the water. The origins of reverse osmosis are based on osmosis, a fundamental action of nature. When a semi-permeable membrane, such as a living cell wall, separates two solutions with differing concentrations of dissolved solids, pure water will flow from the solution containing the lower concentration of solute through the membrane into the solution containing the higher concentration of solute. This movement of water through the cell wall (semi-permeable membrane) can be explained by the fact that the solution containing less solute is at a higher energy state than the more concentrated solution. In order to attain an equilibrium of energy, the movement of water results. By applying pressure to the more concentrated solution, the normal osmotic flow is reversed and pure water is forced through the semi-permeable membrane into the less concentrated solution. In the reverse osmosis process, applied pressure is provided by a pump and basically adds energy to the more concentrated or low energy side to account for the movement of water. Osmotic pressure is the difference between the potential energy of any solution and that of pure water. It is a function of the specific solute and its concentration. In practical terms, it is the minimum pumping energy required to initially produce pure water from a solution of solute at a specific concentration. Reverse osmosis specifically involves the separation of dissolved ionic materials from water. Figure 4 illustrates the osmosis and reverse osmosis principles.

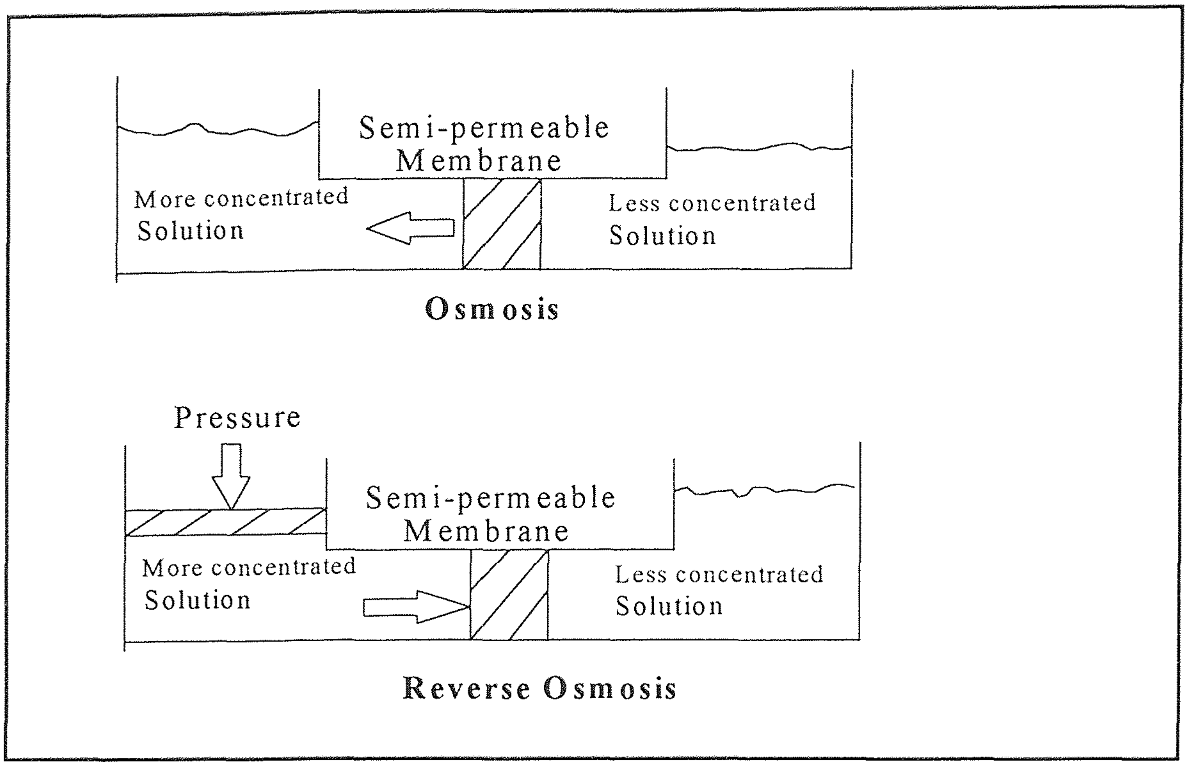

Figure 4 Osmosis and reverse osmosis principles

The higher the ionic charge (or valence) of an ion, the greater its tendency to be repelled from the surface of the membrane. This means, that monovalent salts such as sodium and chloride will tend to pass more readily through the membrane into the pure water side at a higher rate than multivalent salts such as calcium and sulfate. The typical pore size of a reverse osmosis membrane is 5 angstrom units (0.0004 micrometers). In reverse osmosis the rate of production of pure water, or permeate, is a function of the membrane material, the applied pressure, the membrane area, the differential osmotic pressures of the two solutions, and the temperature.

The exact mechanism of reverse osmosis filtration is controversial. There are several theories. Some theories rely on the presence of sub-visible pores in the membrane to describe the phenomenon, Porosity Theory. Some theories disclaim the presence of pores and rely on differences in the rate of diffusion of substances through the membrane, Diffusion Theory. Whatever the mechanism, pores are not seen by scanning electron microscopy, however a concentration gradient is developed across the membrane.

2.2 Membrane Selection

There are two semi-permeable membrane materials which account for the majority of the membranes presently in service. These are cellulose acetate and polyamide. Cellulose acetate was the first commercially available membrane, however, polyamide membranes, namely the thin-film composite membranes, have recently accounted for the majority of membranes in use today. We selected thin-film composite polyamide membranes because of the following advantages over cellulose acetate:

- Thin film composite membranes typically operate at a pressure of 300 psig, as compared to 550 psig for cellulose acetate. This translates into lower electrical cost to operate the feed pump.
- Thin film composite membranes show a better salt rejection. These membranes can reject up to 98% of influent salts as compared to 85-90% for cellulose acetate.

- The water flux per unit area is greater with thin-film composite membranes than with cellulose acetate which requires less membrane surface area, and thereby, allows a smaller size unit for housing requirements.
- The pH range for thin-film composite membranes is between 1-12, which is a much wider range than cellulose acetate. This is beneficial for both operation and cleaning of the membranes.
- The thin-film composite membranes are less susceptible to fouling from suspended solids than the cellulose acetate membranes.
- The thin-film composite membranes are not biodegradable, which is extremely important in the event that the membranes are out of service for any significant amount of time.

Two configurations of semi-permeable membranes are common, flat sheet and hollow fiber. Flat sheet membrane is manufactured by applying a semi-permeable material to a woven or nonwoven cloth. It is manufactured as a continuous sheet and rolled up like a large paper towel roll. Hollow fiber membranes are extruded like fish line with a hole in the center to create a tiny hollow fiber. The flat sheet membrane is used in "spiral wound" reverse osmosis elements. The thin-film composite membranes in the spiral wound configuration, offer several advantages over the hollow fiber-type configuration. Spirals offer easier maintenance, greater design freedom, are less prone to fouling, more forgiving of pretreatment upsets, and deliver the best combination of productivity and long-term performance. The station's choice was the spiral-wound configuration.

Spiral-wound modules consist of two layers of membrane, sandwiched around a permeate carrier, with one end of this sandwich bonded to a plastic permeate tube. The tube is perforated, and the membrane sandwich is attached so that the permeate will enter the tube through the perforations and exit through one end or the other. The polymer "sandwich" construction is rolled around the tube in a spiral fashion, and the feed water enters from one end parallel to the membrane surface. A plastic netting spacer provides a finite separation between the sandwich layers. Under pressure, permeate passes through the membrane layer, and is absorbed by the permeate carrier. It passes in a spiral fashion down to the permeate tube, while the concentrate stream simply passes out the other end of the element. Figure 5 illustrates the spiral wound configuration and flow patterns.

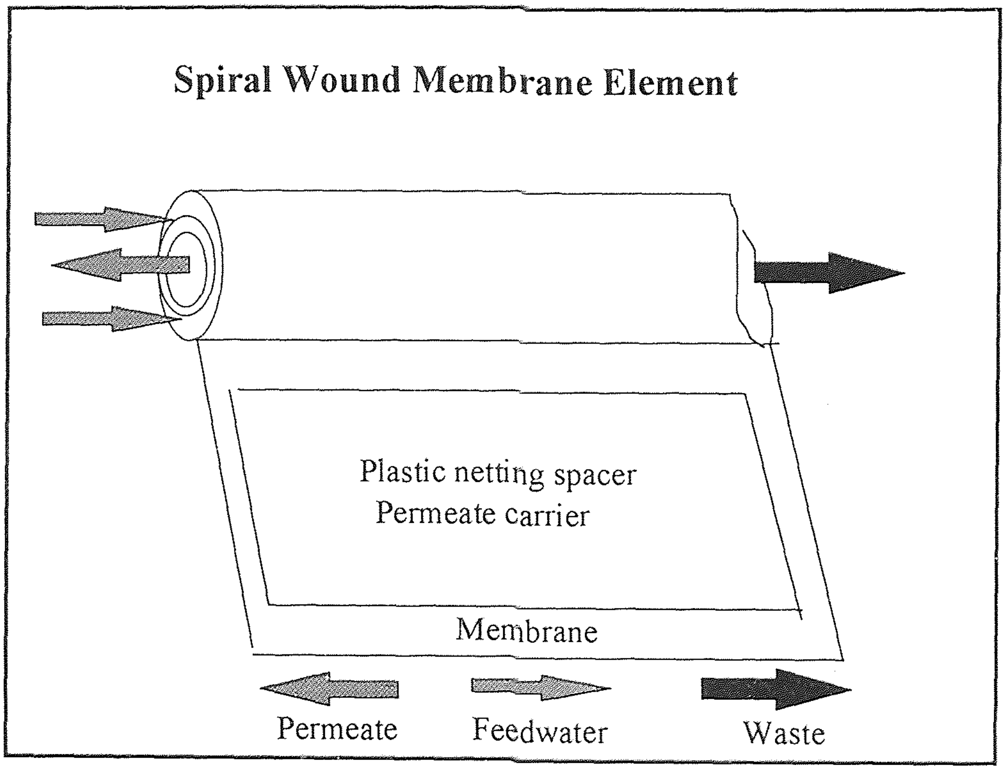

Figure 5 Spiral wound membrane configuration

It is important to note that the concentrate is that portion of the feed stream that is left after the permeate passes through the membrane. The concentrate stream is rich in dissolved materials simply because the membrane rejects them, while allowing pure water to pass through. The concentrate or waste stream typically contains four times the dissolved solids as the feed water stream. The spiral wound configuration offers a good ratio of surface area to volume, and can be operated under turbulent conditions. These turbulent conditions help deter suspended solids from settling and adhering to the membrane surface.

The thin-film composite membrane used at the Hudson station consists of three layers. The underlying support is provided by a nonwoven polyester web which is about 120 microns thick. The microporous polysulfone interlayer is about 40 microns thick and is used for its resistance to compaction. Finally, the fully aromatic, cross-linked polyamide barrier layer is about 2000 angstroms thick and provides the barrier for salt rejection.

2.3 Design Calculations

The design of reverse osmosis membranes is contingent on the flux of water through the membrane, the thickness of the membrane barrier, the viscosity of the feed water as it changes with temperature, and the differential pressure across the membrane. The homogeneous solution diffusion model is the most widely accepted and best fits the reverse osmosis phenomena. This model describes mass transport in the permeate stream of the reverse osmosis system. The basic equations are shown below and represent solvent (water) production and solute (salt) concentration in the permeate streams. With the membrane thickness fixed and operating at a constant temperature, product water flux is controlled by the transmembrane pressure differential. Please note that these design calculations are based on constant temperature and membrane pressure differentials. When operational, both of these conditions will change either by seasonal fluctuations or by fouling of the membrane surfaces. However, in order to acquire baseline data to track system performance, it is necessary to initially hold some values constant. The method for tracking temperature and pressure as variables will be shown later in this paper. The product water flux is indicated in Equation 2.

$$J = A(TNP - NOP) \qquad (2)$$

where, J = product water flux, gallons per square foot per day

A = water permeability coefficient, gallons per square foot per day/psig

TNP = transmembrane pressure, psig

NOP = net osmotic pressure, psig

The transmembrane pressure differential for the system is determined by the difference of the feed and reject pressure values. For our application, the effect of feed water solution net osmotic pressure can be neglected since the osmotic pressure developed by freshwater is negligible compared to the transmembrane pressure differential. When solution osmotic pressure is unknown, but is desired to be considered, a useful rule of thumb is that for every 100 ppm Total Dissolved Solids in the feed and reject streams, an osmotic pressure of 1 psi is developed. Appendix A.1 shows the actual calculations for city water osmotic pressure determination. The estimated average osmotic pressure would then be subtracted from the calculated transmembrane differential pressure to provide the effective transmembrane pressure. The water permeability coefficient for the membranes selected is 0.235 gallons per square foot per day / psig. The following data was applied to Equation 2 to determine the product water flux for our proposed system:

- Total transmembrane pressure differential of 70 psig
- Net osmotic pressure differential of 3.75 psi. This was determined from an inlet Total Dissolved Solids of 150 ppm (1.5 psi osmotic pressure) and a reject total dissolved solids of 600 ppm (6.0 psi osmotic pressure). Typically, a reverse osmosis system will concentrate solute four times in the reject stream, therefore, the reject osmotic pressure will be four times greater than the influent. Applying this data to equation 2 provides a product flux of 15.6 gallons/ square foot/ day.

The salt flux or solute flux through the reverse osmosis membrane is dependent on the concentration of the influent stream and the reject stream, as indicated in Equation 3.

$$S = B\,(COF - COP) \tag{3}$$

where, S = solute flux, pounds per square foot per day

B = solute mass transfer coefficient, gallons per square foot per day

COF = concentration of feed water, ppm

COP = concentration of permeate, ppm

During normal operation, a boundary layer of concentrated salts is formed at the membrane surface. The concentration of dissolved salts in this layer is significantly higher than the concentration in the bulk feed solution. Therefore, the solute flux or membrane salt rejection ability is influenced by the concentration difference between the polarized boundary layer on the membrane surface and permeate. The extent of concentration polarization that occurs in the boundary layer is influenced by the membrane design and the system operating conditions. The selected membranes are designed to operate in the range of feed flow velocities and product recoveries where the boundary layer concentration effect is minimized. Since product recovery affects the extent of concentration polarization that occurs, it is necessary to measure the salt concentrations in the feed and product water streams at the system baseline product recovery condition to determine the salt rejection ability of the membrane. The salt rejection, claimed by the membrane manufacturer, is 98%. This rejection rate is again a baseline value and will vary with membrane fouling and operating conditions. The station stipulated a 95% minimum salt rejection. This number is more realistic for extended operations of 4-5 years. This is not to say that 98% salt rejection can not be obtained, on the contrary, initially 98% or better is achievable, but as the membranes age and become fouled it is not possible to maintain the design solute flux and the rejection rate gradually declines. Equation 4 is used to determine the salt or solids rejection.

$$\frac{(\text{feed water solids}) - (\text{product solids})}{(\text{feed water solids})}(100) = \text{Percent solids rejection} \qquad (4)$$

Subsequently, the following data was used for equation 3, influent concentration of 150 ppm, product concentration of 3.0 ppm, and a mass transfer coefficient of 2.176x E-6 gallons/ square foot /day provides a solute flux of 0.000321 pounds/square foot/day.

The previously stated equations, 2 and 3 are useful for membrane design selection. Equation 5 is used to determine the total area of membrane surface to produce the desired output.

$$A = Q/J \tag{5}$$

where, A = membrane surface area, square feet

Q = flow rate, gallons per day

J = product water flux, gallons per square foot per day

Once the total surface area of membranes is determined, the number of membrane elements can be determined by dividing the area of each element into the total area required. Each element in our selection contained 400 square feet. In order to produce 150 gpm or 216,000 GPD, at a product flux of 15.6 gallons/square foot/day, a minimum of 13,846 square feet of membranes are required when using equation 5. The actual total membrane area will be designed higher.

In order for a feed water source to be filtered by a semi-permeable membrane, pressurized feed water must come into contact with the membrane. Additionally, permeate and the concentrate must be transported away from the membrane. This is accomplished by placing the membrane inside a housing called a membrane element, or simply, an element. Reverse osmosis elements are coupled together inside a fiberglass pressure vessel called a module. The number of elements per module ranges from 1-7, depending on the size of the system; we employed seven elements. Figure 6 illustrates a reverse osmosis element. This configuration is commonly employed to designate elements.

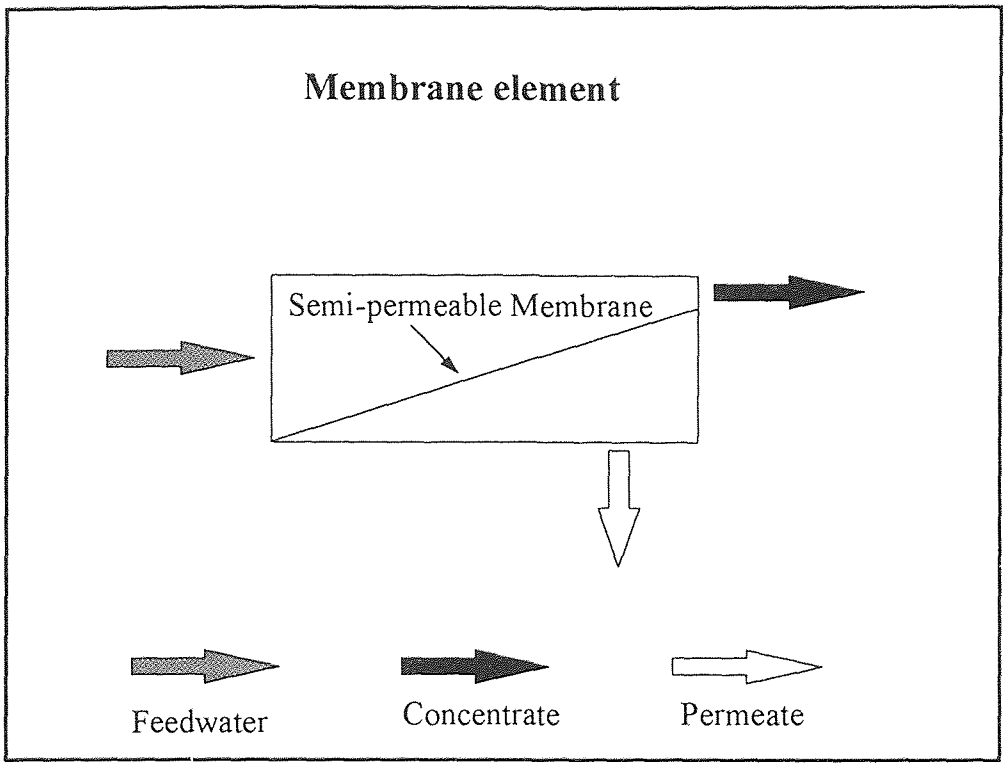

Figure 6 Reverse osmosis element

The limitation of elements per pressure vessel is caused primarily by the need to maintain a relatively high feed water/waste flow rate through the system. The relatively high flow rate promotes turbulence which minimizes scaling and fouling. A pump pressurizes the feed water and forces it into each pressure vessel. Approximately 5-10% of the feed water which enters each element is forced through the membrane. This is called a 5-10%

recovery. This gives an overall recovery for a seven element vessel of around 50%. Fifty percent recovery concentrates the feed water twice. With proper system design, most feed water can be concentrated four or five times before scaling occurs. The Total Dissolved Solids in our water supply allows the membranes to concentrate the feed water four times providing a 75% recovery rate. Concentration / recovery rate relationship is shown in Appendix A.2. This is beneficial for the station and the vendor since the station can recover a high percentage of the feed water as product while the vendor has less concern with scaling and fouling the membranes. The recovery rate is usually a negotiating point between the end user and the vendor. The vendor wants to minimize the recovery to control membrane fouling while the end user wants to maximize recovery to reduce the volume of the waste stream. A recovery of 70-75% is a fair compromise for both parties and was stipulated in our bid specifications.

The first seven element module recovers 50%, the concentrate from these first modules are manifolded together and become feed water to another set of modules. This is called staging. The first set of modules is called the first stage; the second set of modules is called the second stage. There are two first-stage modules for every second-stage module. Figure 7 illustrates the staging process and related topics.

As previously discussed, product recovery is the ratio of water volume recovered as product water compared to the volume of water supplied to the reverse osmosis as feed water. In an ideal situation you would recover all the water, but this is not practical since if all the water was processed and converted to product water, there would be no water available to flush the membrane surface area free of remaining suspended solids.

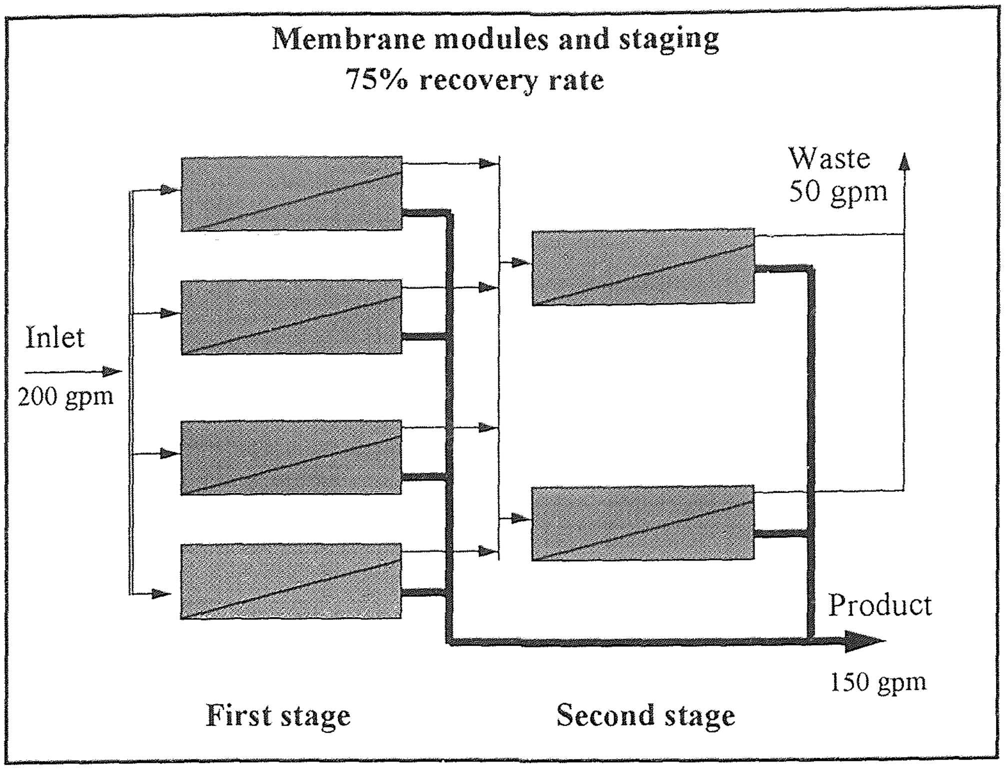

Figure 7 Membrane modules and staging

Consequently, the suspended solids would build up on the surface of the membrane very quickly and gradually restrict the flow of product water until no permeate could be produced. Percentage of water recovery is determined by Equation 6.

$$\text{Percent recovery} = \frac{(\text{feed water flow rate}) - (\text{concentrate flow rate})}{(\text{feed water flow rate})} \times 100 \qquad (6)$$

The station requires 150 gpm from each of its reverse osmosis units. In order to meet this requirement and to maintain a 75% recovery rate, 200 gpm of feed water will be supplied. This leaves 50 gpm as the concentrate waste stream.

The calculations and design methods shown in this section were for demonstrative purposes. Reverse osmosis system design is more commonly accomplished through computer driven software. Appendix A.3 contains computer generated design data for the station's system. The software is provided by courtesy of the Polymetrics Corporation. This software was instrumental for the design of our system and for providing theoretical data and analysis to which actual data could be compared. Table 1 shows the reverse osmosis system components.

Table 1 Reverse osmosis system overview

Membrane manufacturer	Fluid Systems
Membranes type	Thin Film Composite / spiral wound
Product flux	15.6 gallons/square foot/day
Membrane area / element	400 square feet
Number of elements/ module	7
Number of modules / unit	6
Total membrane area / unit	16,800 square feet
Maximum flow rate / unit	262,080 GPD / 182 gpm
Element length	40 inches
Element diameter	8 inches
Design operating pressure	225 psi
Maximum system pressure	330 psi
Pump manufacturer	Tonkaflo
System arrangement	4 modules first stage 2 modules second stage
Recovery rate	75 %
Product flow rate	150 gpm
Concentrate flow rate	50 gpm
Feed water flow rate	200 gpm
Solute flux	0.000321 gallons/square foot/day
Rejection rate	>98 %

CHAPTER 3

PRETREATMENT SYSTEMS

3.1 Pretreatment Overview

The successful long term performance of reverse osmosis membranes depends primarily on proper pretreatment. Without proper pretreatment design, all reverse osmosis systems are destined for failure and significant downtime for maintenance activities.

Reverse osmosis, by its simplest interpretation, is a cross-flow filter, in which the liquid being filtered continuously passes over the filter surface. The filtrate passes through the membrane while the impurities traverse the surface, and in doing so, becomes more concentrated and eventually exits the filter as the concentrated waste stream. Ideally, all the impurities are carried away in a concentrated stream rather than precipitating or adhering to the membrane surface. Since, reverse osmosis is not an ideal application, proper pretreatment of the feed water can increase the efficiency and life span of the membranes. This is accomplished by minimizing fouling, scaling and membrane degradation resulting in the optimization of product flow, product recovery and salt rejection, which is directly related to operating costs and system efficiency.

A marginal pretreatment system will result in excessive cleaning operations, which should not be regarded as a substitute for proper pretreatment. In essence, cleaning does not restore 100% of the efficiency of the membranes, therefore, each subsequent cleaning operation will slightly diminish overall efficiency.

All types of fouling have in common either trapping a material within the reverse osmosis membrane, or chemically depositing on the surface. A non-fouled system operating at constant conditions will remain essentially unchanged with time. An increase in salt passage or an abnormally high transmembrane pressure drops usually indicates fouling. Some examples of fouling are; deposition of organic compounds, metals, biological activity, and large particulate matter. On the other hand, membrane scaling is the result of precipitated dissolved solids as the feed water is concentrated. This concentration is magnified by the tendencies of the salts to polarize on the membrane surfaces. The scaling tendency of water, for our application, was determined using the Langelier Saturation Index, which will be discussed further. Having discussed the difference between fouling and scaling, the methods for controlling these occurrences will be detailed separately.

3.2 Membrane Fouling

Pretreatment for fouling must be a total system approach for continuous and reliable operation, in that, improper design of upstream filters will necessitate the frequent cleaning of downstream filters. The water supply used at the Hudson Generating Station is from the Boonton Reservoir. As a surface water, it typically contains a moderate amount of suspended and colloidal solids, with low calcium and alkalinity levels. The ASTM Committee on Water has proposed a simple qualitative test that can be used in the field to indicate fouling potential due to suspended and colloidal solids. The test is used to evaluate the effectiveness of various pretreatment systems and to predict the extent of membrane fouling and expected frequency of cleaning when operating a reverse osmosis system from a particular water supply. The test is referred to as the "Microporous

Membrane Pluggage Test", or more commonly known as the Silt Density Index (SDI). The test is based on determining the rate of pluggage of a 0.45 micron membrane filter when a known quantity of water passes through at a constant pressure. The SDI is a non-dimensional number, calculated from the rate of pluggage, and is used by most membrane manufacturers as an indication of the maximum level of suspended solids allowable in a feed water in order to maintain membrane performance warranties. There is a general correlation between SDI levels and the degree of fouling. Spiral wound polyamide configuration requires an SDI less than 3.

SDI testing at the station was conducted over a three month period to obtain an average influent SDI value. Samples of raw water were also collected, and analyzed off-site, through pilot treatment systems to determine the best pretreatment and media required to lower SDI values within membrane manufacturer's guidelines. An explanation and procedure will be detailed in the Methodology and Testing Chapter.

The pretreatment at the station consisted of multimedia filters and cartridge filtration. The multimedia filters consist of six skid-mounted vessels containing multilayered filter media. The vessel are constructed of schedule 80 carbon steel and are internally coated with epoxy for corrosion protection. Each vessel is 48 inches in diameter and 60 inches high. The normal system flow rate is 400 gallons per minute. Regeneration or backwashing is initiated when the differential pressure exceeds 10 psig; the system employs differential pressure switches for this purpose. All valving is pneumatically controlled and receives an electrical signal from the main panel programmable logic controller to initiate regeneration, rinsing, or in-service valve sequencing. The logic controller can be programmed on-site for field adjustments. The filters are operated in parallel with all six vessels on line and in service. Regeneration is sequential with one unit backwashing, and the other units remaining in service. It's estimated that the filters will require weekly backwashing. Table 2 describes the type and amount of media in each vessel.

Table 2 Filter media description

Layer 1	Gravel 1/4x1/8 inches	10 cubic feet
Layer 2	Garnet 8-12 mesh	2.86 cubic feet
Layer 3	Garnet 30-40 mesh	6.4 cubic feet
Layer 4	Sand 0.45-0.55 mm	12 cubic feet
Layer 5	Anthracite #1	19 cubic feet

The layers are arranged with layer 2, small mesh garnet positioned on top of the gravel support, then progressing upward with the larger mesh garnet, sand and finally anthracite. This arrangement will assure that larger particles will be entrained in the anthracite and sand without plugging the garnet, leaving the garnet available to filter the smaller particles in the feed stream. Figure 8 illustrates the filter.

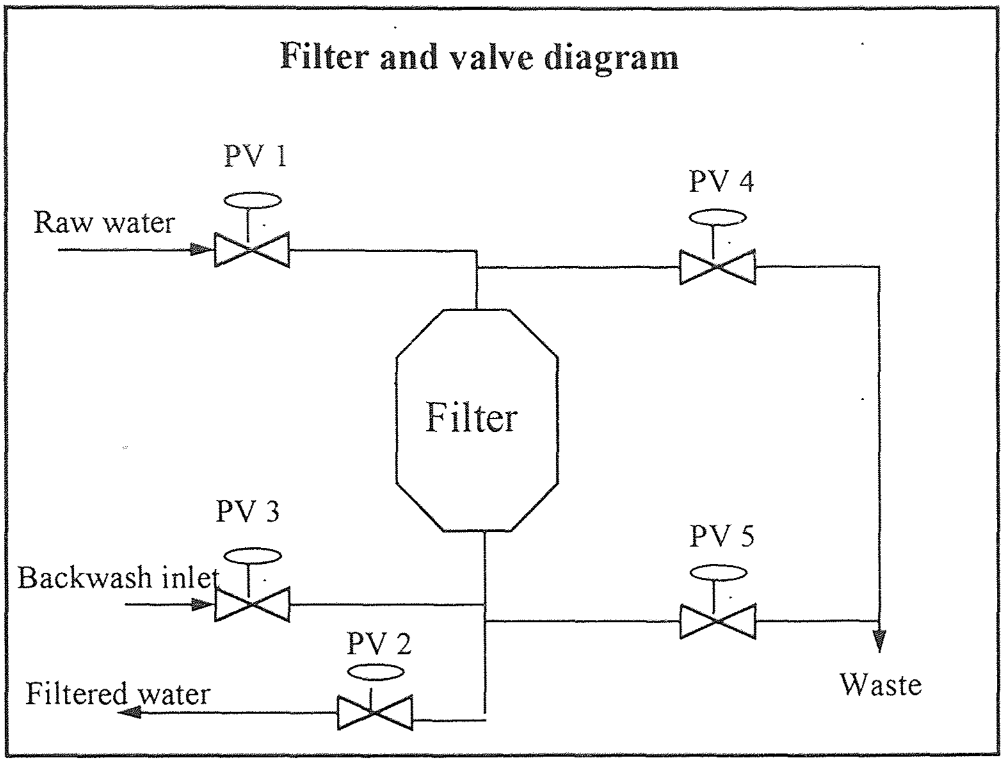

Figure 8 Filter and valve arrangement

The normal raw water influent flow for this vessel is through pneumatic valve PV 1 into the vessel, and exiting through PV 2 to the cartridge filters. The normal flow is 65-70 gallons per minute per vessel. During backwashing of one of the other vessels, this flow can increase to 80 gallons per minute. Backwash operations are initiated by closing PV 1 and PV 2, and opening PV 3 and PV 4. This allows water to enter through the bottom of the vessel and lifts the media to scour and remove the entrained suspended solids. Backwashing continues for approximately 20 minutes at 180 gpm for a total backwash volume of 3600 gallons. The backwash inlet line is equipped with a flow restricting orifice to protect the integrity of the layers. After completion of the backwash cycle, the fast rinse cycle begins. PV 3 and PV 4 close, PV 1 and PV 5 open. The fast rinse continues for 6 minutes at 120 gpm, for a total fast rinse volume of 720 gallons. The fast rinse cycle serves the dual purpose of compacting the filter media and washing out the remaining suspended solids before the unit is returned to service. After this cycle, PV 5 is closed, and PV 2 is opened. The system is ready for service. Total backwash time is approximately 26 minutes, with a total volume of 4320 gallons.

The effluent of the multimedia filters will produce a water with suspended solids no greater than 10 microns in size. The effluent of the multimedia filters is sent to 1 micron absolute cartridge filters. This system consists of four vessels, each equipped with six polypropylene woven cartridge filters. The filters are two inches in diameter and 30 inches long. Water enters the vessel and filters into an inner filtrate collection tube. The cartridges are woven so that gradually the weave constricts to increasingly smaller pore size. This will allow for the collection of larger particles on the outer surface of the filters, leaving the inner surfaces available to collect and entrain the smaller particles. The cartridges are replaced manually when the differential pressure increases to 10 psig. The absolute rating indicates that all particles greater than 1 micron are guaranteed to be removed from the feed stream.

As previously stated, the SDI utilizes a 0.45 micron filter. The reverse osmosis membrane pore openings are 0.0005 microns, which is a significantly smaller opening. Therefore, the argument taken by some, is what easily passes through the SDI membrane filter test would readily foul a reverse osmosis membrane, and puts certain doubts as to the credibility of the SDI test. On the other hand, the argument is, that the SDI is a counter flow test, whereas, reverse osmosis is a cross-flow process. What may be caught in the membrane filter for the SDI test would not necessarily be entrained on the reverse osmosis membranes. Although the SDI measurement is not 100% accurate and there are questions of its validity, it does provide a guide in determining the magnitude of fouling from suspended and colloidal solids. Empirical correlation between the SDI and fouling has shown that the SDI must be 3 or less to minimize the rate of fouling of thin film composite membranes and to obtain successful long term (4 years or longer) performance.

3.3 Membrane Scaling

Scaling is due to concentrating the ions of the feed water beyond the saturation point of a soluble salt. Scaling usually occurs in the second stage where the feed water is most concentrated. In reverse osmosis, dissolved solids are concentrated, depending on the percent of recovery and the influent salt concentration. For our water supply, calcium carbonate precipitation is the mineral of concern.

The method used to determine the potential for calcium carbonate precipitation is the Langelier Saturation Index. The Langelier Saturation Index (LSI) was derived from the second ionization constant of carbonic acid and the solubility product of calcium carbonate, therefore, the index uses calcium hardness, total alkalinity, pH, Total Dissolved Solids, and temperature to determine whether precipitation will occur. Equation 7 illustrates the LSI. Information gathered for this equation was collected from the Permutit *Water and Wastewater Treatment Data Book.* It is not in the scope of this report to list all the tables used to calculate the LSI, but this book, as well as many others, can be used to reference the appropriate tables.

$$pH\ saturation = (9.3 + A + B) - (C + D) \qquad (7)$$

where, A = total dissolved solids ppm

B = temperature, degrees Fahrenheit

C = calcium hardness, ppm as calcium carbonate

D = total alkalinity, ppm as calcium carbonate

$$Saturation\ index = pH\ water - pH\ saturation$$

Saturation Index Value	**Interpretation**
Negative value	Dissolve calcium
Zero value	chemical balance
Positive value	Precipitate calcium

The values used in this equation to determine the LSI were derived from the raw water influent. These values were then multiplied by four to project the concentration in the second stage elements. Appendix B.1 shows the analytical concentration values. The projected values are: total dissolved solids 600 ppm, calcium hardness 152 ppm, total alkalinity 140 ppm, temperature, a maximum of 70 degrees F., and pH of 7.3. Utilizing these values, the LSI is projected to be - 0.31. Since the LSI is within the calcium dissolving range, and in conjunction with the turbulent conditions generated in the elements, it was decided not to employ sulfuric acid injection at start-up. The design option was to operate the system without any scale inhibition, but to monitor performance, and implement acid injection if required.

There are two methods for controlling calcium carbonate precipitation, polymer scale inhibitor and acid. Polymers will tolerate a LSI of (+1.5) to (+ 2.0) without calcium precipitation. It was decided not to use polymers since they typically contain either organic or inorganic phosphates, or organic acids and the station is not permitted by the New Jersey Department of Environmental Protection (NJDEP) to discharge these compounds from our waste treatment plant into the Hackensack River. If required, sulfuric acid would be injected prior to the cartridge filters to promote turbulence and rapid mixing of the acid with the water. A pH probe, located after the cartridge filter housing, sends an electrical signal to the main control panel, which in turn sends an electrical signal to the injection pump. As the pH rises above or falls below the set point, the pump rate increases or decreases respectively. Adjusting the feed pH by acid addition converts the bicarbonate ions to carbon dioxide, thereby reducing the potential of calcium carbonate formation. Calcium sulfate, a more soluble salt, is produced from this reaction. In the event acid injection is required, the influent pH would be maintained at 5.5 to 6.0. Appendix B.2 shows the LSI at multiple pH ranges for the raw water supply.

Other mineral constituents can precipitate and scale membranes; three of the most common are silica, strontium, and barium. The concentrations of barium and strontium in our influent, when concentrated, are well below the solubility limits of 0.20 ppm barium sulfate and 10 ppm strontium sulfate. Silica can concentrate to 150 ppm before precipitation will occur. As Appendix B.1 shows, actual values are below the solubility limits.

From an ion exchange perspective, it is undesirable to use sulfuric acid since carbon dioxide is generated from the acidification of bicarbonate alkalinity.

$$Ca(HCO_3)_2 + H_2SO_4 \rightleftarrows CaSO_4 + H_2O + CO_2$$

The reverse osmosis membranes do not remove carbon dioxide gas, allowing the gas to react downstream to form bicarbonate alkalinity. This alkalinity will be ionically removed by the anions in the demineralizers.

$$CO_2 + H_2O \rightleftarrows H_2CO_3 \rightleftarrows H + HCO_3$$

This is undesirable since it reduces anion capacity and results in a shorter demineralizer service cycle. For this reason, a forced draft decarbonator/clear well was incorporated in the design to reduce the level of carbon dioxide in the product water. Decarbonation occurs when the water is sprayed into the top of the decarbonator tower and flows downward over the surface of the tower packing material. As the water flows down through the packing material, filtered air is forced upward. The counter-current flow of air and water interact at the surface area provided by the packing, and carbon dioxide is stripped from the water during this interaction and exits through the tower vents. Water is collected in the clearwell beneath the decarbonator section. The decarbonator used for our application is constructed of fiberglass, has ten feet of packed material in the column, and is designed to remove carbon dioxide from an influent of 20 ppm to approximately 5 ppm.

The final pretreatment system is the plants' existing carbon filters. The filter's sole purpose is to dechlorinate the city water prior to being used in the reverse osmosis system. Thin film composite membranes are easily oxidized by either free or combined chlorine.

Each of the two vessels contains 230 cubic feet of Calgon 400 FS activated carbon. These filters are capable of removing total chlorine residuals to below 0.1 ppm, which is an acceptable influent for the membranes. Table 2 summarizes the pretreatment system components.

Table 3 Pretreatment system components

Filter vessel manufacturer	Lakeside Water Treatment
Vessel size	48” diameter / 60” high
Number of vessels	6
Media type	gravel, sand, garnet, charcoal
Design flow rate	400 gpm
Backwash procedure	Automatic
Backwash initiation	10 psig pressure differential
Backwash volume	4320 gallons / vessel
Removal efficiency	95 % > 10 microns
Cartridge filter manufacturer	Filterite
Number of cartridge filters	4
Cartridges / vessel	6
Cartridge size	2” diameter / 30” long
Removal efficiency	1 micron absolute
Acid injection	Baume 66 sulfuric acid
Acid pump manufacturer	Liquid Metronics Inc.
Pump capacity	Proportional 0-14 GPD
Decarbonator manufacturer	Polymetrics
Design flow rate	300 gpm
Construction material	Fiberglass
Blower design	1600 cfm at 3” of water column
Clearwell capacity	1000 gallons

CHAPTER 4

TEST METHODOLOGY

4.1 Test Methods

To evaluate the performance of the reverse osmosis system, it was decided to utilize both on-site and outside laboratories for water analyses. The station is equipped with facilities to test reactive silica, conductivity, pH, SDI, chlorine, and carbon dioxide. Outside laboratories were used for more extensive analytical testing and to verify on-site results.

This chapter is grouped in two sections; continuous and grab sample testing. Continuous testing, silica, pH, and conductivity, were employed to determine the performance of the reverse osmosis system. Grab sample testing, SDI, carbon dioxide, and chlorine, were utilized to monitor the efficiency of the pretreatment systems.

The plant is a certified lab for pH, temperature, and chlorine as per our New Jersey Pollution Discharge Elimination System (NJPDES) permit. PSE&G's, state certified, Research and Testing Laboratory in Maplewood, New Jersey was the outside laboratory used in this report. This chapter will detail only on-site test methods.

4.2 Continuous Monitoring

The Rosemount model, Compu-sol, was used for conductivity analysis and is based on *Standard Methods* "Conductivity 2510 Laboratory Method." This meter measures the ability of a water to conduct electrical current, and is a direct measurement of the total ionized (dissolved) solids in the water. Conductance is inversely proportional to electrical resistance ie, the higher the water purity, the higher its resistance to passage of an electrical current, and therefore, has a lower conductivity. Inversely, the higher the mineral content of the water, the higher the conductivity. The test is not specific for any one ion, but rather is a measure for total ionic concentration. The basic unit of electrical resistance is the ohm; since electrical conductivity is the reciprocal of resistance, the term mho, is used as the basic unit of conductivity.

The Rosemount meter incorporates two platinum electrodes, spaced 0.01 centimeters apart. The electrode is housed in a stainless steel cell holder. Effluent water from the reverse osmosis system is transported through stainless steel tubing to the bottom of the cell holder; the system is hard-piped to prevent atmospheric carbon dioxide from dissolving in the water stream and interfering with the conductivity test. As water flows up past the cell, the conductivity of the water is measured between the two electrodes. This conductance provides a millivolt signal, which is amplified by a pre-amp and transmitted to the meter's electronics. This millivolt signal is converted to a digital readout. The resolution of the meter is 0.01 micromhos, with an accuracy of 1%. The separation of the electrodes dictates the meter's range. We selected a separation of 0.01 centimeters, making our range 0.01 to 100 micromhos. Since conductivity is temperature dependent, the unit incorporates an internal temperature sensor, which automatically adjusts the conductivity readings to 68 degrees F. The compensation range is between 32-145 degrees F.

On occasion, a hand-held conductivity meter, manufactured by the Myron L Company, was also utilized for water analysis. This meter was used for higher conductivity samples, such as city water, and concentrate waste from the reverse osmosis system. A hand-held meter was used for this application because carbon dioxide interference at these high conductivity levels is considered negligible. This meter works on the same principle as the Rosemount model.

pH measurements were necessary to control acid injection. The analysis is based on *Standard Methods* "4500-H+ B Electrometric Method", pH measurements were determined using Great Lakes' model 670 pH meter. This method utilizes two electrodes which are in contact with a potassium chloride electrolyte solution. The voltage of the electrode known as the reference electrode is fixed, while the voltage of other electrode varies with the hydrogen ion concentration of the sample. The voltage difference between these two electrodes is dependent on the hydrogen ion concentration of the water. The difference in potential generates a millivolt signal which is detected by a potentiometer. The signal is then amplified to a voltage signal by a pre-amp, and is sent to the meter's electronics, where it is converted to a digital output The meter has a resolution of 0.01 pH units with an accuracy of 1%. Since pH is temperature dependent, the meter incorporates an internal temperature device to compensate for and modify varying water temperatures.

Continuous silica analysis of the reverse osmosis effluent stream was accomplished using the Hach Series 5000 silica analyzer. This analyzer is a continuous, wet chemical, colorimetric determination of reactive silica, which will determine silica concentrations between 0-5000 ppb at an accuracy of 1 ppb, and a resolution of 0.01 ppb. The analysis is performed using *Standard Methods* "4500-Si D Molybdosilicate Method." The unit provides semi-continuous analysis of a water sample stream by measuring discrete samples an 8 minute cycle. A programmable automatic calibration system is provided to ensure continuous accuracy.

The molybdosilicate method is used to measure molybdate-reactive silica. The sample is collected in a 50 ml sample cell to which acidic molybdate solution is added to react with any silica and phosphate present to form molybdosilicic and molybdophosphoric acids. After a delay for mixing the reaction, citric acid/surfactant reagent is added. Citric acid will mask molybdophosphoric acid present and will react with excess molybdate. This prevents molybdate from producing an interfering blue colored compound. The surfactant, a wetting agent, will minimize air bubble formation on the sample cell walls. Light absorbance through this solution is measured to determine a sample blank reference absorption. Color formed at this point is identical to the final color of a zero ppb silica sample. This provides a zero reference and will compensate for any background turbitity and color inherent in the sample. Finally, amino acid is added to reduce molybdosilicic acid to a blue colored solution. The amount of color formed is directly proportional to the silica concentration of the sample. Light absorbance through the solution is measured at 810 nm. This absorbance is compared to the sample blank reference absorbance, and silica concentration is calculated accordingly.

The analyzer's operations are fully controlled by the system's programmable logic controller and electronic solenoid valves, which allow sample and chemical reagents to enter the cell at the proper times. The logic controller activates the cell mixer and adjusts the lamp intensity. Reagents are supplied to the analyzer by pressurizing the reagent containers and using solenoid valves to regulate reagent dispensation. The reagent containers are enclosed in a separate reagent compartment, located directly beneath the cell holder. The reagents are supplied in 2 liter bottles which are sufficient for a month of continuous operation. The analyzer is automatically calibrated weekly with a 500 ppb standard solution. The logic controller will automatically lock out the sampling system, on the date and time entered by the user to initiate calibration. The analyzers' electronics will automatically calibrate the analyzer according to the calibration results, after which it resumes normal sampling operations. Another feature of this unit we found most beneficial was its ability to analyze grab samples. The analyzer incorporates an inlet funnel mounted above the sample cell. Since reagent volume and mixing times are held constant, precision is enhanced by eliminating the potential for operator error. We found this unit to be extremely accurate and helpful in gathering silica data.

4.3 Grab Sample Testing

The Silt Density Index (SDI) test follows methods prescribed by the *American Society of Testing and Materials Committee on Water* "Microporous Membrane Pluggage Test." The method is based on determining the rate of pluggage of a 0.45 micron membrane filter when water is passed through the filter at a constantly applied pressure. The SDI is calculated from the percentage of flow decay over a period of time at 30 psig. The membrane fouling test consists of an inlet ball valve, a pressure regulator to maintain 30 psig, and a filter container with a removable top for setting the 0.45 micron membrane filter. Figure 9 illustrates the SDI test equipment.

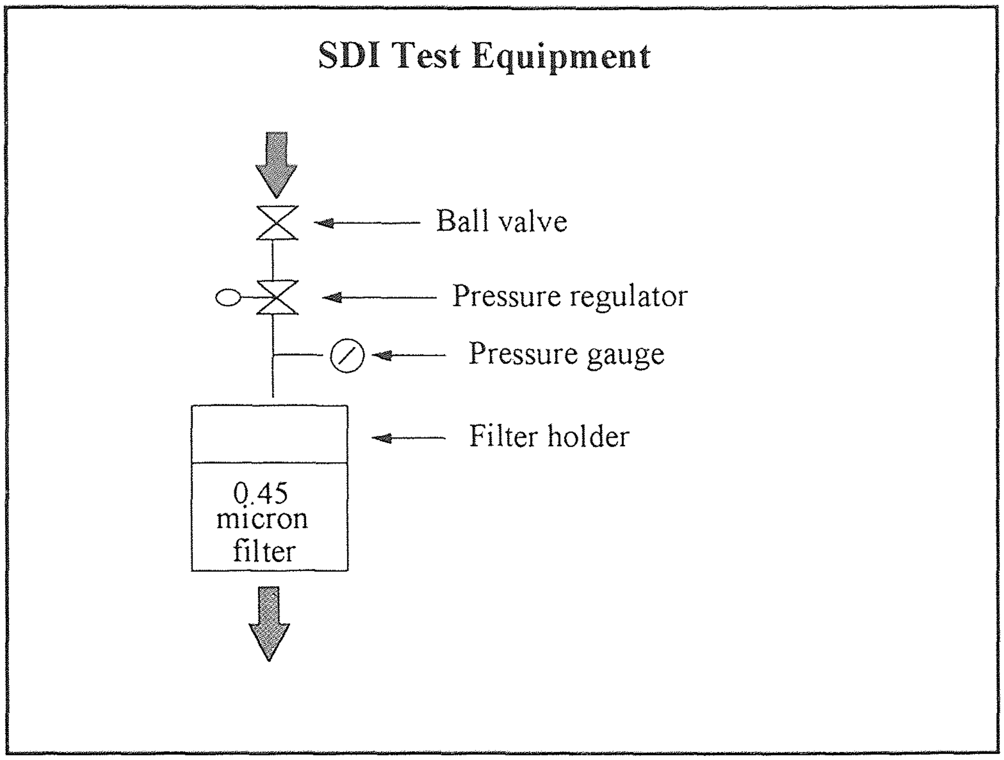

Figure 9 SDI test equipment

A digital stop watch that records seconds is essential to optimize test efficiency. The test will determine flow decay over a 15 minute period. SDI test procedures are as follows:

1. Install the SDI tester without a 0.45 micron filter.
2. Turn on the water supply and open the inlet valve. Allow water to flow through the tester for five minutes to flush out the tester, tubing, and valves.
3. Close the inlet valve, and using tweezers, place a 0.45 micron filter into the filter holder. Tighten the filter holder.
4. Fully open the inlet valve and adjust the pressure regulator to 30 psig. Once the pressure is set, close the inlet valve.
5. Use a 250 ml graduated cylinder to collect the water sample.

6. Open the valve, and with a stop watch, immediately measure the time required to collect a volume of 200 ml and record that time as T(0) seconds. Leave the valve open and the water flowing after the 200 ml mark has been reached.

7. Allow the water to flow and immediately at fifteen minutes of elapsed test time, measure the amount of time in seconds it takes to fill the beaker to the 200 ml mark and record as T(15) seconds.

8. Using the time recordings from the above test procedure, insert these values into equation 8, as follows:

$$SDI = [\ 1 - \{\ T(0)\ /\ T(15)\ \}\ \{100\}\ ]\ /\ \text{elapsed time} \qquad (8)$$

example: $T(0) = 10$ seconds
$T(15) = 40$ seconds
elapsed time = 15 minutes
$SDI = [\ 1 - \{10/40\}\{100\}]\ /\ 15\ =\ 5$

Note: If complete pluggage occurs before 15 minutes of elapsed time, then use 10 minutes. Do not use less than 5 minutes.

The carbon dioxide test was used to periodically determine decarbonator efficiency. The analysis for carbon dioxide utilized *Standard Methods* "4500-CO$_2$ C Titrimetric Method" for free carbon dioxide. A water sample is titrated to the phenolphthalein endpoint with sodium hydroxide standard solution. The reaction of sodium hydroxide with carbon dioxide, as carbonic acid, occurs essentially in two steps;

first, a reaction with carbonic acid produces bicarbonate; second, the further addition of sodium hydroxide produces carbonate. Because the conversion of carbon dioxide to bicarbonate is complete, at a pH of 8.3, phenolphthalein can be used as a color indicator for this titration. The sodium hydroxide must be of high quality and free of sodium carbonate. The test procedure is, as follows:

1. Fill the mixing bottle to the 15 ml mark with a water sample.
2. Add one drop of phenolphthalein indicator to the sample.
3. Add 0.01 N sodium hydroxide solution drop by drop, counting each drop.
4. Swirl the bottle gently to mix after each drop is added. Continue adding drops until a light pink color forms and persists for 30 seconds.
5. Each drop of sodium hydroxide solution is equal to 2 mg/l of carbon dioxide.
6. Care must be taken while swirling to minimize the loss of carbon dioxide from the water sample as a result of aeration.

The samples were collected from the decarbonator clearwell and immediately analyzed on-site to prevent the loss of carbon dioxide to the atmosphere. Precision of this method is 10% of the known concentration.

The chlorine test utilized in this report determined total chlorine residual from the effluent of the carbon filters. The analysis utilized *Standard Methods* "4500-Cl G DPD Colorimetric Method." The test for total chlorine utilized the Hach N,N-diethyl-p-phenylenediamine (DPD) colorimetric method. DPD is readily oxidized by free chlorine to form a red color that is directly proportional to the chlorine residual. Potassium iodide is added to the reaction to determine combined chlorine residuals. Combined chlorine oxidizes the iodide to iodine; the liberated iodine reacts with DPD to form a red color. The red color is compared colorimetrically with standard solutions to determine total chlorine residual. The reagent used in this test contains both DPD, potassium iodide and appropriate buffers. The test procedure is as follows:

1. Fill the mixing bottle to the 5 ml mark with sample.
2. Add one packet of DPD total chlorine reagent powder.
3. Shake the contents vigorously and allow to stand for three minutes. The three minute reaction time is required to allow for the conversion of iodide to iodine by combined chlorine.
4. After three minutes, insert the sample cell into the standard holder and compare with standards to obtain total chlorine residual.

This test was conducted on-site to prevent the loss of chlorine to atmosphere. Please note, it was not necessary to quantify the amount of chlorine in the effluent of our carbon filters but to qualify that there was no appreciable amounts present.

CHAPTER 5

SYSTEM OVERVIEW AND CONTROL INSTRUMENTATION

5.1 System Overview

There was insufficient room inside the station to accommodate the reverse osmosis unit and ancillary equipment. A remote enclosure 30' by 70', originally used to house forklifts, was used to house the new system. PSE&G engineering modified the building with reinforced concrete floors and supports to handle the vibration of the new equipment. Station electricians installed 440 volt electric feed and safety disconnects to the building. All motors for pumps and fans utilize 440 volt feed to reduce energy requirements. Station pipe fitters were used to install 6" feed water piping from the existing carbon filters to the new building and, 4" product and concentrate piping were run to the demineralizers and the chemical waste basins, respectively. All piping was constructed of either 316 stainless steel or fibercast blend plastic, for its corrosion resistance. All piping inside the building was schedule 80 polyvinyl chloride (PVC).

Initially, city water is dechlorinated in the station's carbon filters and enters the multimedia filters. Effluent of the multimedia filters enters the cartridge filters and supplies feed water to the reverse osmosis feed pump. The station decided to use two 150 gpm reverse osmosis units. Normal station water consumption is approximately 125 gpm, requiring one reverse osmosis unit, or skid, in-service. On occasion, water consumption increases to 200+ gpm, in which case the second reverse osmosis unit would be activated for service. After the feed water is pressurized and purified in the reverse osmosis skid, the product water enters the decarbonator. The decarbonator clearwell acts as a storage tank for the product water and is pumped by either of two 300 gpm centrifugal pumps to the demineralizers. Figure 10 illustrates this flow pattern.

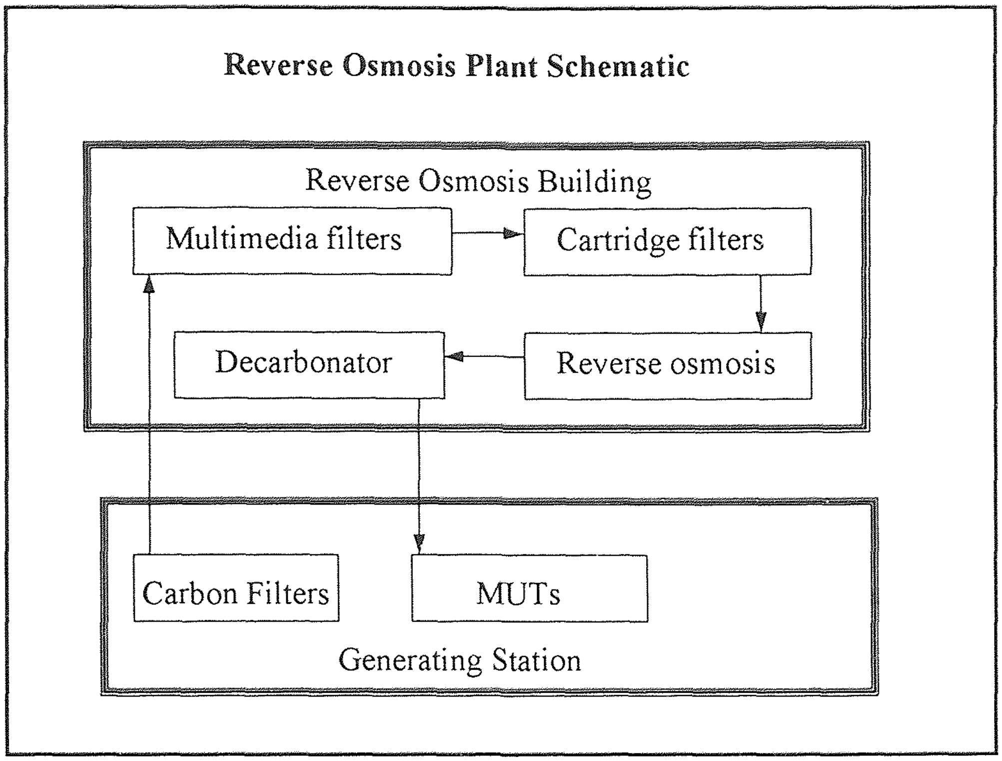

Figure 10 Reverse osmosis plant flow pattern

Located on the discharge of these pumps is a pneumatic actuated throttling valve. This valve receives an electrical signal from the clearwell level controller, and the signal, depending on the level in the clearwell, will open or close the valve to maintain approximately 70-75% level at all times. Consequently, the pump will supply either 150 or 300 gpm, depending on station demand. Each pump is equipped with a check valve on its discharge to prevent back flow from the demineralizer storage tanks when the reverse osmosis system is off-line.

Additional equipment supplied to this building was two gas fired space heaters with thermostatic controllers to maintain building temperature at approximately 60 degrees F. This was provided for concern of freezing the membranes if the system was off-line for an extended period of time. Since sulfuric acid might be used in this system, an eyewash and safety shower system was installed next to the acid storage area. Acid, supplied in 55 gallon drums, was placed over a teflon-lined secondary containment, to contain a spill. Lastly, new locks were installed for system security.

All equipment, reverse osmosis units, multimedia filter vessels, cartridge filter skids, decarbonator, discharge pumps and control panels were provided, pre-fabbed and skid-mounted by the vendor, for quick and easy installation. Construction started in December, 1993, and was completed in May, 1994.

5.2 Plant Automation

The determining factor for automating the initiation and deactivation of the reverse osmosis units was the level in the demineralization storage tanks. Each tank is equipped with a Rosemont differential pressure level transmitter. Differential pressure, as determined by the transmitter, is due to the static head in the storage tanks. Each tank is 33' high, or 396 inches of water column. The transmitter is calibrated between 0-396 inches, which corresponds to an electrical output of 4-20 milliamps (mA); such that 4 mA is equal to 0 feet of water, 20 mA is equivalent to 33 feet of water. This electrical outlet is a proportional band over its entire range. Therefore, mid-scale, 12 mA, corresponds to 16.5 feet of water. This electrical signal is sent to the station's Bailey Net90 computer to supply level indication to the control room and demineralizer plant monitoring booth. We interfaced the station computer with the reverse osmosis units' programmable logic controller. When the storage tank level falls below 24 feet, the computer sends a milliamp signal to the reverse osmosis building and activates one of the skids. This skid will stay in service until the storage tank level is again 32 feet. At that time, the skid will shut down. In the event of high water demand, such as a tube leak, and one skid cannot supply sufficient quantities of water, the second skid will be activated when the storage tank level reaches 16 feet. In this case, both skids will stay in service until the storage tank level is 32 feet.

The determining levels for activation and deactivation of the skids was agreed upon by both the station and the vendor. The vendor wanted a low tank level before activation, whereas the station desired to keep the tank levels as high as possible at all times. The vendor's concern was with frequent starting and stopping of the reverse osmosis skids because the most vulnerable time for the membranes is during startup, when the surge of pressure from the feed pump tends to imbed particles in the membrane. On the other hand, the station did not want tank levels to drop below 10 feet for concern that a vortex may be created in the storage tanks when the condenser make-up pump was activated, thereby drawing air into the condensers. Large amounts of air in the condensers would break vacuum and automatically take either or both units off- line. The final level points were satisfactory for both parties.

By automating the system, the demineralizers can be kept in service at all times. Before the reverse osmosis system, the demineralizers were removed from service daily for concern that they would prematurely exhaust during the unmanned hours and contaminate the storage tank. The demineralizer was put back in service by the next shift. This is a labor intensive operation taking at least 20 minutes, by an experienced operator. However, the reverse osmosis system alleviates this concern with premature breakthrough by its extended service cycle.

Each reverse osmosis skid can be activated automatically as previously discussed, or by a manual override. We didn't experience a need for the manual override, and the system was left in automatic for the entire evaluation period.

5.3 Monitoring Instrumentation

The following monitoring instrumentation was used to monitor performance or to notify of an alarm condition.

Stainless steel, turbine type flow meters, with an accuracy of 0.5%, were installed by the station on the feed water and permeate lines. These meters monitor the amount of water required by the reverse osmosis system and verify the vendor's product output meter. Since the station is being charged $3.17 per 1000 gallons, we installed highly accurate meters to verify their billings.

The station installed a continuous flow silica analyzer and conductivity meter on the product output piping to monitor contractual performance of the reverse osmosis system. The analyzer and meter are equipped with output signals that are sent to a Yokagowa chart recorder to continuously trend performance. The trend readings were reviewed daily by the demineralizer operator and archived.

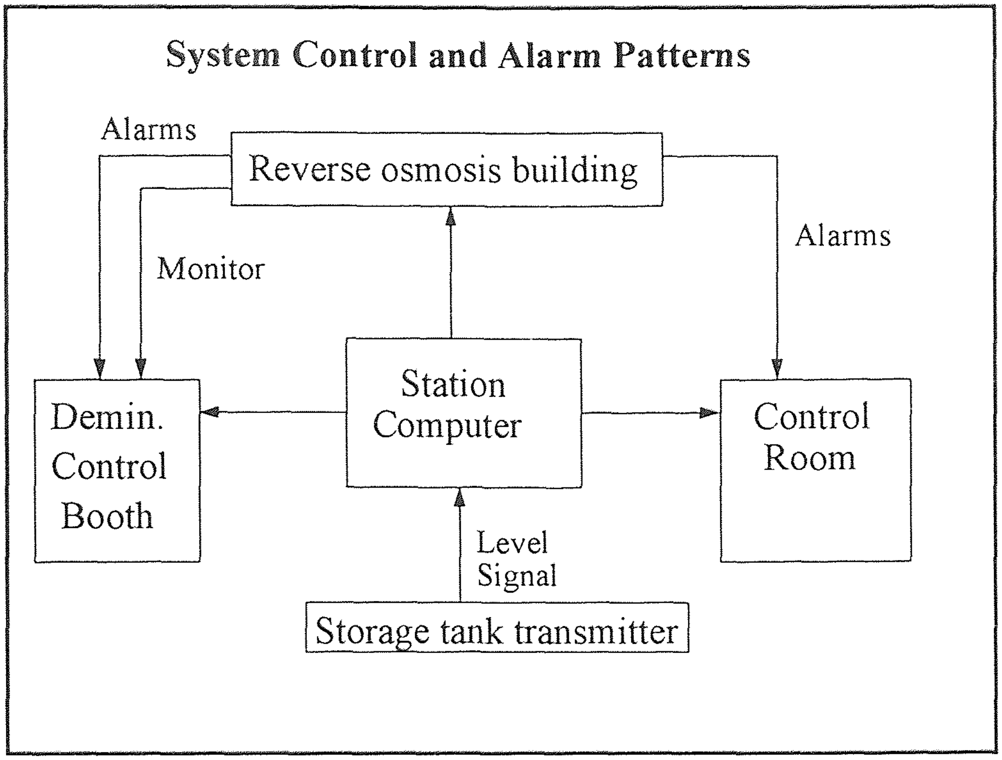

Figure 11 Reverse osmosis plant control

The control panel in the reverse osmosis building features local alarm indication and a common acknowledgment switch. Alarms activate for: high product conductivity, low building temperature, low or high feed water pH, low or high water inlet pressure, high clearwell level, and pump malfunctions. Since this building is in a remote location, and does not require continuous attention, it was necessary to install alarms that would be activated in the station's control room, which is manned 24 hours daily. It was not feasible to run separate lines to the control room for each alarm, therefore, one line was run from the panel's main alarm contact switch. When any of the local alarms activate, the main contact is also activated. In turn, the control room's annunciator panel will signal "reverse osmosis building malfunction". In the event of an alarm condition, an operator is dispatched to the building to further investigate the problem. Figure 11 illustrates the control and instrumentation systems. Table 4 shows the alarm set points.

Table 4 Reverse osmosis alarm set points

High product conductivity	10 micromhos
Low feed water pH	5.0
High feed water pH	8.0
Low concentrate pH	4.0
Low building temperature	40 degrees F
Low feed water pressure	15 psi
High feed water pressure	100 psi
Low clearwell level	45 %
High clearwell level	90 %
Pump alarm	Pump trip

CHAPTER 6

SYSTEMS PERFORMANCE

6.1 System Startup

The system was activated for testing on May 2, 1994, with a one week shakedown period to flush the membranes with city water, clean the filter media, and to configure the programmable logic controller. The membranes are flushed for 24 hours with dechlorinated city water to expand the membranes and support materials, and to remove the preserving agent in which the membranes were packed. The preserving agent used was a 1.5 % solution of sodium bisulfite and a 0.02 % solution of glutaraldehyde which prevents bacteria and algae from multiplying during storage and shipping. The media was backwashed and rinsed to remove dust and fine particles prior to service. Considerable time was spent configuring the programmable logic controller and testing all possible control and alarm scenarios to ensure adequate control while in the automatic mode. The system was put on-line and started producing water on May 9, 1994.

Reverse osmosis membranes require an initial break-in period; normally 50 hours of operating time. This allows time for expansion of the membranes and to gradually acclimate the membranes to design operating pressures. Putting a system on-line at design specifications too quickly could stress and irreversibly damage the membranes. During the break-in period the recovery rate is kept relatively low at 65% and rejection rates were held at 85-90%. Initially, effluent water quality for the first 50 hours of operation was approximately 8.5 micromhos conductivity and 320 ppb silica. After the initial break-in period, the recovery rate and solids rejection were adjusted to design criteria. Water quality increased to consistently maintain 4.2 micromhos conductivity and approximately 100 ppb silica. Base line data to monitor system performance is based on the data collected after the first 50 hours of operation.

In general, the operation of a reverse osmosis system is fairly simple, requiring only a brief daily inspection to check for damaged components and /or leaks, minor adjustments for flow and pressures to ensure that the system is operating according to the design parameters, and collection and analysis of data to determine the long term maintenance requirements. To collect and compile this data, inspection sheets were developed for the operating group. Appendix C.7 contains a copy of the actual inspection sheet. The basic inspections were performed daily to ensure proper operation while other activities were performed either semi-weekly, weekly or semi-monthly. Table 5 illustrates the inspection activities and the frequency in which they were performed. Once the operators were familiar with the new equipment, daily readings could be accomplished within 15 minutes. The more time consuming activities such as semi-monthly SDI testing required approximately 30 minutes to perform. Even these inspections did not prove to be labor intensive operations. It was decided to focus attention for this report on reverse osmosis skid number 2, because specifics of the system can be detailed without being redundant with results from skids 1. Both skids performed on par with each other.

Table 5 Inspection activity and frequency

Feed water conductivity	Semi-weekly
Feed water silica	Semi-weekly
Feedwater flow and accumulator	Daily
Feed water temperature	Semi-weekly
Feed water SDI	Semi-monthly
Feed water chlorine residual	Weekly
Feed water pH	Weekly
Filter differential pressure	Weekly
Filter effluent SDI	Semi-monthly
R-O unit in-service	Daily
Hours of operation	Daily
Product flow	Daily
Product conductivity	Daily
Product silica residual	Daily
Product totalizer	Daily
Product pH	Weekly
Feed pump pressure	Semi-weekly
First stage pressure	Semi-weekly
Product pressure	Semi-weekly
First stage product conductivity	Semi-weekly
Second stage pressure	Semi-weekly
Second stage inlet conductivity	Semi-weekly
Second stage outlet conductivity	Semi-weekly
Concentrate pressure	Semi-weekly
Concentrate flow	Daily
Concentrate pH	Weekly
Concentrate conductivity	Daily
Silica rejection	Daily
Normalized product flow	Semi-weekly
Recovery percent	Daily
Decarbonator influent CO_2	Weekly
Decarbonator effluent CO_2	Weekly
CO_2 reduction	Weekly
MUT in-service	Daily
MUT effluent conductivity	Daily
MUT effluent silica	Daily
MUT totalizer	Daily
Comments	Flow adjustment, leaks, etc.

6.2 Demineralizer Water Quality

Demineralizer effluent water quality exceeded expectations. An immediate observation was that the demineralizer effluent water quality was not dependent on flow. Historically, the plant, designed for 125 gpm flow, experienced variable water quality conditions when the demineralizer flow was increased. Water quality decreased as flow increased beyond 150 gpm. Even though water quality diminished with increased flow, it was still necessary to keep the demineralizers in-service to meet station water requirements. Figures 12 and 13 illustrate this condition.

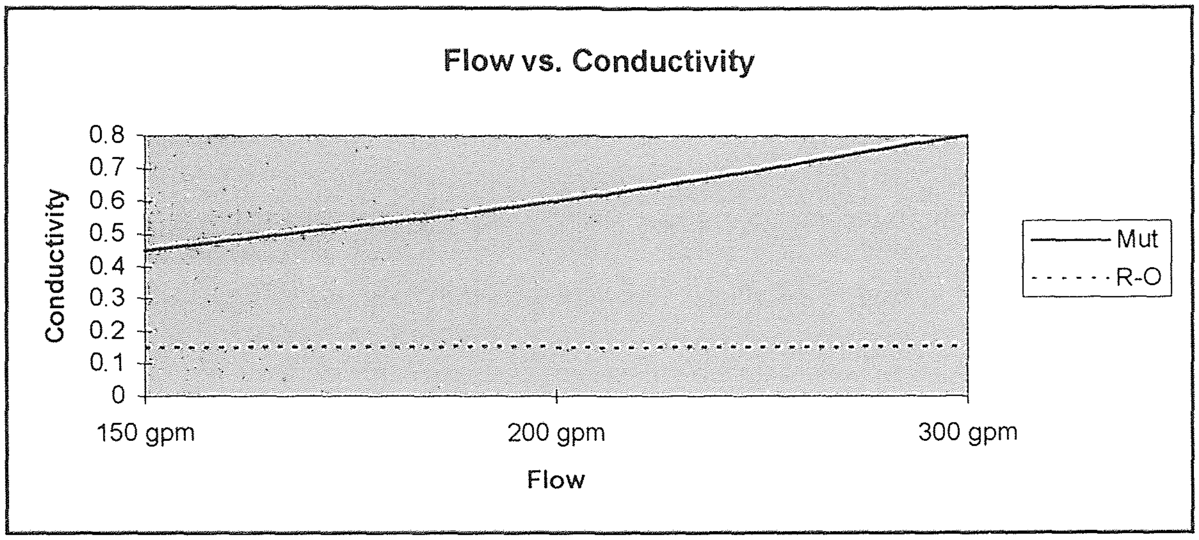

Figure 12 Demineralizer flow / conductivity relationship

Reduced water quality is due to the rapid movement of water through the demineralizer and the inability of the resins to kinetically remove the ionic species as efficiently at this high flow rate. As the two graphs show, water quality, using reverse osmosis, is not dependent on flow. Water quality was consistently good at either the low flow rate of 150 gpm, or at the high flow rate of 300 gpm being attributable to the low ionic loading at the demineralizer influent. Previous to the reverse osmosis system, an operator would have stayed on an overtime shift to monitor demineralizer performance, since at a high flow rate operation was close to our exceedance limit of 15 ppb silica or 1.0 micromhos conductivity, and there was concern that the demineralizer would prematurely break and contaminate the storage tank. Fortunately, this is no longer the case and the need for operator overtime to monitor the demineralizer has been eliminated.

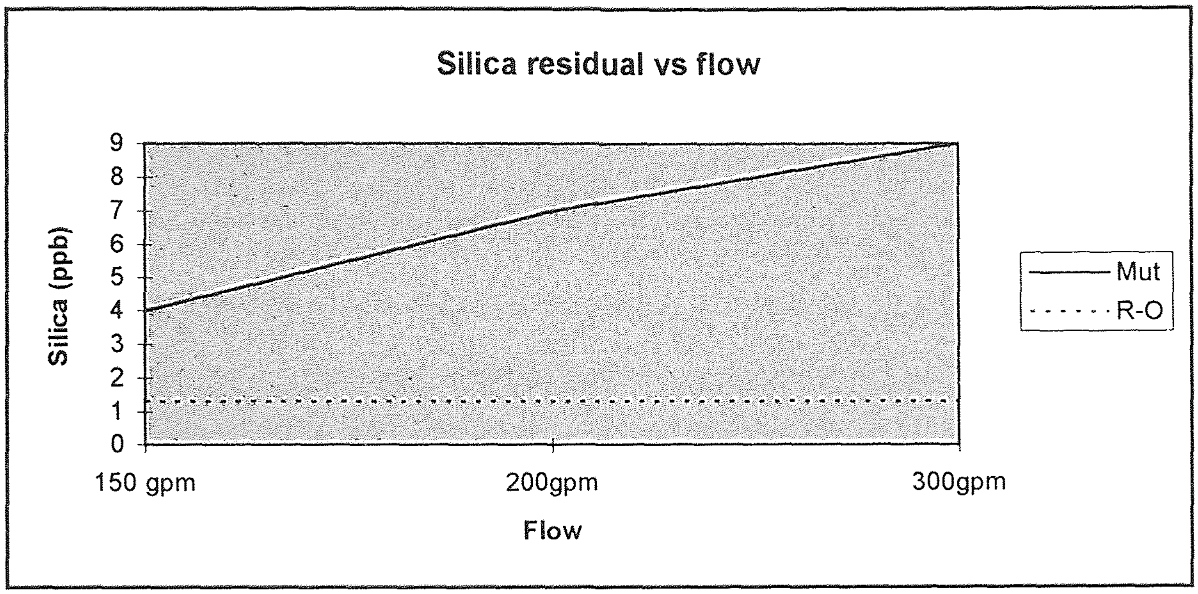

Figure 13 Demineralizer flow / silica relationship

On average, the water quality before the installation of the reverse osmosis system was approximately 0.4 - 0.8 micromhos conductivity, and 5-11 ppb silica. After the reverse osmosis system was installed, conductivity averaged 0.15 micromhos, with a low value of 0.08 micromhos, while silica averaged 1.5 ppb with a low 0.4 ppb silica. This increase in water quality was so dramatic, that our first reaction was to check our instrumentation, which was found to be accurate. This was a significant benefit to the station since we were no longer operating close to exceedance limits. Even though there were other problems with the reverse osmosis system, demineralizer effluent quality was never an issue. The reason for the higher effluent water quality is due to the lower ionic loading to the demineralizers. Table 6 shows the influent characteristic differences of the major ionic species.

Table 6 Demineralizer influent characteristics

Parameter	City water (mg/l)	R-O Permeate (mg/l)
Total dissolved solids	150	2.93
Total organic carbon	0.316	<0.1
Chloride	37.1	1.61
Alkalinity	35.2	Below detection
Calcium as $CaCO_3$	38.0	<0.02
Magnesium as $CaCO_3$	24.4	<0.004
Sulfate	15.4	Below detection
Hardness as $CaCO_3$	62.4	Below detection
Conductance	250 micromhos	4.3 micromhos
Silica	7.44	0.123
Potassium	1.02	<0.13
Sodium	21.5	0.741

6.3 Demineralizer Throughput

Demineralizer throughput dramatically increased from an average of 300,000 gallons to almost 7,000,000 gallons per service cycle, a 2300% increase. This was higher than the predicted 2000% increase, which was based on a 95 % rejection rate. Table 7 illustrates the demineralizer service cycles during the evaluation period.

Table 7 Demineralizer service cycles

Demineralizer	Date	Through-put (gallons)
21	5/9/94 - 6/12	6,747,220
22	6/12 - 7/12	6,686,300
21	7/12- 8/3	6,324,970
22	8/3 - 9/1	6,774,970
21	9/1 - 11/3	5,684,590
22	11/3 - 12/2	6,598,130
21	12/2 - 1/20/95	6,340,892

Total throughput for this evaluation period was 45,157,072 gallons with an average service cycle of 6,451,010 gallons. This provided a 2150% service cycle increase. Figure 14 graphically depicts the demineralizer service cycles during the evaluation period. There are two reasons for the slight inconsistency of service cycles. The first being a slight decrease when sulfuric acid injection was implemented to control calcium carbonate precipitation; even though the decarbonator was used during this period, it was unable to completely remove the excess carbon dioxide generated during the acidification process.

Consequently, it was left to the demineralizer anion resins to remove the carbon dioxide, which resulted in slight loss of anion capacity. Subsequently, service cycles were reduced by approximately 300,000 gallons, indicating a 4 - 6 % reduction. The second reason for reduction in throughput is the result of varying operator procedures for conducting a regeneration. Even though there are standard operating procedures for regenerating a demineralizer, no two operators perform an identical regeneration. This is particularly evident when a new operator performed the fifth regeneration which showed a large variation from the others. Nonetheless, even with inconsistencies and slightly reduced throughput, the demineralizer service cycle increased from one day to as much as four weeks before requiring regeneration. This not only reduced chemical and water usage, but it allowed the operator to perform other tasks besides daily regenerations. Some of these

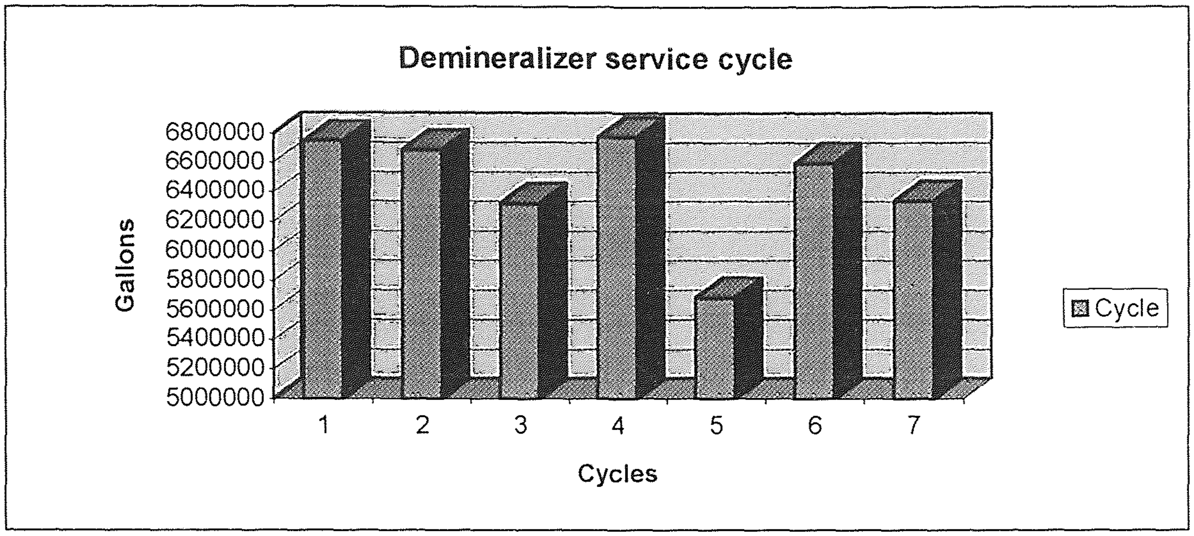

Figure 14 Demineralizer service cycle

tasks included plant chemistry, and general maintenance and repair work. Previously, these functions were performed by other plant personnel.

6.4 Filter Performance

This section will detail the performance of the multimedia and cartridge filters. As previously discussed the filters are essential for lowering the SDI below 3 to control membrane fouling and to comply with the manufacturer's warranty. Figure 15 graphically shows the SDI removal efficiency of the filters. As shown, city water or influent SDI reaches a maximum of 7 in mid-December with a low of 2.8 in May. Generally, the influent SDI was below 5 for most of the year. However, in late October the SDI rapidly started to trend upward and continued to rise until early January, when it began to recede. This has been attributed to a phenomena known as "lake turnover" which occurs in northern regions during the late fall and winter months. Since the station receives its water from the Boonton reservoir, we experienced this phenomena. During the summer months, lakes or reservoirs are strongly stratified with two distinct layers. The epilimnion represents the warmer upper layer, and the colder lower region is known as the hypolimnion. A vertical plane of maximum temperature difference, called the thermocline, is located between these two layers.

The reason for the vertical temperature differences is due to the low conductivity of heat and absorption in water. Therefore, only the top 10-12 feet of the water body is radiantly heated. During autumn, surface temperatures begin to fall and subsequently the thermocline penetrates deeper in to the reservoir; resulting in a turning of the lake sediment. This sediment contains appreciable amount of decaying organic matter that is entrained in the water phase. This entrained matter is colloidal by nature and does not readily settle. As such it remains in the water supply. The higher SDI values that we experienced were directly related to "lake turnover." When temperatures equalize, the mixing action ceases, and SDI values declined.

Only during this mixing action did our filter effluent SDI exceed 3 when a maximum value of 4.1was analyzed in mid-December. Generally, effluent SDI values were below 3 with a minimum value of 1.8 in May. Table 8 shows the SDI values used in figure 15 and the percent reduction as a result of the filters. As a general rule of thumb, the SDI will be reduced by 50% using multimedia and a minimum of 3 micron filters.

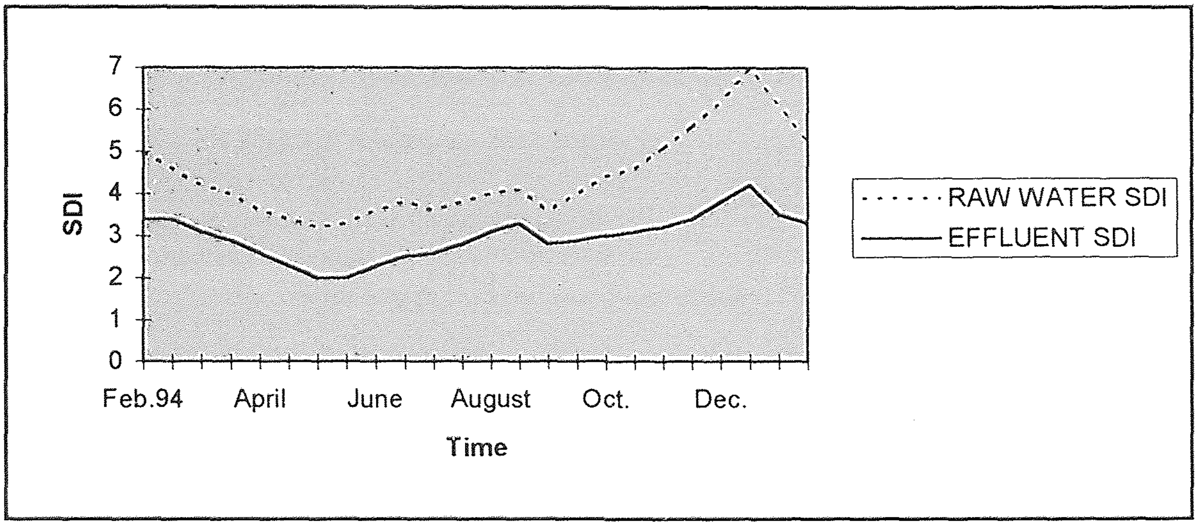

Figure 15 Influent / effluent SDI

We never attained 50%, but this was attributed to our relatively low influent SDI values. Salt and brackish waters, which have much higher SDI values, might attain this efficiency.

Table 8 SDI value

Date	Influent SDI	Effluent SDI	% Reduction
February 1994	5.0	3.4	32
	4.6	3.4	21
March	4.2	3.1	26
	4.0	2.9	28
April	3.6	2.6	28
	3.4	2.3	32
May	3.2	1.8	41
	3.3	2.0	39
June	3.6	2.3	36
	3.8	2.5	34
July	3.6	2.6	28
	3.8	2.8	26
August	4.0	3.1	22
	4.1	3.3	20
September	3.6	2.8	22
	4.0	2.9	28
October	4.4	3.0	32
	4.6	3.1	33
November	5.1	3.2	37
	5.6	3.4	39
December	6.2	3.8	39
	7.0	4.2	40
January 1995	6.1	3.5	43
	5.2	3.3	37

6.5 Permeate Quality

Initial product water quality averaged approximately 4.2 micromhos conductivity from an influent of 250 micromhos, providing better than a 98% rejection rate. During the course of the evaluation, the rejection rate was never lower than 96.9%. Table 9 provides a overview of the product water quality after the initial break-in period. The table also shows the city water or influent water quality and the percent rejection factor for each constituent of the water supply.

As noted in Chapter Two, reverse osmosis membranes will reject divalent ions more readily than monovalent ions. The monovalent ions in the product water (sodium, chloride, and potassium) constituted a majority of the mineral content, whereas, they account for only 40% of the city water. On the other hand, the divalent ions such as calcium and magnesium, which account for a large majority of the city water salts are practically non-existent in the product water. The divalent ions show a membrane rejection rate greater than 99.9%, whereas, the monovalent ions show a 95-96% rejection rate. Our analysis confirms that divalent ions are more readily rejected than monovalent ions. Reactive silica, which has a slight negative charge, was rejected at 98.3%. This was higher than the 86.1% rejection rate experienced with colloidal silica. The testing was performed by PSE&G's Maplewood Research and Testing Laboratory in accordance with the USEPA *Methods for Chemical Analysis of Water and Wastes*, USEPA *Test Methods for Evaluating Solid Waste Physical / Chemical Methods*, and *Standard Methods for the Examination of Water and Wastewaters, 18th edition*. Appendix C.1 details the methods.

Table 9 Water quality analysis

Parameter	City water	Product	Rejection %
Total Dissolved Solids	150 ppm	2.93 ppm	98
Conductance	250 micromhos	4.3 micromho	98.3
Chloride	37.1 ppm	1.61 ppm	95.6
Alkalinity	40.2 ppm	< MDL	N.A.
Calcium as $CaCO_3$	38.1 ppm	< 0.02 ppm	> 99.9
Magnesium	24.4 ppm	< 0.004 ppm	> 99.9
Sulfate	15.4 ppm	< MDL	N.A.
Reactive silica	7.44 ppm	0.123 ppm	98.3
Colloidal silica	1.3 ppm	0.18 ppm	86.1
Potassium	1.02 ppm	< 0.13 ppm	N.A.
Sodium	21.5 ppm	0.741 ppm	96.5
Manganese	0.008 ppm	< 0.003 ppm	N.A.
Total Organic Carbon	0.316 ppm	< 0.1 ppm	N.A.
Barium	< 0.1 ppm	< 0.1 ppm	N.A.
Iron	< 0.02 ppm	<0.02 ppm	N.A.
Copper	< 0.01 ppm	< 0.01 ppm	N.A.

MDL Method Detection Limit

N.A. Not Applicable. Constituents detected but not quantified since they are below MDL.

A mass balance was performed across the reverse osmosis stages to illustrate how the unit provides this quality of purified water and to demonstrate that each individual element has a higher rejection rate than the total unit. Each of the two stages will recover 50% of its influent and reject greater than 98.3% of its solids. Therefore, the second stage influent, which is the concentrate waste of the first, will have twice the solids content of the first stage. The final stage, which is the concentrate blowdown will have four times the solids content as the first stage. The first stage, with four modules will produce 100 gpm product water, while the second stage will produce 50 gpm. Figure 16 illustrates the flow conditions and water quality.

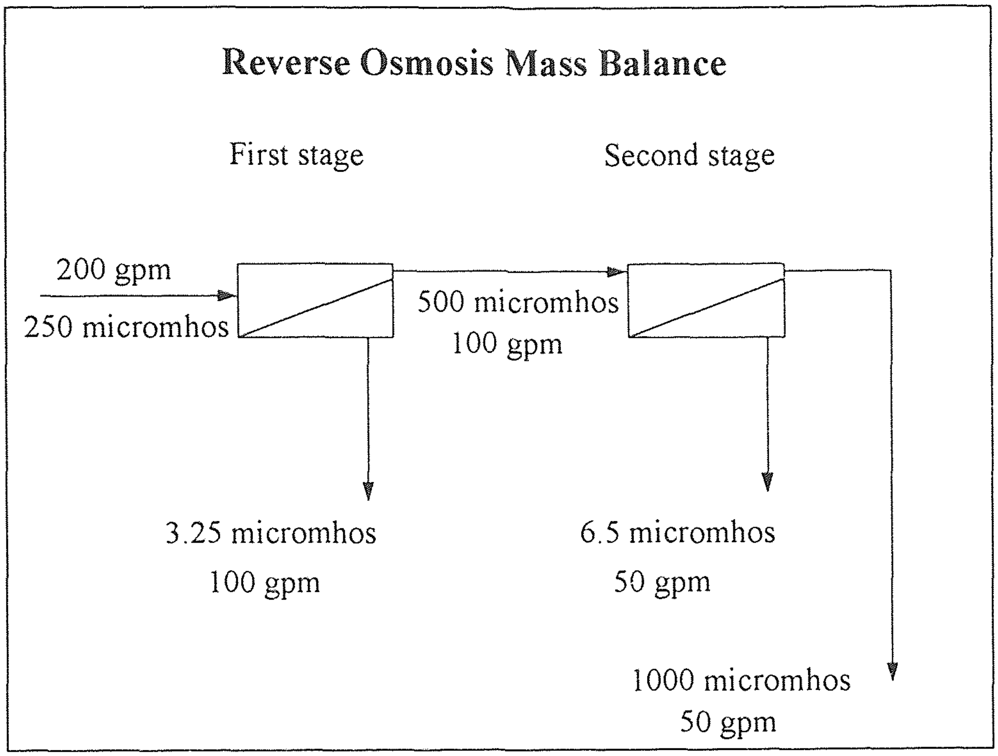

Figure 16 Reverse osmosis mass balance

In order to obtain a first stage product conductivity of 3.25 micromhos and 6.5 micromhos with the second stage as analytical data shows, the individual element rejection must be 98.7% to obtain an overall conductivity of 4.3 micromhos.

$$\frac{[(100\text{ gpm})\ (250 \times 1- 0.987)] + [(50\text{ gpm})\ (500 \times (1-0.987)]}{100\text{ gpm} + 50\text{ gpm}} = 4.33\text{ micromhos}$$

6.6 Performance Indicators

The proper analysis of reverse osmosis is essential for the successful operation of the system. The regular collection of appropriate data and its proper interpretation is essential to trouble free operation. To evaluate system performance it is necessary to compare permeate flow and salt rejection at the same conditions. In essence, it is necessary to convert the collected data obtained at actual conditions to a set of selected baseline conditions, thereby standardizing or normalizing the data. Once the appropriate system data is collected and normalized, a basis for identifying gradual or sudden performance changes is well established. The data collected is used to track three key performance elements: salt rejection, product flow rate, and feed water to concentrate pressure differential. Feed water and reject pressure differential values tend to vary as a function of both temperature and flow rate. Therefore, it is imperative to collect this data at baseline operating conditions to properly observe changes in temperature and feed water flow rate. This baseline data was collected after the initial membrane break-in period. Significant changes in the normalized data indicate that corrective action is required or will soon need attention.

Membrane flux is primarily dependent on transmembrane pressure differential and temperature. It is necessary to determine the effective operating transmembrane pressure and temperature to track system productivity. Reverse osmosis membranes flux rates are determined at 77 degrees F; a temperature swing either way will affect flow rate due to the increased or decreased water viscosity, which plays an important role in determining the flux rate and productivity rate. A temperature difference of 2 degrees F. can change product flux 2- 3%. The observed product flow rate can be normalized to the baseline operating conditions by multiplying the observed product flow rate by the ratio of the baseline temperature to actual operating temperature. The calculated product flow rate is further adjusted for varying pump pressures. Baseline operating conditions for the startup of the unit is 225 psig at the first membrane stage. This is not the maximum pressure output that the pump can develop but the manufacturer's recommended pressure for the system. The pump is capable of developing 330 psig at the first stage inlet. As the membranes become fouled and/or scaled, it will be necessary to adjust the pump pressure to maintain the specified product flow. Consequently, transmembrane pressure differential will increase accordingly. It is necessary to track this pressure differential and to normalize the data to determine when the membranes require cleaning. The observed product flow rate can be normalized to the baseline operating conditions by multiplying the observed product flow rate by the ratio of the baseline transmembrane pressure differential to the actual transmembrane pressure. Normalized product flow is calculated by using equation 8.

$$NPF = MPF \times TCF \times (225 / MFP) \tag{8}$$

where, NPF = normalized product flow

MPF = measured product flow

TCF = temperature correction factor

MFP = measured feed pressure

A decrease in normalized product flow of 15-20% indicates that the membranes require chemical cleaning. To continue beyond 20% of the normalized product flow rate is against the recommendations of the membrane manufacturer and can cause serious and irreparable damage to the membranes.

In addition to normalized flow, an increase in the pressure drop from the baseline conditions across individual stages is also an indication of membrane performance. This indicator can be used to troubleshoot fouling or scaling conditions and to isolate the cause. However, the pressure drop across a reverse osmosis system is not the most reliable method of determining when to clean, since the normalized flow loss is usually observed before reliable pressure increases are accurately detected.

Another indicator of system performance is an increase in the salt passage or conversely, a decrease in salt rejection. The increase in salt passage is usually an indication of a mechanical leak in the system such as a faulty O-ring seal, or a physical tear in the membrane. An increase in salt passage exceeding 5% will require internal inspection of the membranes to determine the reason for the high permeate solids. Percent salt reject and percent salt passage can be determined as follows:

$$\%\ Salt\ Rejection = [(Cf - Cp)/Cf] \times 100 \tag{9}$$

$$\%\ Salt\ Passage = [Cp/Cf] \times 100 \tag{10}$$

where, Cp = Permeate conductivity, micromhos

Cf = Feed water conductivity, micromhos

6.7 Reverse Osmosis System Operations

The following graphs will detail operating conditions and the problems encountered during the evaluation period. Due to the volume, it is not practical to graph all the data collected; therefore, data is graphed by operational hours with 100 hour increments. Appendix C.2 - C.6 presents actual data. The first graph is the recovery rate as shown in Figure 17.

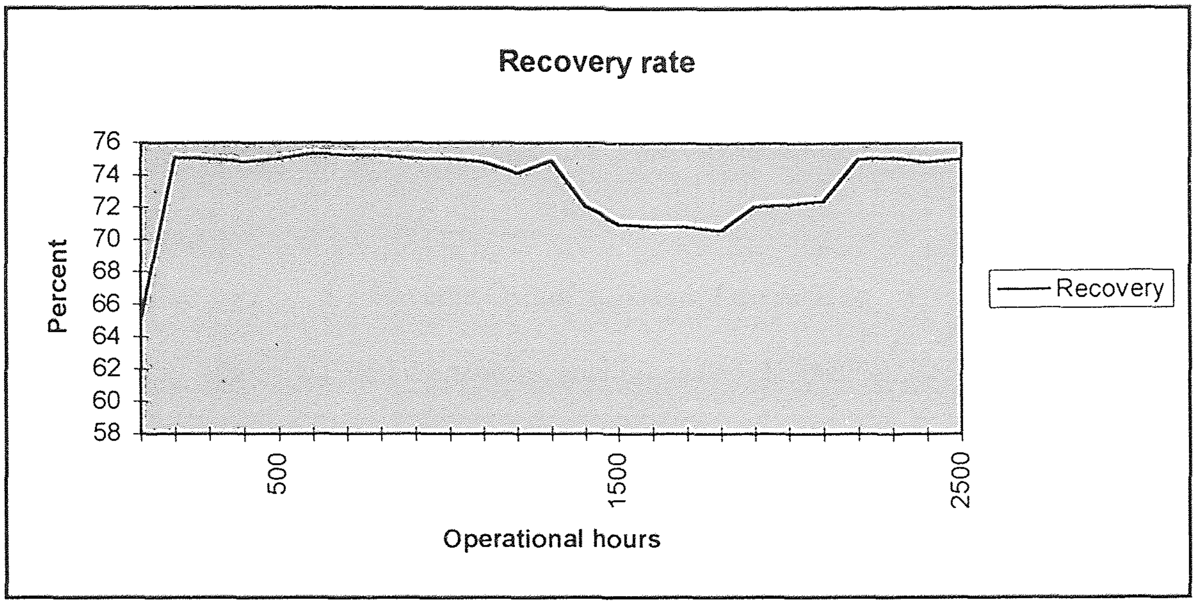

Figure 17 Recovery rate

The graph shows that after the initial break-in period, when the recovery was maintained at 65%, the rate was held consistently at approximately 75%. In the beginning of September we started to experience high pressure drops across the second stage. In an effort to prevent membrane scaling we lowered the recovery rate to 72% and finally 70 % by operational hour 1600. Lowering the recovery rate will reduce the concentration of solids at the membrane surface which reduces the potential for scaling. This was only a temporary fix to forestall irreparable damage until we could trouble shoot the problem and take corrective action. Once the problem was rectified, recovery rate was again held at 75%.

To address the problem with membrane scaling it is necessary to review the normalized product flow (NPF) as shown in figure 18.

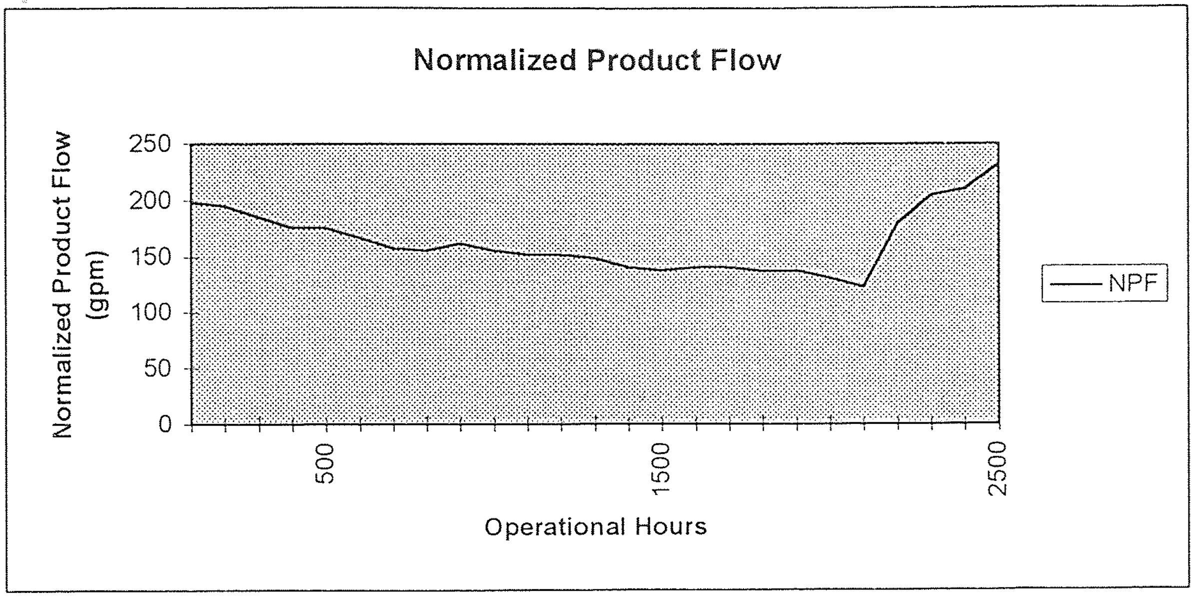

Figure 18 Normalized Product Flow

The graph shows that the normalized product flow stabilized at approximately 160 gpm at operational hour 1000. The decline in NPF was originally due to the higher water temperatures in the summer and not from differential pressure increases. However, observations from the graph, shows that at operating hour 1200 the NPF started to decline even though temperature was constant. This is a clear indication of membrane fouling or scaling. Remedial action is required whenever the NPF falls below 10-20% of baseline conditions. Considering 160 as baseline, then remedial action was required when the NPF reached 144-128 gpm. Before appropriate action can be taken, it is necessary to determine if fouling or scaling is causing the high pressure differentials. Figure 19 illustrates the pressure drops across each stage.

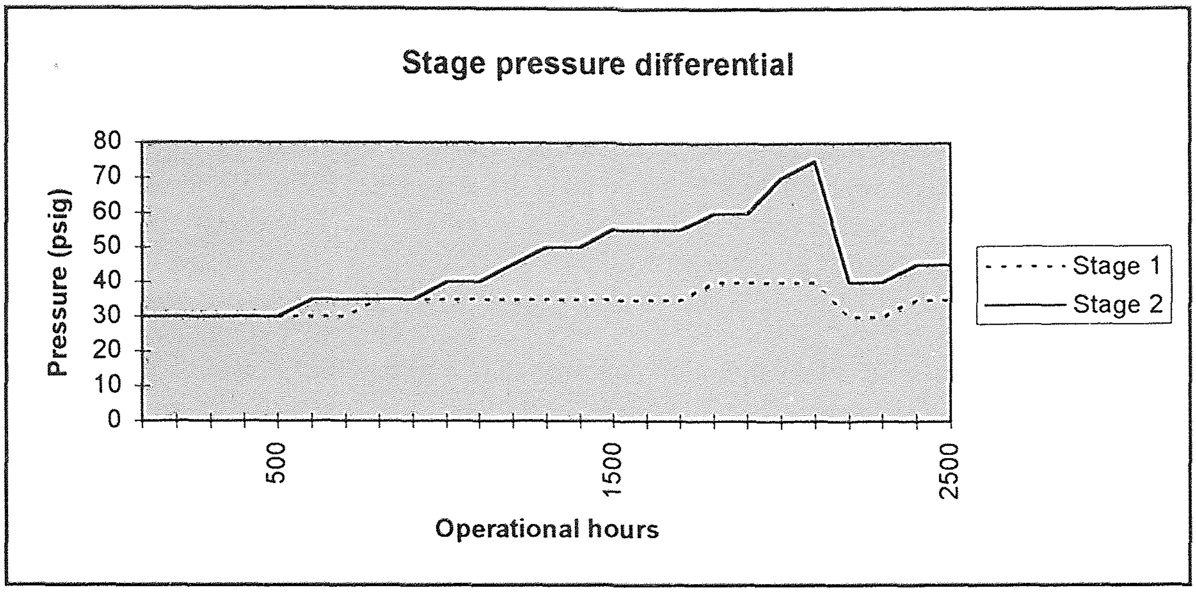

Figure 19 Individual stage pressure differential

This graph clearly shows that the pluggage causing the high pressure drop was in the second stage. Second stage pluggage is the result of scaling. Colloidal and suspended solids have a 99.9% removal efficiency in the first stage and would not be present to foul the second stage membranes. We felt that the scaling was due to calcium carbonate precipitation, and immediately implemented sulfuric acid injection. Acid was fed to the inlet of the cartridge filters to maintain a 5.5 feed water pH. The LSI was lowered to

-1.97, well within the calcium dissolving or corrosive range. The acid injection had a dual purpose: first, to control further precipitation, and second, to attempt to dissolve the precipitates and forestall removing the membranes for cleaning. Acid injection was initiated on September 18th, at approximately the 1800 operational hour, to maintain a 5.5 pH in the feed water. Two in-place acid cleanings were performed during this time period which involved removing the unit from service and isolating the system influent and effluent valves. Sulfuric acid was injected into the membrane modules and soaked for 3-5 hours with a pH 2 solution, after which, the modules were drained, flushed with permeate, and air sparged. Both of these cleanings failed to dissolve the precipitation and restore membrane performance.

It became immediately apparent to us at this time that acid injection should have been employed from the start-up of these units. Even though the Langelier Saturation Index calculation performed in Chapter 4 showed that the water was slightly corrosive at our operating conditions, it did not take into account the concentration polarization at the membrane surfaces, despite the natural water turbulence in the elements. Unfortunately, this was a time of peak water demand and the system could not be removed from service for the three days required for off-site membrane cleaning. It was decided to keep the system in service until station operating conditions would allow its removal. During that time, we lowered the recovery rate to 70% to reduce the concentration polarization factor, and increased the pump pressure to attain the required flow. Increasing pump pressure will increase flow, but there is also a chance of damaging the membranes. When membranes are heavily scaled, excessive pressure can irreversibly damage the membrane by a phenomena called *telescoping*. Telescoping occurs when the membrane becomes unraveled due to the axial force applied to it by the pump pressure. By the end of October, our NPF was greater than 19% of baseline conditions and the units could only produce a maximum of 135 gpm. The first stage inlet pressure at this time was 310 psig, indicating that there was little pump head remaining before plant productivity would significantly suffer. Fortunately, station operating conditions allowed us to remove the reverse osmosis units from service for cleaning during the first week of November.

Membrane cleaning was accomplished off-site at Polymetrics New Haven service center. This facility is equipped with a membrane cleaning skid that allows the membrane elements to be cleaned separately at operating flow rates. The membranes are cleaned separately by stages. The first stage was slightly fouled with organic and colloidal contamination. These elements were cleaned with a 0.1% solution of sodium hydroxide, ethylenediaminetetracetic acid and trisodium phosphate. The second stage, heavily fouled with calcium carbonate, required a 0.5% solution of hydrochloric acid to dissolve the calcium deposits, followed by the caustic solution as used in the first stage. Cleaning was performed at 120 degrees F and involved circulating the solutions for 1-2 hours followed by a 4-5 hour soak period. This procedure was repeated twice. The higher temperatures aids in dissolving mineral deposits and sloughing organic material from the membranes. Multiple circulating and soaking cycles are required in order to dissolve blockage, flush it out, dissolve more blockage, and so on. The membranes are placed six to a skid with parallel cleaning flow so that the contaminants removed from one do not re-enter into another element. After cleaning, the membranes were rinsed with demineralized water until the effluent was clear and the conductivity was within acceptable limits. It took three days to remove the membranes, clean them, and reinstall. During the cleaning process, Polymetrics supplied mobile demineralizers, at no charge, to meet station water requirements.

The system was back on-line by November 10th. As figure 19 indicates, pressure drops were within recommended limits and water production was back to design specifications of 150 gpm. The cleaning process was a success.

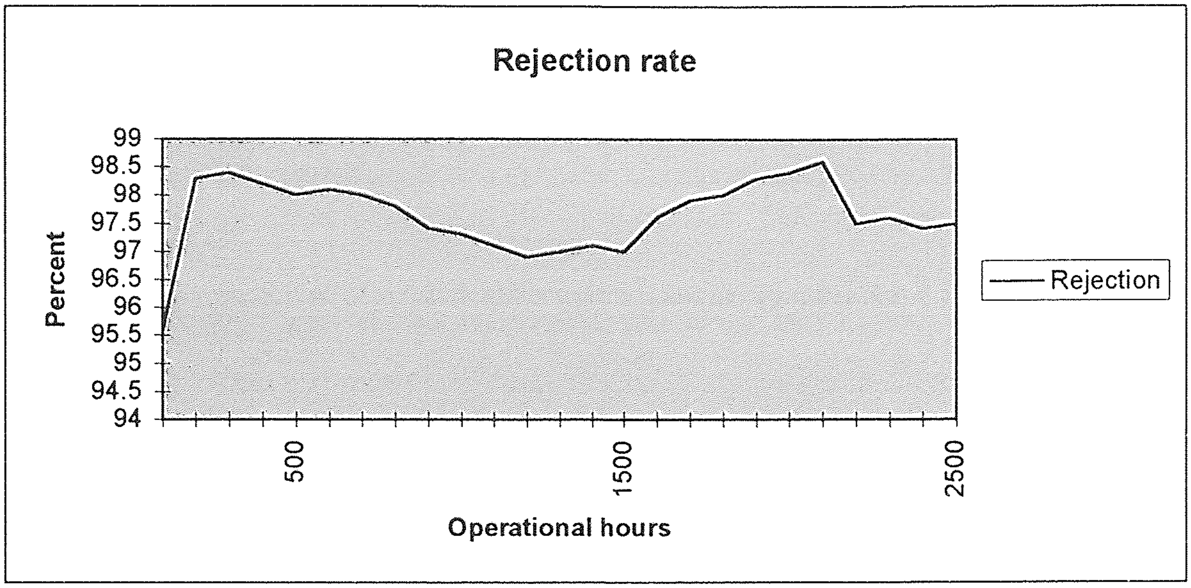

Figure 20 Rejection rate

An interesting observation was detected prior, during, and after the chemical cleaning. Whenever membranes become fouled or scaled, there is a greater potential for salt passage to increase or conversely, salt rejection to decrease. However, figure 20 shows that the opposite occurred; our rejection rate slightly increased when the membranes were scaled. Initially the rejection rate based on conductivity was 98.3%. Up to approximately 1200 operating hours, prior to scaling, the rejection rate slightly trended down until it reached a low of 96.9%. After that point, it steadily increased until it reached a maximum of 98.6% immediately prior to cleaning. This can be explained by the operating changes we made at this time and by a phenomena called the "ripening effect." Salt passage is dependent on the solute flux coefficient and the concentration gradient across the membrane. For our purposes, the solute flux coefficient remained constant during this period. When the recovery rate was lowered, the salt flux was decreased, since the concentration gradient was less, providing a lower conductivity permeate. The permeate was further diluted by increasing the pump pressure to maintain specified flow. Since solute flux is not dependent on pressure, the additional flow simply diluted the permeate.

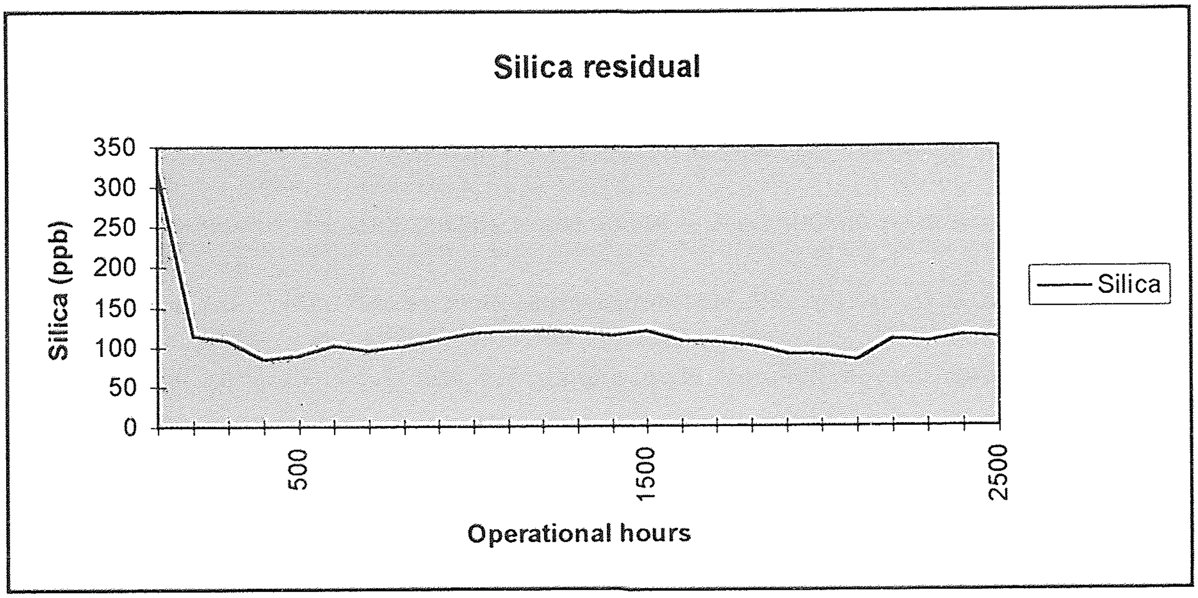

Figure 21 Permeate silica residual

Finally, as the membranes become fouled, their openings become smaller due to the build-up and assisted particle rejection. This is similar to media filters that become matted or ripened and entrain smaller suspended solids. After cleaning, the operating conditions were standardized and rejection resumed its normal 97.5 - 98% rate. Figures 21 and 22 show permeate silica and conductivity, during the evaluation period, respectively.

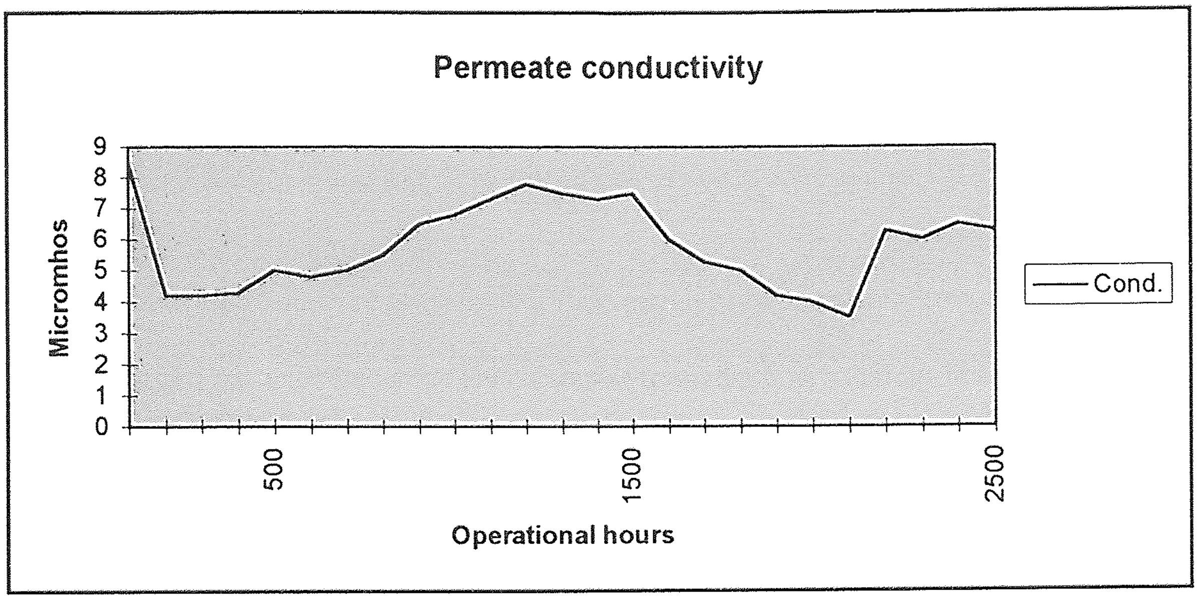

Figure 22 Permeate conductivity

Neither silica or conductivity were adversely affected by the scaling problems encountered. On the contrary, both values decrease during this time to provide a very high quality permeate. During the evaluation, neither silica or conductivity approached the minimum acceptable levels of 250 ppb silica and 10 micromhos conductivity; the one exception being the initial break-in period.

6.8 Decarbonator Performance

Permeate carbon dioxide levels remained consistently at approximately 5 mg/l. After sulfuric acid injection was implemented, the permeate carbon dioxide levels increased to an average of 16 mg/l, with a high value of 20 mg/l. The decarbonator reduced this value to approximately 7 mg/l, indicating a performance efficiency of 55%. This an acceptable removal efficiency considering the low carbon dioxide residuals present. Overall, the carbon dioxide levels increased by 2 mg/l which reduced demineralizer service cycle by 4%. However, if the decarbonator was not used, carbon dioxide levels would have increased by 11 mg/l, and demineralizer service cycles would have been drastically reduced. Since the service cycles are proportional to carbon dioxide loading, its projected that demineralizer through-put would have been reduced by 22% or 1,500,000 gallons. The decarbonator, being a low cost and maintenance item, is a sound investment for increasing demineralizer through-put.

During the evaluation, total chlorine residuals were maintained below 0.10 mg/l from the carbon filter effluent. An inspection of the membranes, during the cleaning process, showed no indication of chlorine oxidation.

CHAPTER 7

COST ANALYSIS AND BENEFITS

7.1 Cost Analysis

The costs associated with leasing the reverse osmosis system are offset by the savings realized by its implementation and use. The main cost savings realized will be the direct result of less demineralizer regenerations, including reduced usage of sodium hydroxide, sulfuric acid, demineralization, and city water. Additionally, we expect to experience longer life from the ion exchange resins since they will not be subjected to osmotic shock from frequent chemical regenerations; reduced labor requirements to man the demineralization plant during regular and overtime hours, and finally, elimination of the need to rent mobile demineralization units. The following paragraphs will detail each of these specific costs and explain how the benefits were derived. This data is based on five years of operational data from the demineralization plant and current costs, as of March, 1995.

In comparing the cost for previous demineralizer operations to the reverse osmosis/demineralizer operations, costs were based on dollars ($) /1000 gallons. These costs were further based on an annual requirement of 60,000,000 gallons of demineralized water and an average of 300 regenerations per year. Each cost is detailed as follows:

Carbon filter media replacement cost is approximately $15,000 every two years. Considering 60,000,000 gallons per year, our cost is $0.125 /1000 gallons. The carbon filters are backwashed daily with 4000 gallons of city water. At $1.74 /1000 gallons for city water, our cost is $0.04 /1000 gallons.

During the course of the evaluation, we backwashed the multmedia filters on a weekly basis. Each filter required 4320 gallons of city water to backwash and rinse. For all six filters, this requires 25,920 gallons weekly, at a $1.74 /1000 gallons, this resulted in a cost of $0.04 /1000 gallons.

The total cost to rent the reverse osmosis unit is $3.17 /1000 gallons. This cost includes the two reverse osmosis units, pretreatment system, media and cartridge replacement, and technical assistance, on an as-needed basis. The station, by contract, has no additional expenses associated with this lease.

Caustic requirement for each demineralizer regeneration is 300 gallons. The previous average was 300 regenerations annually, or 90,000 gallons of caustic soda per year. At $1.05 per gallon of 50% sodium hydroxide, our cost was $1.58 /1000 gallons. With the new system, we project 10 regenerations per year, requiring 3,000 gallons of 50% sodium hydroxide. Our cost will be $0.05 /1000 gallons.

Acid requirement for each demineralizer regeneration is 150 gallons. Based on 300 regenerations annually, or 45,000 gallons of acid per year. At $0.53 per gallon of Baume 66 sulfuric acid, our cost was $0.39 /1000 gallons. With the new system, we project 10 regenerations per year, requiring 1500 gallons of Baume 66 sulfuric acid. Our cost will be $ 0.01 /1000 gallons.

The demineralizer requires 60,000 gallons of demineralized water per regeneration. At 300 regenerations per year, the water requirement for regeneration was 18,000,000 gallons annually. This meant over 25 % of our water production went into regeneration and was not available for boiler make-up. Basing the cost for this demineralized water at a modest $7.00 /1000 gallons our cost was $2.10 /1000 gallons. Using these numbers, based on only 10 regenerations per year, our cost is $0.07 /1000 gallons.

City water is used to backwash and separate the resins prior to demineralizer chemical regeneration. Approximately 6000 gallons of city water is required for this purpose. Considering $1.74 /1000 gallons, our cost was $0.05 /1000 gallons. Considering only 10 regenerations per year, our cost is less than $ 0.01 /1000 gallons.

The demineralizer resins were routinely replaced every four years, due to breakdown of the resin and loss of exchange capacity from chemical shock. Graver Chemical, our resin supplier, informed us that we can conservatively expect a 50% increase in resin life as a result of less frequent regenerations. At a cost of $90,000.00 per demineralizer, previous resin replacement cost was $0.666 /1000 gallons. Our new cost will be $0.444 /1000 gallons.

During the previous five years, the plant had allocated approximately $70,000.00 per year for renting mobile demineralizer units to supplement plant production. During the evaluation period, we did not require rental equipment to supplement our needs and do not feel that we will need to in the future. Our previous cost for this service was $1.16 /1000 gallons. This is a total cost savings.

Considering five year's data, the plant required approximately 1100 hours annually of overtime labor requirements to monitor demineralizer performance, or to perform regenerations. At a cost of $30.00 hourly, the annual expense was $33,000.00, or $0.55 / 1000 gallons. During the evaluation period, no overtime was required, nor do we feel that overtime will be required in the future. This is also a total cost savings.

The city water requirements for the reverse osmosis system would be expected to be higher than the demineralizers because of the 25% reject stream. However, this is not the case. We estimate that only 45,000,000 gallons per year of reverse osmosis product water is required for plant needs due to less regenerations. It requires approximately 60,000,000 gallons of feed water, based on a 25% reject stream to produce 45,000,000 gallons of product. Historically, the plant has produced 60,000,000 gallons annually. Therefore, with either system 60,000,000 gallons of city water is a requirement. The cost for city water is $1.74 /1000 gallons.

One of the drawbacks of a reverse osmosis system is the power requirements for pump operations, however, the cost to the station to use its own electric power is only $0.02 per kilowatt hour. Power cost requirement is based on the following formula.

$$\text{Cost/1000 gallons} = \frac{(0.0001455) \times \text{pressure (psig)}}{\text{efficiency}} \qquad (11)$$

where, kilowatt-hour is $0.02
pressure = 225 psig
efficiency = 80% (pump and motor)

Cost = $0.04 /1000 gallons

Table 8 summarizes the cost savings, showing the demineralizer system, the demineralizer and reverse osmosis system, and the percentage of cost reduction of each system. Briefly, the table will illustrate that the savings are significant. This conservative estimate shows a savings of $2.66 /1000 gallons, (based on an annual usage of 60,000,000 gallons) or $159,600.00. The cost savings are not limited to those illustrated in the following chart, when considering that 15,000,000 gallons are no longer required for regenerations. The plant will require only 45,000,000 gallons annually.

Table 10 Cost analysis

Item	Demineralizer	Reverse osmosis Demineralizer	Reduction
Carbon filter media	0.125	0.125	0
Carbon filter backwash	0.04	0.04	0
Multi-media backwash	0	0.04	Increase
Reverse osmosis lease	0	3.17	Increase
Caustic soda	1.58	0.05	96.8
Sulfuric acid	0.39	0.01	97.4
Demineralized water	2.10	0.07	96.6
City water, regeneration	0.05	0.001	98
Ion exchange resins	0.666	0.444	50
Demineralizer rentals	1.16	0	100
Overtime labor	0.55	0	100
City water, feed water	1.74	1.74	0
Power requirements	0	0.04	Increase
TOTAL COST	8.40	5.74	2.66

Multiplying 45,000,000 gallons by $5.74 /1000 gallons yields an operating cost of $258,300.00. Considering our previous operating cost of $504,000.00, this provides an annual saving of ***$245,700.00***. The station has a five year contract to lease the reverse osmosis system and will realize a ***$1,228,500.00 cost*** saving over that period.

7.2 Ancillary Benefits

In addition to the immediate cost savings, we have been able to reduce chemical discharge to the waste treatment plant. The station discharges its process water into the Hackensack River, regulated under NJPDES. By eliminating 290 regenerations, we eliminate neutralizing and discharging 87,000 gallons caustic soda, and 43,500 gallons sulfuric acid from the waste treatment plant outfalls. This is an environmentally forward attitude, and reflects the Department of Environmental Protection's direction for pollution prevention.

There is always pressure on a reverse osmosis system to utilize the waste stream for beneficial use. A method was studied and implemented involving utilization of this waste stream to reduce nitrogen oxide emissions. Number 1 unit, which burns gas or oil, needed to reduce its nitrogen oxide emissions to remain in compliance with Phase II of the Clean Air Act Amendments. Water injection into the burners lowers flame temperature, which, in turn, lowers nitrogen oxide emissions. Testing done at the station showed a 30% nitrogen oxide reduction with city water injection. We proposed using the reverse osmosis waste and filter backwash streams as an alternate. After analysis confirmed that the waste stream did not contain any materials that would damage the burners and boiler tubes, the idea was approved by the corporate engineering department as a viable source and adopted into their design. The water injection system would require approximately 250 gpm at full unit load. The reverse osmosis system could supply, at most, 100 gpm, therefore, this source can only supplement water requirements; still this is a significant savings to the station. Not all of the reverse osmosis waste stream will be available for water injection since No.1 unit does not operate at 100% capacity, but projections indicate that we can recoup 75% of the waste stream. This translates into additional savings of ***$19,575.00*** per year, and further reduces the hydraulic loading to the waste treatment plant by ***11,250,000*** gallons per year.

While the demineralization plant was performing frequent regenerations, plans were initiated to automate the regeneration system which involved installing new valves and electric actuators to interface with a programmable logic controller and computer. The expectation was that an automated regeneration system would standardize procedures, provide consistent service cycles, and reduce labor requirements. Expected cost for this project was approximately **$120,000.00**. The implementation of the reverse osmosis system, and the need to regenerate only 10 times annually made this automation project unnecessary.

Colloidal silica removal is an important station consideration. Ion exchange equipment can optimally remove only 5% colloidal silica. This removal is accomplished more by filtration than ion exchange, since the silica is not in its ionic form. Our analytical results showed that the reverse osmosis unit rejected greater than 86% of the influent colloidal silica. Both soluble and colloidal silica vaporize above 450 psig in the boiler and condense downstream in the turbine. The problems occur when the silica precipitates on the turbine blades, causing vibrational disturbances, blade erosion, and loss of efficiency. Fortunately, not all colloidal silica will precipitate in the turbine due to numerous variables such as: pH, steam velocity, temperature, and silica concentration, all of which control deposition. The 86% rejection rate, which is standard for reverse osmosis, is certainly a benefit to the station. It would be difficult to quantify the total amount of silica that will no longer precipitate or the cost savings associated with it, however this removal will undoubtedly increase turbine efficiency and reduce the frequency for turbine overhauls.

Finally, the operations of the demineralization plant have been considerably more reliable. This allows the plant to better support unit operations and lowers our product cost.

CHAPTER 8

FINAL ANALYSIS

8.1 Recommendations

Our experience with operating the reverse osmosis system has taught us some valuable lessons. Foremost, it is not prudent to base acid injection requirements solely on the Langelier Saturation Index, and turbulent conditions within the element modules. Our experience shows that scaling will occur by concentration polarization at the membrane surface even with a slightly corrosive LSI value. Researching the LSI and concentration polarization relationship indicates that a conservative design would maintain an LSI at -1.5 to -2.0. This provides an adequate safety margin, and more than compensates for the concentration polarization. In our case, this meant maintaining the feed water pH at approximately 5.5. This was accomplished with a minimal amount of sulfuric acid. Our cost for the acid was less than $.01 /1000 gallons; well worth the investment. After the membrane cleaning in November, and with continued acid feed, we have not experienced the high pressure drops in the second stage, indicating scale precipitation is under control.

It was decided to clean the membranes when the normalized product flow decreased by 10% from baseline conditions. We felt that waiting until normalized product flow dropped 15-20% was too long and could be detrimental to the membranes. In our case, this will mean semi-annual cleaning. The membranes were cleaned in mid-April, 1995, approximately six months after the initial cleaning, when the normalized product flow decreased by 11% from baseline conditions. Our plan is to operate through the peak electric demand months of May through October by starting off with clean membranes, and removing the membranes for cleaning immediately after that period.

In retrospect, it was prudent to lease the reverse osmosis system when considering the problem the station had with membrane scaling. Membrane cleaning is a costly venture, and should be left to professionals. It was comforting to rely on the vendor's experience and expertise, knowing that they were responsible for the membranes. Because of the size of our system, even small trial and error mistakes could be expensive.

8.2 Conclusions

The implementation of the reverse osmosis process was an unqualified success! The three critical objectives: plant reliability, cost reduction, and improved water quality were achieved far beyond expectations. In fact, due to the positive experience realized at Hudson Generating Station, other stations in PSE&G's fossil fuel department are in various stages of implementing their own reverse osmosis systems.

The demineralization plant is a functioning and reliable component of the station. Since implementation, the plant has not required mobile demineralizers to supplement water needs. Chemical deliveries have been reduced from one to two per week to less than one delivery every four months. This is a significant safety benefit for plant personnel who receive chemical deliveries, as well as an environmental benefit resulting from the reduced potential for spillage.

Plant operators are no longer performing daily regenerations. During the evaluation period, May, 1994 through January, 1995, operators performed only seven demineralizer regenerations. Previously, this many regenerations would have been performed in just one week during peak operating conditions. These operators are now performing routine maintenance activities, plant chemistry, and other station duties. Off-shift labor requirements are no longer necessary, and the midnight shift demineralizer schedule has been canceled. The infrequent regenerations, and plant automation have completely eliminated this financial burden.

Finally, the reverse osmosis system eliminated the crisis management mindset of the plant and provided us the luxury of reliability.

Even though cost benefits have been discussed, the savings realized are significant. The station will realize a greater than $250,000 annual cost reduction. This cut the plant operation and maintenance budget in half. Any user of a reverse osmosis system will encounter the dilemma of utilizing the waste stream. In many applications, it is simply discharged as waste, however, the station had a water requirement that the reverse osmosis concentrate stream could fulfill. Approximately 75% of the waste stream is captured for beneficial use.

Water quality is vastly improved. The quality of water produced from the demineralizer is almost of sufficient quality to be used directly into the units. This has increased the life of the condensate polishers and reduced their regeneration frequency. During the evaluation period, the demineralizer storage tanks were never contaminated. Previously, premature break-thorough of the demineralizers resulted in high silica and conductivity levels in the storage tanks. Since the implementation of the reverse osmosis system, silica levels never exceeded 5 ppb, and conductivity was maintained below 0.5 micromhos.

In the final analysis, reverse osmosis is an innovative technology that alleviated the water quality and reliability difficulties that were occurring at the Hudson Generating Station.

APPENDIX A

DESIGN CALCULATIONS

A.1 City Water Osmotic Pressure Calculation

Parameter	City water (mg/l)	Molecular Weight	Molality (moles /liter)
Chloride	37.1	35.5	0.0010
Alkalinity as HCO_3	35.2	61	0.00058
Calcium	38	40	0.00095
Magnesium	24.4	24.3	0.0010
Sulfate	15.4	96	0.0002
Silica as SiO_2	7.44	60	0.0001
Sodium	21.5	23	0.00093
Potassium	1.02	39.1	< 1.0 x E-6
Barium	< 0.10	137.3	< 1.0 x E-6
Copper	< 0.010	63.5	< 1.0 x E-6
Iron	< 0.020	55.8	< 1.0 x E-6
Manganese	0.008	55	< 1.0 x E-6
Nickel	< 0.30	58.7	< 1.0 x E-6
Zinc	< 0.005	65.4	< 1.0 x E-6
Phosphorus	0.015	31	< 1.0 x E-6
Total Molality			0.00476

Osmotic Pressure = 1.12 (temperature degrees Kelvin) (total molality)

Design temperature = 25 degrees C + 273 = 298 deg.K

Osmotic Pressure = (1.12) x (298) x (0.00476) = 1.59 psi

This calculation shows that the rule of thumb of 1 psi for 100 ppm Total Dissolved Solids is accurate and reliable.

A.2 Concentration / Recovery Rate Relationship

The concentration of feed water solid in the waste stream based on recovery rate is determined by the following formula:

$$C_f = 1 / (1 - R)$$

where, C_f = Concentration factor

R = Recovery rate

Concentration factor	Recovery rate (percent)
2	50
2.5	60
3.3	70
4	75
5	80
6.7	85
10	90

A.3 Computer Generated Design Data

The software will prompt the user for specific information as shown below. The user enters this data (shown in italics) and the design data is generated. The following is the design information used at Hudson Generating Station, courtesy of the Polymetrics Corporation.

Input:

Feed:	*200gpm*
Recovery:	*75%*
Temperature:	*10 C*
Raw water:	*Raw water data entered*
Stages:	*2*
Membrane type:	*BW8040*
Water source	*Surface*
SDI	*3-5*

Output:

Stages	1	2
Number of modules	4	2
Number of elements / module	7	7
Total elements	28	14

Fouling factor:	0.85
Permeate flux (GFD)	15.6
Osmotic pressure (feed)	1.4 psi
Osmotic pressure (reject)	5.6 psi
Average pressure	3.5 psi
Feed pressure	260 psi
Reject pressure	185 psi
Average pressure	230 psi
Permeate flow	150 gpm

Stage	Element	Permeate (GPD)	TDS (mg/l)	Feed flow (gpm)	Feed TDS (mg/l)
1	1	5722	2	50	162
	2	5586	2	46	195
	3	5468	2	42.1	212
	4	5367	3	38.3	233
	5	5281	3	34.6	258
	6	5208	3	31	288
	7	5146	4	27.3	326
2	1	4953	4	47.5	374
	2	4824	5	44.1	403
	3	4711	5	40.7	436
	4	4611	6	37.5	473
	5	4523	6	34.4	517
	6	4446	7	31.1	569
	7	4378	8	28	630

Stage	Total	1	2
Reject (gpm)		95.1	50
Reject (mg/l)		374	706
Permeate (GPD)	216,003	151,110	64,893
Permeate (mg/l)	4	3	6

Permeate ion (mg/l)	Total	Stage 1	Stage 2
Calcium	0.3	0.2	0.5
Alkalinity	0.4	0.3	0.7
Chloride	0.8	0.6	1.3
Magnesium	0.1	0.1	0.2
Sodium	0.9	0.6	1.7
Sulfate	0.9	0.6	1.4
Silica	0.2	0.1	0.2
Potassium	0.2	0.1	0.3

Reject ion (mg/l)	Feed	Stage 1	Stage 2
Calcium	39.1	86.7	182.5
Alkalinity	42.4	91.3	176.1
Chloride	39	81.4	153.5

Magnesium	17	35.7	67.5
Sodium	22.1	46.1	86.8
Sulfate	16.2	34.1	69.6
Silica	6.8	13.5	27.4
Potassium	0.9	1.8	3.7

	Feed	Reject
pH	7.2	8.1
Ionic strength (molal)	0.0045	0.015
Langelier Saturation Index	-2.13	-0.36

APPENDIX B

SCALE CONTROL

B.1 Concentration Factors

Parameter	City water (ppm)	Analytical Concentrate (ppm)	Concentration factor
Total dissolved solids	150	580	3.86
Total organic carbon	0.316	0.637	2.02
Total phosphorus	0.015	0.058	3.87
Chloride	37.1	174	4.69
Alkalinity	35.2	152	4.32
Calcium as $CaCO_3$	38	182	4.79
Magnesium as $CaCO_3$	24.4	90	3.69
Sulfate	15.4	66.2	4.3
Conductance	250	1040	4.16
Silica	7.44	31.7	4.26
Barium	< 0.01	0.01	Not quantified
Strontium	< 0.02	< 0.02	Not quantified
Copper	< 0.01	0.024	Not quantified
Iron	< 0.02	0.041	Not quantified
Potassium	1.02	4.71	4.62
Sodium	21.5	92.2	4.29
Manganese	0.008	0.031	3.88
Total suspended solids	< 4	< 4	Not quantified

Note: Total organic carbon only doubled its original value. This is due to the first stage removing 99.9% of the colloidal organic matter in the feed water, leaving none to be concentrated in the second stage.

B.2 Langelier Saturation Index

Based on the following raw water data:

Calcium	160 mg/l
Alkalinity	180 mg/l
TDS	600 mg/l
Temperature	70 degrees F

Then pH saturation = 7.47, and LSI = pH (water) - pH (saturation)

pH	LSI
5.0	-2.47
5.5	-1.97
6.0	-1.47
6.5	-0.97
7.0	-0.47
7.5	+0.33
8.0	+0.53

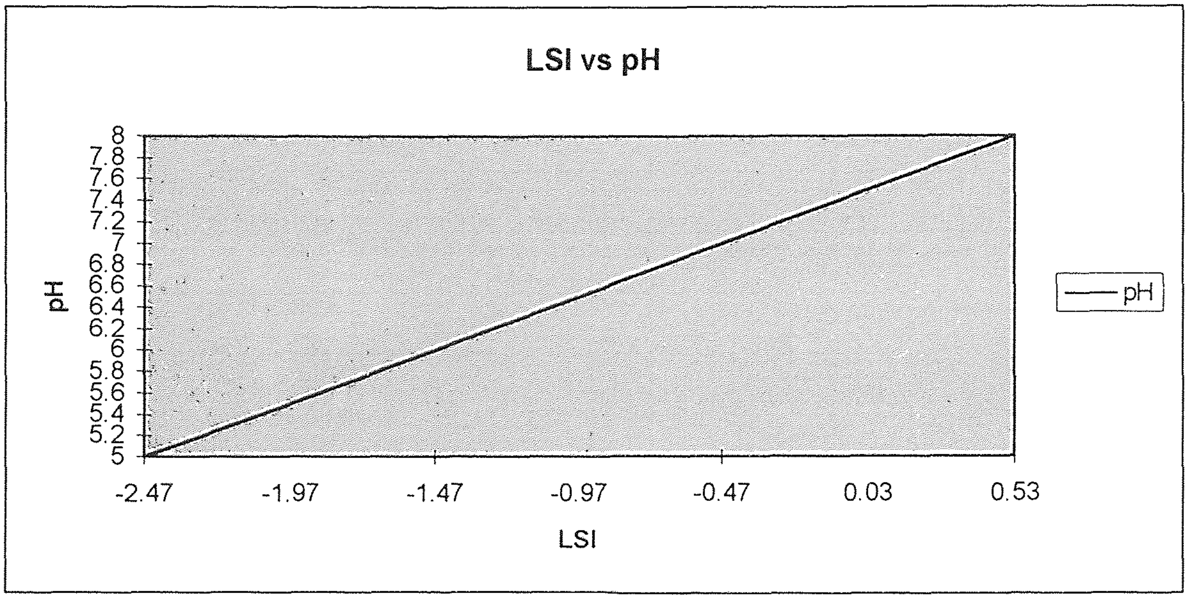

Figure 24 Langelier Saturation Index vs pH

APPENDIX C

EXPERIMENTAL DATA

C.1 Summary of Methodology

Parameter	Method Detection Limit (mg/L)	Method
Barium	0.100	USEPA 208.1
Calcium	0.020	USEPA 215.2
Copper	0.010	USEPA 220.1
Iron	0.020	USEPA 236.1
Potassium	0.130	USEPA 258.1
Magnesium	0.004	USEPA 242.1
Manganese	0.003	USEPA 243.1
Sodium	0.020	USEPA 273.1
Nickel	0.030	USEPA 249.1
Strontium	0.020	APHA 4500-Sr
Zinc	0.005	USEPA 289.1
Total Dissolved Solids	1.0	USEPA 160.1
Total Phosphate	0.01	APHA 4500-P
Alkalinity	10	USEPA 310.1
Chloride	0.1	USEPA 325.3
Hardness, $CaCO_3$	10	USEPA 130.2
Sulfate	10	USEPA 375.3
Conductance	1% of value	USEPA 120.1
Silica	0.010	APHA 4500-Si-E

C.2 Reverse Osmosis Permeate Data

Date	Operational hours	Silica (ppb)	Conductivity (micromhos)
	< 50	320	8.5
May 20	100	115	4.2
27	200	108	4.2
June 13	300	85	4.3
26	400	90	5.0
July 1	500	102	4.9
9	600	96	4.8
14	700	101	5.0
22	800	110	5.5
August 2	900	118	6.5
10	1000	117	6.8
15	1100	120	7.3
20	1200	121	7.8
24	1300	119	7.5
September 1	1400	115	7.3
4	1500	120	7.5
12	1600	108	6.0
19	1700	106	5.3
24	1800	101	5.0
October 7	1900	92	4.2
16	2000	90	4.0
28	2100	83	3.5
November 14	2200	110	6.3
December 2	2300	106	6.0
27	2400	115	6.5
January 16	2500	112	6.3

C.3 Recovery Rate and Rejection Rate

Date	Operational hours	Recovery Rate (Percent)	Rejection Rate (Percent)
	< 50	65	95.6
May 20	100	75.5	98.3
27	200	75	98.4
June 13	300	75.1	98.2
26	400	74.8	98
July 1	500	75	98.1
9	600	75	98
14	700	75.3	97.8
22	800	75.2	97.4
August 2	900	75.1	97.3
10	1000	75	97.1
15	1100	74.8	96.9
20	1200	74.1	97
24	1300	74.9	97.1
September 1	1400	72.1	97
4	1500	71	97
12	1600	70.8	97.6
19	1700	70.8	97.9
24	1800	70.6	98
October 7	1900	72.1	98.3
16	2000	72	98.4
28	2100	71.9	98.6
November 14	2200	75.1	97.5
December 2	2300	75	97.6
27	2400	74.8	97.4
January 16	2500	75	97.5

C.4 Normalized Product Flow Data

Date	Operational Hours	Actual Flow (gpm)	Temperature Degrees F	Feed Pressure (psi)	Normalized Product Flow
	< 50	150	60	200	198
May 20	100	150	60	225	195
27	200	150	64	225	185
June 13	300	150	66	230	176
26	400	150	66	230	176
July 1	500	150	69	230	169
9	600	150	72	230	160
14	700	150	73	230	158
22	800	150	73	230	158
August 2	900	150	72	235	155
10	1000	150	72	240	152
15	1100	150	72	240	152
20	1200	145	72	245	149
24	1300	145	73	250	139
September 1	1400	145	72	250	140
4	1500	145	72	255	138
12	1600	145	71	255	140
19	1700	145	71	255	140
24	1800	145	70	265	138
October 7	1900	140	66	270	138
16	2000	135	65	280	131
28	2100	135	63	310	123
November 14	2200	150	62	240	180
December 2	2300	150	54	240	205
27	2400	150	50	250	210
January 16	2500	150	44	250	232

C.5 Temperature Correction Factor (TCF)

The membrane's permeate capacity is designed at 25 degrees C or 77 degrees F. To correct for temperature variations the following formula was used:

TCF = 1.03 power (25 - T)

where, TCF = Temperature Correction Factor

T = Temperature, degrees C

Degrees F	Degrees C	TCF		Degrees F	Degrees C	TCF
34	1.1	2.03		64	17.7	1.23
36	2.2	1.96		66	18.9	1.19
38	3.3	1.92		68	20	1.16
40	4.4	1.85		70	21.1	1.12
42	5.6	1.79		72	22.2	1.08
44	6.7	1.72		74	23.3	1.05
46	7.8	1.67		76	24.4	1.02
48	8.9	1.61		77	25	1.00
50	10	1.56		78	25.5	0.98
52	11.1	1.51		80	26.6	0.95
54	12.2	1.47		82	27.7	0.92
56	13.3	1.40		84	28.9	0.89
58	14.4	1.37		86	30	0.86
60	15.5	1.32		88	31.1	0.83
62	16.6	1.28		90	32.2	0.81

C.6 Reverse Osmosis Stage Pressure Drop Data

Date	Operational Hours	First Stage (psi)	Second Stage (psi)
	< 50	25	25
May 20	100	30	30
27	200	30	30
June 13	300	30	30
26	400	30	30
July 1	500	30	30
9	600	30	35
14	700	30	35
22	800	30	35
August 2	900	35	35
10	1000	35	40
15	1100	35	40
20	1200	35	45
24	1300	35	50
September 1	1400	35	50
4	1500	35	55
12	1600	35	55
19	1700	35	55
24	1800	40	60
October 7	1900	40	60
16	2000	40	70
28	2100	40	75
November 14	2200	30	40
December 2	2300	30	40
27	2400	35	45
January 16	2500	35	45

C.7 Reverse Osmosis Logsheet

REVERSE OSMOSIS LOG SHEET

NJIT MASTERS THESIS STEVE LEON

	DATE DAY	DATE DAY	DATE DAY	DATE DAY	DATE DAY	DATE DAY	DATE DAY	DATE DAY	DATE DAY	DATE DAY
FEEDWATER										
COND (µmhos)										
SILICA (ppb)										
FLOW (gpm)										
TOTALIZER (gal x 100)										
pH										
TEMPERATURE (°C)										
SDI										
CHLORINE (TOTAL) (ppm)										
FILTERS										
MULTI-MEDIA (psi)										
CARTRIDGE (psi)										
CARTRIDGE SDI										
R-O SYSTEM										
UNIT IN-SERVICE										
HOURS										
PRODUCT FLOW (gpm)										
PRODUCT COND (µmhos)										
PRODUCT pH										
PRODUCT SILICA (ppb)										
PRODUCT TOTALIZER (gal x 100)										
R-O PUMP PRESSURE (psi)										
1st STAGE PRESS. (psi)										
PRODUCT PRESSURE (psi)										
1st STAGE O/L COND. (µmhos)										
2nd STAGE PRESS. (psi)										
2nd STAGE I/L COND (µmhos)										
2nd STAGE O/L COND. (µmhos)										
CONCENTRATE PRESS (psi)										
CONCENTRATE FLOW (gpm)										
CONCENTRATE pH										
CONCENTRATE COND.										
% REJECT (COND.)										
% REJECT (SILICA)										
NORMALIZED PRODUCT FLOW										
PERCENT RECOVERY										
DECARBONATOR										
CO IN (ppm)										
CO OUT (ppm)										
CO REDUCTION (ppm)										
MUT										
MUT IN-SERVICE										
COND. (µmhos)										
SILICA (ppb)										
TOTALIZER (gal x 100)										

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