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# Membrane Technologies for Nitrate and Atrazine Removal from a Surface Water Source

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# REPORT SUMMARY

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## Background

Water treatment suppliers need to be able to remove agricultural by-products, including nitrates and pesticides, from drinking water. Conventional treatments, such as lime softening and multi-media anthracite/sand filtration, do not adequately remove these organic and inorganic dissolved solids. Membrane filtration is a promising alternative technology that is becoming more prevalent in the U.S. Recent advances have made membrane systems more effective and less costly. However, the retentate from these systems must be treated as industrial waste. Microfiltration and ultrafiltration can remove suspended solids and natural organic material while nanofiltration and reverse osmosis can reduce the level of dissolved solids. This report discusses these and other types of membrane filtration.

## Objective

This study had two main objectives—the construction of a membrane filtration pilot plant that could be operated in Decatur, Illinois and the operation of that plant in the form of a feasibility study. This study would be concerned with evaluating the cost and efficiency of contaminant removal from surface water.

## Approach

In the pilot plant, ultrafiltration was used as a pretreatment for reverse osmosis. The system was monitored throughout the pilot study, measuring data in such a way that comparisons could be made between these results and some of EPRI's other studies in the area of nitrate removal.

## Results

Nitrate and atrazine were found to be significantly reduced when reverse osmosis methods were used. In addition RO offers a cost-effective nitrate removal solution. However, ultrafiltration did not appear to be very effective as a pretreatment method for reverse osmosis; there were only minor differences between water that had been pretreated with ultrafiltration technology and water that had not been pretreated in this manner.

Specifically, the following results were obtained:

1. Pretreatment with coagulation aids was necessary to prevent rapid build-up of organic and cake fouling on the membrane. Sulfuric acid and sodium hexametaphosphate was found helpful to prevent scaling. If blending is required then no additional post-treatment was required. If no blending occurs, then the RO permeate requires a caustic soda addition.

2. The UF retentates are higher in suspended solids and total organic carbon. The NF/RO retentates were found to have resolved solids concentrations below 1000 mg/L TDS. After the wastewater is produced, these rejects could be wasted to the wastewater treatment plant or used for field irrigation. The rejects may not necessarily be beneficial for agricultural production if used as the sole nitrogen and pesticide source.

3. The TFC-HR membrane provided the best nitrate removal but the older cellulose acetate membranes were not as effective for nitrate removal. All membranes tested reduced atrazine levels, with the HR-PA, TFC-HR, and CPA2 membranes reducing the atrazine levels below 0.1 ppb.

4. More long-term tests are required to determine cleaning frequencies. However, normal operation for NF/RO systems would indicate cleaning cycles every 4-6 months with this type of water quality. For the UF systems, the cleaning frequencies occurred about every 4 to 5 days using a general type soap, with coagulation. More intense cleaning was required every 3-4 months.

5. Remote operation was shown to be feasible and allowed for changing water quality. The remote operation instantly alerted operators when system failures occurred.

6. Literature review indicated that several factors can cause membrane fouling. These include cake or jell layer formation; precipitation or scaling of the membrane; adsorption of organic compounds onto membrane surfaces, and biofouling. As part of another project, an attempt was made to evaluate the surfaces of the membranes but no reliable results were obtained.

7. The cost for a 10 MGD (37,850 $m^3$/d) RO treatment system were determined to be in the range of $6.9-$9.3 million. Maximum operating costs would be about $4100 per day.

### EPRI Perspective

EPRI's Municipal Water and Wastewater Program was created to help member utilities address the energy needs of the more than 60,000 water systems and 15,000 wastewater systems in the United States. These facilities are among the country's largest energy consumers, requiring an estimated 75 billion kWh nationally, about 3% of the annual U.S. electricity use.

Clean drinking water and effective wastewater treatment are vital services needed in all communities. These safeguards protect the public health, strengthen the community infrastructure, and provide a foundation for economic growth. Agricultural by-products in drinking water can pose a health risk. By demonstrating the feasibility of membrane technologies, EPRI can play in role in increasing water safety and also electrical load. Other EPRI publications that are concerned with the use of electrotechnologies to improve the safety of drinking water include *Ozonation in Drinking Water Treatment for the Reduction of Herbicide Levels* (CR-104077), *UV Disinfection for Water and Wastewater Treatment* (CR-105252), *Demonstration of Biological Denitrification of Drinking Water for Rural Communities* (CR-108884) and *Ultraviolet Disinfection of Drinking Water* (TA-107716).

**TR-111650**

**Interest Category**

L3004 Municipal Water and Wastewater

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# 1

# INTRODUCTION

Water treatment suppliers have a number of concerns related to existing and possible future legislation to reduce public health risks. Water quality reports for a handful of Midwestern cities indicate an ever increasing concern that agricultural by-products, specifically nitrates and pesticides, have contaminated both ground and surface waters. Unfortunately, these pollutants pose a threat to public health because conventional treatment plants with lime-softening and multi-media anthracite/sand filtration are inadequate for the removal of organic and inorganic dissolved solids.

On the forefront of alternative technologies, membrane filtration offers the ability for treatment systems to meet USEPA standards set forth in both the safe water drinking act (SWDA) and the yet to be approved, interim enhanced surface water treatment rule (IESWTR). Various types of filtration can be applied to a particular water source to meet the needs of individual plants. Microfiltration (MF) and ultrafiltration (UF) can remove suspended solids consisting of bacteria and viruses, as well as natural organic material (NOM); this can reduce enteriological infections along with risks associated with carcinogenic disinfection by-product (DBP). Nanofiltration (NF) and Reverse Osmosis (RO), or hyperfiltration (HF), directly provide high removals of dissolved material ranging from pesticides (i.e. Atrazine) to $Ca^{+2}$, $Mg^{+2}$, $SO_4^{-2}$, $Na^+$, $NO_3^-$, and $Cl^-$, ions. The combination of filtration systems with powdered activated carbon (PAC), sedimentation, or conventional pretreatment can significantly increase membrane life while rendering a consistent, viable source of municipal water.

Recent developments in materials and manufacturing processes have markedly reduced costs of membrane systems for full scale use. The present spiral wound RO membrane elements can produce high quality permeate with less surface area and lower pressures than older systems. Compared to the cellulose acetate (CA) blend membranes of 5-10 years ago, thin-film composite (TFC) or polyamide (PA) style RO elements offer better rejection of monovalent ions with higher specific flux over a broader range pH. Improved quality of membrane element manufacturing, including the thin film itself and the element connections, increases membrane reliability as a barrier to organisms such as *Cryptosporidium* and *Giardia*. With years of test data, membrane manufacturers can now guarantee turbidity removal below 0.1 NTU for their hollow-fiber UF modules at normal operating conditions.

In addition to these benefits, there are also some important drawbacks. As permeate is produced, a more concentrated retentate is formed. Recovery rates of up to 80% of the feed water can be achieved leaving the remaining 20% to be treated as industrial wastes in either lagoons, deep-injection type wells, or wastewater treatment plants. Should the membrane fail there is a possibility of infectious agents passing through the system. Another drawback of the process includes the amount of energy required to pressurize the system, in the range of 100-800 psi (689-5516 kPA) for permeate production. Finally, repeated cleaning of the membranes can irreversibly impair the membranes causing a reduction in performance.

Without cleaning and pretreatment chemicals, however, increased detrimental degradation of membrane performance over time could occur. Several types of fouling can diminish the performance of a system. Cake or gel-layer formation involves the collection of particles on the membrane surface causing an increased resistance to transmembrane flow. With concentration of various ions beyond their saturation point, precipitation on the membrane surface can occur. In the case of absorptive fouling, organic material adsorbs to membrane surfaces reducing flux and the fundamental ability of the membrane to filter various components. Long term irreversible fouling will limit the operational time span for a given element regardless of rejuvenation technique chosen (Mallevialle et al., 1996).

In comparison to other methods of nitrate removal, reverse osmosis is competitive with ion exchange (IX) and biological denitrification. With the use of IX resins, anionic resins could effectively remove all nitrates present producing only higher concentrations of chloride in the IX effluent. Another factor for design and operation of an IX system includes the use of a concentrated brine solution to regenerate the resin.

Biological denitrification removes nitrates using anaerobic towers of denitrifying bacteria. This process can remove up to 85% of nitrates present. Filtration of the effluent is required to remove the various bacteria sloughed off from the process. It is not known if these bacteria or the organic substrates required in the process could affect the amounts of TOC and DBP precursors present. However, the perception of drinking water treated with bacteria often concerns consumers. Costs of biological denitrification can be as high as \$5.00/1000 gallons (\$1.32/m$^3$), compared to the \$0.60/1000 gallons (\$0.16/m$^3$) for IX and \$1.05/1000 gallons (\$0.28/m$^3$) for high pressure RO (Turner, 1998). Lower pressure RO system costs are around \$0.55 per 1000 gallons (\$0.15/m$^3$) (EPRI, 1997). The secondary health benefits offered by membrane filtration, such as DBP and bacteriological decontamination, could indirectly make up for their higher cost.

Regardless of technology, the use of membrane filtration in the United States is becoming more prevalent. Malcolm Pirnie, Inc. recently completed construction of the largest membrane treatment facility in the world at Del Rio, Texas to produce over 28 MGD (106,000 m$^3$/d) of potable water from reclaimed water sources. Smaller cities such as Toluca, IL use RO to produce waters free of radionucleides. The USEPA's Information Collection Rule has set an unprecedented path for the municipal use of membranes by implementing bench and pilot scale RO testing at a variety of U.S. cities. With further understanding and implementation of membrane filtration processes, public drinking water suppliers will be able to tackle their contaminant problems and hopefully meet both present and future regulations.

# 2

# RESEARCH OBJECTIVES

As part of the study conducted by the University of Illinois at Urbana-Champaign, goals for the project were set as a guide to project activities. The two-year study had two primary tasks. The first task consisted of constructing a pilot plant that could be operated in Decatur, Illinois, while the second task was to operate the system in the form of a feasibility study.

## Task 1

During the first year, various components of the pilot plant were acquired and assembled into one full system. The system was to consist of an existing UF membrane plant donated for use by the USEPA Risk Management Laboratory and a RO membrane plant on loan from the Illinois DNR Waste Management and Research Center. Once received, both pilots were to be outfitted with electrical transducers and controls and coordinated using LabView software for remote operation and control. During the first year of the project this goal was met by assembling a new UF pilot to be placed in series with the modified WRMC RO plant in Decatur, IL. This task was near completion in early December, 1997.

Another part of the first task was to develop a testing protocol that allowed for comparison of results with EPRI's other studies in the area of nitrate removal. Three facilities, consisting of reverse osmosis, ion exchange, and biological denitrification, could then be evaluated for cost and efficiency of contaminant removal. This goal was met by obtaining seven different RO membrane elements. A laboratory regime was also set up to measure important water quality parameters throughout the pilot study.

## Task 2

With the system in place at the South Water Treatment Facilities, Decatur, Illinois, the second task involved operating the system and collecting perfomance data. With this protocol, the following questions could then be addressed:

1. In considering pretreatment prior to membrane filtration, what types of pretreatment would be required for fouling and scaling reduction? Can PAC

improve the operation of the UF pilot by preventing organic fouling on the UF membrane? After RO, what types of post-treatment might be required?

2. What is the chemical makeup of the UF and NF/RO rejects? What can be done to the waste after it is produced? Are these rejects beneficial for use in agricultural production?

3. Which types of NF/RO membrane elements are effective in the removal of pesticides and nitrates?

4. For longer-term operation, what are the necessary backflush and cleaning frequencies for each type of membrane? What rate of water production could be maintained by the system, considering possible seasonal variability?

5. Is remote operation of a computer controlled membrane pilot plant feasible?

6. What causes fouling of the UF and NF/RO membranes? Can surface analysis tools observe differences between individual membranes to explain why one performs better than another with a given chemical pretreatment to prevent fouling?

7. What would be the estimated full scale capital and operating cost for a system to serve a community the size of Decatur, Illinois?

For the pilot study, a short assay on each of the RO elements was conducted with ultrafiltered and tap water for a range of feed pressures. Following the short assays, longer term tests were conducted to prove the viability of nitrate and pesticide removal using the RO system. With the experimental data, the ultimate goal for the project is to provide Decatur, Illinois with a thorough assessment of membrane technologies, specifically UF and RO, to help them consider compliance issues related to current and future drinking water regulations.

# 3

# LITERATURE REVIEW

## Fundamentals of Membrane Processes

Membrane processes involve the use of a thin film or porous material to separate one compound or phase from another. Specific membrane materials can directly remove particles or dissolved solids to produce water that can meet any requirement. Mechanical and electrochemical forces drive membrane processes to create water (or permeate) lower in solids. From the process, a concentrated retentate is generated that can be further concentrated or wasted according to state and federal regulations.

### *Materials*

Since the first membrane materials were described by Reid and Breton and by Loeb and Sourirajan in the late 1950's, numerous materials have been developed to improve the capacity and performance of membrane filters. For a given treatment stream, a certain membrane material can be selected from an assortment of candidates. Inorganic, ceramic, and polymeric synthetic membranes are growing in use across the United States for both municipal and industrial processes. While many membrane materials are available, three materials will be covered here because of their use during the project.

#### Cellulose Acetate

Cellulose diacetate (CDA), is a hydrophilic chain of cellulose polymers derived by the acetylation of cellulose acetate, using formamide and acetone. When sulfalane and polyethylene glycol are added, cellulose triacetate (CTA) is formed. The mixture of CDA and CTA in a CA type membrane allows for high strength with chemical durability (Montgomery, 1985). The disadvantage of this material can be found in its narrow temperature and pH operating range. These limits for CA type membranes fall within a pH of 2 to 6 and a maximum temperature of 35°C (95°F). Free chlorine concentrations above 1 ppm for continuous flow and 50 ppm on a short-term basis can also hinder CA membrane performance. Due to the fact that microbial growth could occur on CA membranes, it is necessary to disinfect the membrane surfaces. Because both can potentially harm the membrane, careful monitoring of free chlorine must be undertaken to ensure that disinfection occurs without overexposing the membrane material (Cheryan, 1986). Although previous experiences with reverse osmosis (RO) plants have indicated poor rejections of nitrate (49%), calcium and sulfate rejections were much higher (85%-98%) (Montgomery, 1985). CA is more common than polyamide for use in nanofiltration and ultrafiltration applications.

### Polysulfone

Polysulfone (PS) type membranes can be implemented to work in a broader range of operating conditions than CA. For the most part, PS membranes are hydrophobic with a pH tolerance of 1 to 13 for temperatures up to 75°C (167°F). For disinfection purposes, concentrations of free chlorine to 50 ppm are tolerable on a regular basis. Unfortunately, the weakness of the PS material lies in its inability to withstand pressures above 25 psi (172 kPa) and its high potential to irreversibly foul (Cheryan, 1986).

### Polyamide

In order to increase strength in the material with improved chemical resistance, polyamide (PA) type materials were developed from aromatic polymers and nitrogen containing amide groups. With new techniques in interfacial polymerization originally developed by Cadotte in the early 1970's, micron thin layers of polyamide can be applied to porous materials of increased structural strength (Riley et al., 1976). The polyamide substance can withstand pH between 3 to 11 and temperatures to 40°C (104°F). Compared to CA RO elements, PA elements show higher rejection of salts (>97%) at lower pressures (200-400 psi or 1379-2758 kPa). Polyamide is even more resilient to chemical and biological attack. However, should free chlorine exist in the feed water, care should be taken to dechlorinate the water fully to prevent damage to the material which could lead to decreased rejection of dissolved material.

### Thin-Filmed Composite

By using combinations of materials, a thin-filmed composite (TFC) membrane can be created by layering materials onto one another. By doing so, a slightly permeable, weak membrane made of polyamide can be placed onto a stronger and more porous polysulfone or polyester material that can withstand high pressures. Mixtures of polymers and organic solvents can be used to produce the type of membrane needed for a particular application or operating condition. A typical thin film composite membrane consists of interfacial polymerization of PA on a supporting layer of PS.

### Manufacturing

Several manufacturing techniques can be used to form a variety of modules. These modules vary in size to produce water at rates of a few milliliters per minute to thousands of gallons per day. Membrane filtration systems can be grouped according to the membrane and plant configuration. The most common modules are tubular, plate and frame, spiral wound, and hollow fibers. Modules for higher pressure operations, such as RO and NF, are usually spiral wound units. Lower pressure units, such as UF, utilize hollow-fiber configurations.

When casting a membrane, two different pore structures, mircoporous and anisotropic, can be formed. In order to create a membrane, thin polymer films are spread onto a stronger backing material that can withstand the pressures involved in a given membrane filter. The combination of layers can be extruded to form either a hollow fiber or sheet of material that can then be assembled into a spiral-wound element or hollow fiber module. A symmetrical, or mircoporous pore formation creates a uniform diameter through the thickness of the membrane material and is more common in inorganic membranes. In asymmetric, or anisotropic membranes, pore diameter changes through the membrane. Most thin-filmed composites are asymmetric membranes (Cheryan, 1986).

Depending on the amount of surface area required for filtration, more hollow fibers or sheet surface area can be included. Six to seven elements are then placed into pressure vessels which are connected in series. While this report will not cover more on the technical details of membrane manufacturing, the following represents the types of elements used for this study.

#### Spiral Wound

Figure 3-1 depicts a typical spiral wound element. As sheets of membrane material are manufactured, a spacer and permeate collection tube is placed between the two opposing leafs on the backside of the membrane surface. The sheet edges are glued together around the collection tube and spacer. A feed spacer is then placed on the membrane surface and both are wrapped around the permeate collector. The feed spacers promote turbulence through the membrane element to prevent detrimental concentration-polarization while serving as a support structure. While the unit can not be backwashed due to the nature of the thin film membrane, the compact spiral wound element forms an inexpensive system that can be used continuously with regular shutdowns for chemical cleaning and maintenance (Cheryan, 1986).

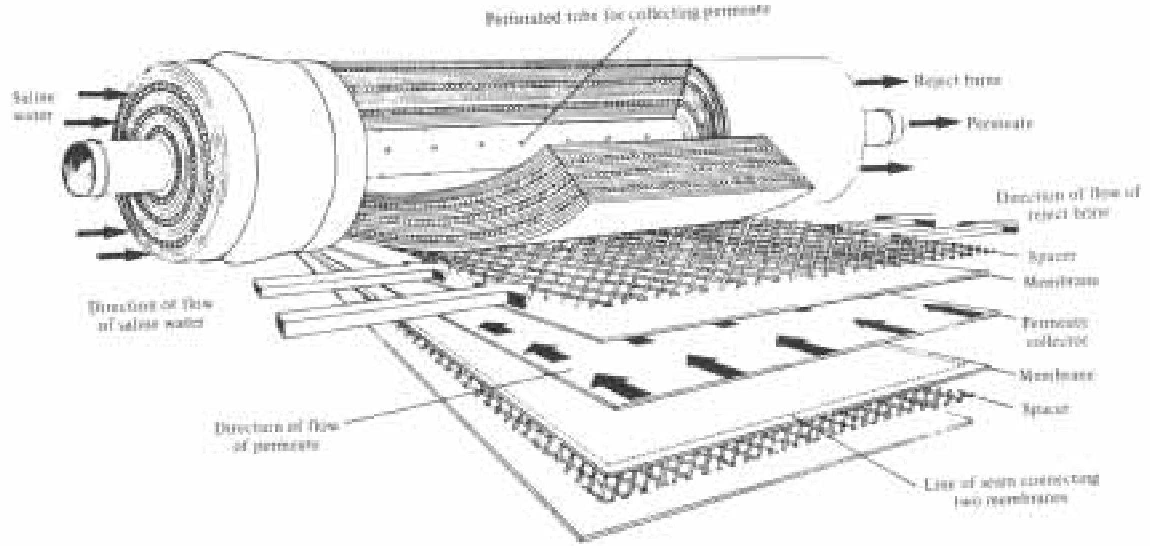

**Figure 3-1**
**Schematic of a Spiral Wound Element, *Water Supply and Pollution Control* (1993)**

Should errors occur during assembly and installation, it is possible for o-rings, membrane glue joints, and the membrane material itself to allow passage of microbial contaminants (Taylor et al., 1996). It has been suggested that particle counters at permeate outlets could detect passage of pathogenic contaminants through the system.

### Hollow Fiber

A typical hollow fiber module is illustrated in Figure 3-2. Depending on the formation of the asymmetrical porous structure, filtration of the water can occur from the outside-in or inside-out. Figure 3-2 represents the inside-out configuration of a hollow fiber module. Unlike the spiral wound module, hollow fibers can be backflushed to reduce fouling and wash away gel-layer buildup on the surface of the hollow fiber. Over time, the hollow fibers will require chemical cleaning, but at reduced frequencies in comparison with spiral wound elements because of the backwashing capability. Operation of hollow fiber polymeric UF modules is limited by maximum pressures less than 29 psi (200 kPa) (Cheryan, 1986).

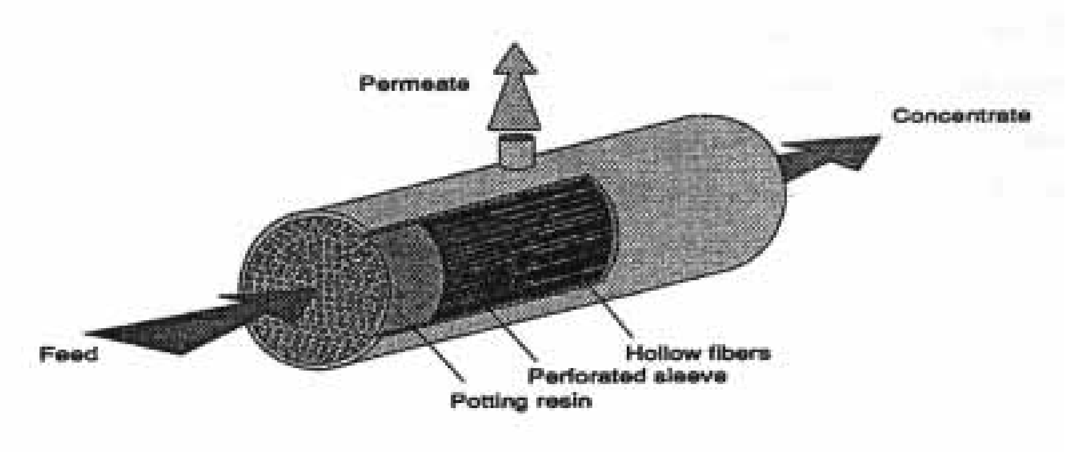

**Figure 3-2**
**Schematic of a Hollow Fiber Module, *Water Processes Membrane Filtration* (1996)**

### *The Filtration Spectrum*

Figure 3-3 illustrates the range of solids removal for a given membrane filtration process. As seen in the graph, each process can remove a particular range of contaminant from bacteria to ions. Filtration requires increasing energy (or pressure) with decreasing size of the material to be removed.

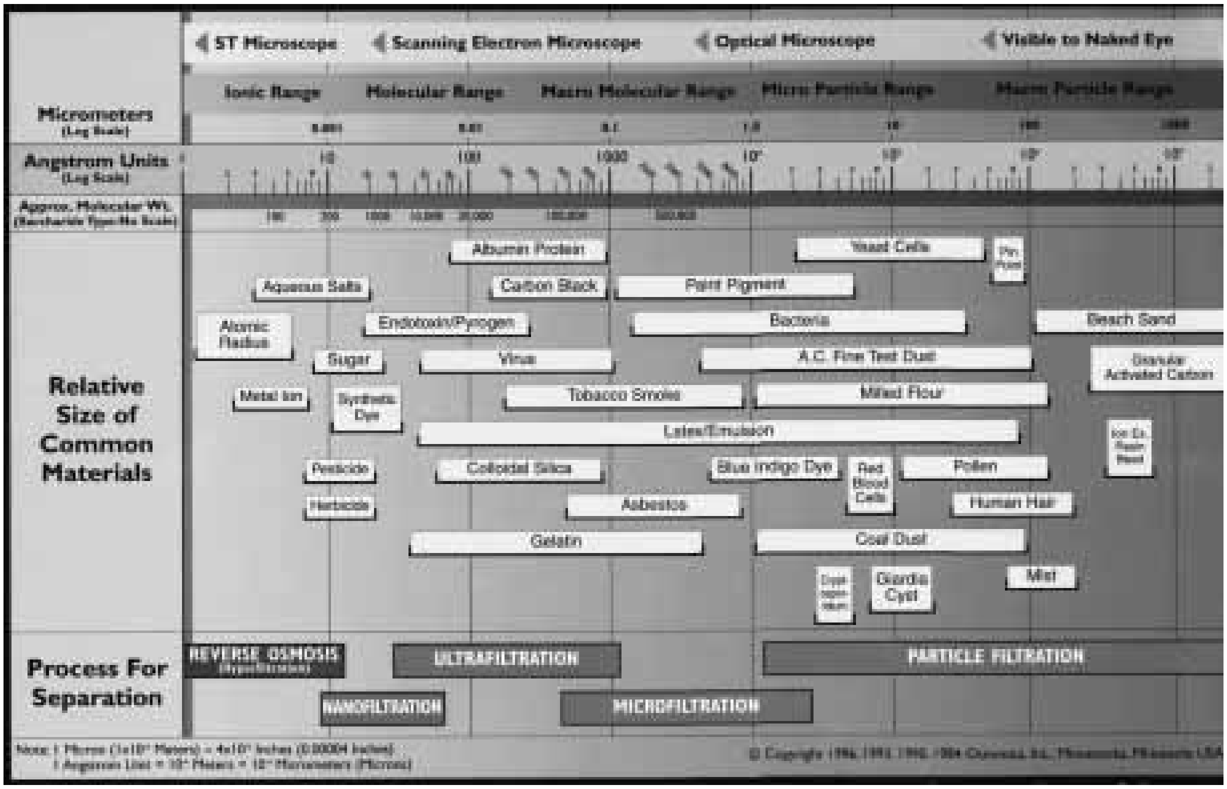

**Figure 3-3**
**The Filtration Spectrum, Osmonics, Inc. (1996)**

## Ultrafiltration—UF

Ultrafiltration (UF) uses a porous type membrane to remove all particulate and colloidal material and larger molecular weight compounds in the range of 1,000 to 100,000 Daltons. UF modules are rated by their molecular weight cut-off (MWCO) as an approximation of the smallest particle to be rejected by a particular membrane (Cheryan, 1986). During the manufacturing process, variations occur in the material that allow membranes with the same estimated MWCO to behave differently. Under most circumstances, UF provides significant protection from viruses and bacteria if the membrane remains uncompromised. UF membranes usually consist of cellulose acetate or polysulfone hollow fibers. Other configurations use spiral-wound elements similar to RO, but operated at lower pressures.

## Nanofiltration—NF

When considering membrane processes tighter than UF, nanofiltration (NF) can provide high rejections of ions from 0.001 microns in diameter to low MWCO's of 1000 Daltons. NF is a diffusional process where the transport of dissolved solids does not necessarily occur through pores, but by diffusing through the membrane material.

Commonly used as a softening process at pressures of 100-200 psi (689-1379 kPa), NF offers cost savings over RO systems when lower overall salt rejections are allowable. Compared to UF, NF membranes can furnish removal of color, turbidity, DBPs, and hardness. Virtually all bacteria and viruses can be removed by NF, but UF membranes are typically less expensive if used for this particular purpose (Taylor et al., 1996). NF/RO Membranes with MWCO's of 200-300 Daltons can achieve reductions of natural organic matter (NOM) in the range of 90% (Nilson and Digiano, 1996).

#### Reverse Osmosis—RO

Reverse osmosis (RO), or hyperfiltration, membranes have the capacity to remove solids 0.0001 microns in diameter or larger, creating a water practically void of all dissolved material. Originally designed for the desalination of brackish seawater, high-pressure RO elements (800-1000 psi or 5516-6895 kPa) are able to reduce TDS from 34,000 mg/L to 500 mg/L for potable drinking water. Today, lower pressure RO elements (200-400 psi or 1379-2758 kPa) offer an affordable treatment option for the removal of dissolved solids from fresh water sources. Like NF, RO is a diffusional process. Differences in membrane material and manufacturing are responsible for the higher rejections and required pressures. Along with turbidity removal and softening, RO is very effective for the removal of dissolved organic carbon. Of particular interest, pesticides and DBP pre-cursors can be removed directly from source waters depending upon feed water quality. Most monovalant ions, for example nitrate ($NO_3^-$) and sodium ($Na^+$) can be separated as well (Taylor et al., 1996).

### *Membrane Terminology*

The following equations are representative descriptions of membrane filtration models.

#### UF Membrane Processes

When using UF processes to treat fresh water sources, osmotic pressure is negligible. The following equations derived from membrane models can then be simplified and applied as a basic understanding of the UF process. The following represents some of the rudiments of UF filtration.

##### Transmembrane Pressure

Transmembrane pressure, $P_{tm}$, refers to the amount of pressure applied across the membrane filter, as described in Equation 3.1 (Jacangelo, 1996).

$$P_{tm} = \left( \frac{P_i + P_o}{2} \right) - P_p \qquad (3.1)$$

Where:

$P_i$ = Element Inlet Pressure
$P_o$ = Element Outlet Pressure
$P_p$ = Permeate Pressure

### UF Membrane Flux

**Transmembrane Flux**
At any given temperature or pressure for a given UF module, permeate flow can then be used to determine the amount of water crossing a membrane surface. The Transmembrane Flux, $J_t$, can be defined as Equation 3.2 (Jacangelo, 1996).

$$J_t = \frac{Q_p}{A_m} \tag{3.2}$$

Where:

$Q_p$ = Permeate Flow
$A_m$ = Membrane Surface Area

**Specific Flux**
The specific flux, $J_s$, allows for the normalization of the transmembrane flux with the transmembrane pressure. Equation 3.3 represents the specific flux (Jacangelo, 1996).

$$J_s = \frac{J_t}{P_{tm}} \tag{3.3}$$

**Normalized Specific Flux**
For the Aquasource UF modules, it is important to take into account temperature effects when comparing flux rates at different conditions. Equation 3.4 represents the normalized specific flux, $J_{s,20}$, at 20°C (68°F) with an error less than 0.6% (Vickers, 1998).

$$J_{s,20} = J_s \cdot (1.7921 - 0.058648T + 0.0011474T^2 - 0.000009745T^3) \tag{3.4}$$

Where:

$T$ = Temperature (°C)

**Gel Layer Formation**
As the formation of a cake or gel layer occurs on the surface of the UF membrane, a resistance to flux occurs. According to the gel polarization model, the hydraulic

resistance, $R_h$, can represent the sum of the intrinsic membrane, $R_m$, irreversible fouling, $R_i$, and cake, $R_c$ resistances (Anselme and Jacobs, 1996).

$$R_h = R_m + R_i + R_c \tag{3.5}$$

Over the length of a filtration cycle, the resistance increases and causes flux decline as represented in Equation 3.6 (Wetterau and Clark , 1996).

$$J_t = \frac{P_{tm}}{\mu \cdot R_h} \tag{3.6}$$

Where:

$\mu$ = Fluid Viscosity

### UF Feed Water Recovery

Feed water recovery for a UF module, $R_{f\text{-}UF}$, is an estimation of the amount of filtered water that can be produced versus the amount of feed water pumped into the membrane module. This is an important parameter to determine module efficiency and for planning future needs for pumping and waste storage (Jacangelo, 1996).

$$R_{f-UF} = \frac{Q_p \cdot t_p - Q_b \cdot t_b}{Q_f \cdot t_p} \tag{3.7}$$

Where:

$Q_b$ = Backwash Flow
$Q_f$ = Feed Water Flow
$t_p$ = Time for Permeate Production
$t_b$ = Time for Backwash Procedure

### RO Membrane Processes

With increased concentrations of dissolved solids, osmotic pressure may become a factor in designing for RO treatment. For instance, the feed water from *Lake Decatur* may exert an osmotic pressure of 10 psi (69 kPa), while brackish seawater can exert osmotic pressures above 300 psi (2068 kPa).

### Osmotic Pressure

Osmotic pressure, $\pi$, represents the amount of pressure that can be created between a concentrated solution and pure water separated by a permeable membrane. Equation 3.8 represents the osmotic pressure for a single electrolyte solution (Robinson and Stokes, 1955).

$$\pi = c_T \cdot R \cdot T \cdot \phi \tag{3.8}$$

Where:

$c_T$ = Total Anionic and Cationic Molar Concentration
$R$ = Ideal Gas Constant (8,312 *Pa/M-K*)
$T$ = Temperature (°K)
$\phi$ = Molar Osmotic Coefficient

When considering solutions near infinite dilution, the molar osmotic coefficient, $\phi$, reaches unity. The osmotic coefficient will vary between particular membrane materials (Mariñas and Urama, 1996).

### RO Membrane Flux

**Water or Solvent Flux**

The water flux, $J_v$, for the RO process is similar to the transmembrane flux of the UF process except for additional consideration of osmotic pressure (Mariñas and Urama, 1996).

$$J_v = A \cdot (P_{tm} - \sigma \cdot \Delta\pi) \tag{3.9}$$

Where:

$A$ = Water Permeation Coefficient
$P_{tm}$ = Transmembrane Pressure
$\sigma$ = Reflection Coefficient
$\Delta\pi = \pi_f - \pi_p$ = Difference in Osmotic Pressure
$\pi_f$ = Feed Osmotic Pressure
$\pi_p$ = Permeate Osmotic Pressure

For a reflection coefficient of zero, the membrane has no particular selective capability for either solvent or solute, and osmotic pressure would not be a factor in determining flux (similar to UF membranes). At $\sigma = 1$, the equation would represent an ideal RO membrane where the mechanism of transport relies solely on diffusion created by the concentration gradient between solute and solvent (Spiegler and Kedem, 1966).

In regards to the water permeation coefficient, $A$, bench and pilot studies with single and multi-component solutions are required to estimate this empirical constant. The estimation of $A$ is done by varying feed pressures for various solutions.

**Solute Flux**

As water passes through a membrane, it is possible to estimate the flux of other molecules as well. Equation 3.10 represents the solute flux, $J_c$ (Mariñas and Urama, 1996).

$$J_c = J_v \cdot c_p = B \cdot \Delta c + (1-\sigma) \cdot J_v \cdot \bar{c} \tag{3.10}$$

Where:

$c_p$ = Permeate Solute Concentration
$B$ = Solute Permeation Coefficient
$\Delta c = c_w - c_p$ = Difference in Effective Concentration
$c_w$ = Solute Concentration at Membrane Wall
$\bar{c}$ = Arithmetical mean of $c_w$ and $c_p$

The solute permeation coefficient, $B$, is another empirical constant determined through further testing with RO membrane elements.

**Temperature Corrected Flux**

In order to correct for differences in water viscosity at dissimilar temperatures, the temperature corrected flux, $J_{v,25}$, at 25°C (77°F) can be estimated from a water flux, $J_v$, at a given temperature $T$ in degrees Celsius. While used during this project to compare the different membrane elements at one specific temperature, it should be noted that equation 3.11 only takes into account the differences in feed water viscosity and not intrinsic membrane characteristics. Unlike the temperature correction equation (Equation 3.4) for use with the Aquasource UF membrane, this standard equation (3.11) is used for RO flux temperature correction. The correction factor is given by (Taylor et al., 1996):

$$J_{v,25} = \frac{J_v}{1.03^{(T-25)}} \tag{3.11}$$

### RO Feed Water Recovery

The feed water recovery, $R_{f\text{-}RO}$, estimates the amount of permeate recovered from the feed flow into the RO element. The RO recovery from a spiral wound element does not require backflushing parameters compared to UF recovery (Taylor et al., 1996).

$$R_{f-RO} = \frac{Q_p}{Q_f} = \frac{Q_p}{Q_r + Q_p} \tag{3.12}$$

Where:

$Q_r$ = Retentate Flow

### Solute Rejection

For individual components of the feed solution, the solute rejection, $X_i$, is given by:

$$X_i = 1 - \frac{c_p}{c_f} \tag{3.13}$$

Where:

$c_f$ = Feed Solute Concentration

### Concentration-Polarization

Concentration polarization is important to membrane operation and modeling. It represents increased concentrations of rejected solutes near the membrane surface. Figure 3-4 represents the concentration-polarization layer of thickness δ at the membrane surface (Taylor et al., 1996).

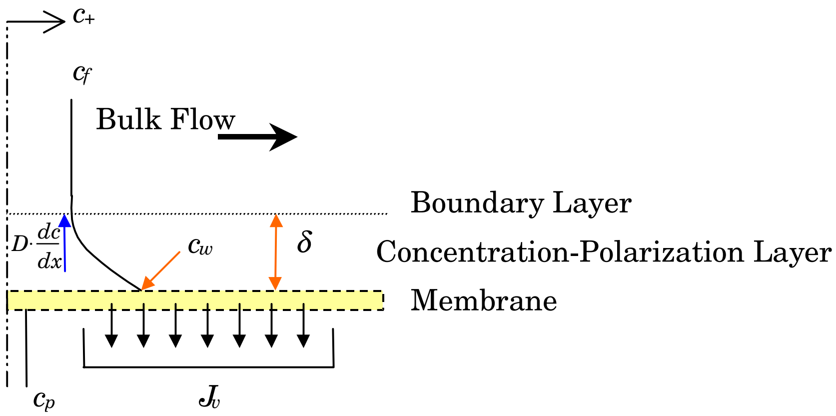

**Figure 3-4**
**Schematic of Concentration-Polarization Layer at the Membrane Surface, *Water Treatment Membrane Processes* (1996)**

The concentration-polarization layer is described with the classic film model (Mariñas and Urama, 1996).

$$\frac{c_w - c_p}{c_f - c_p} = \exp\left(\frac{J_v \cdot \delta}{D}\right) \tag{3.14}$$

Where:

$D$ = Molecular Diffusion Coefficient

### *Membrane Fouling*

#### Fouling and Flux Decline

Several factors can lead to decreases in membrane performance. As mentioned in the previous section, cake and concentration-polarization layer formation can inhibit flux by forming either a resistive solids layer or concentration gradient.

In the case of cake layer formation, the thickness of such a layer will increase flux resistance in proportion to cake thickness and properties. Particle charge and diameter will affect the formation of such a cake layer over the filtration cycle. Over time, the cake can act as an alternative sieve to remove additional particles; it can also compact as pressure is increased. Either backflushing or chemical cleaning can physically remove cake layers.

Unlike cake formation, concentration-polarization limits flux by creating a concentration gradient. This gradient increases the electroviscosity or electrochemical potential at the surface leading to increased osmotic pressure. Crossflow velocity can reduce the formation of the concentration-polarization layer. Elimination of the layer is accomplished once the feed flow has been stopped.

Due to the formation of concentrated layers near the membrane surface, precipitants may form on the membrane. Precipitation or scaling on the membrane occurs when concentrations of salts have exceeded their saturation point. Solutions with barium, calcium, magnesium, iron, and maganese are of concern because they precipitate when combined with the proper amounts of sulfates, carbonates, and hydroxides. The formation of salt crystals on membrane surfaces act as an irreversible cake layer which must be chemically cleaned.

Within the makeup of a resistive layer, compounds that can adsorb directly to the membrane surface can permanently impair membrane flux. The ability of organic compounds or NOM to adsorb onto membrane surfaces is affected by solution chemistry (including pH), membrane hydrophobicity and surface charge (Jucker and Clark, 1994). After adsorption onto the membrane, the organic material can indirectly foul the membrane further by cementing colloidal deposits in the cake layer at the membrane surface. The nature of this adsorptive fouling is largely irreversible when considering backwash operations. Chemical cleaning is often required to dissolve the organic material through the use of caustic or enzymatic cleaners (Wiesner and Aptel, 1996).

In addition to adsorptive fouling, biological growth can occur on membrane surfaces and attack the membrane material, reducing its efficiency. The build up of biofilms on filtration surfaces can occur rapidly over days or slowly over months, depending on the feed water source. As the build-up occurs flux resistance increases as in the case of cake layer formation. Bacterial digestion of the membrane material can cause destructive microchannels in the surface material that render elements useless. Control of bacteriological growth can be accomplished using prechlorination/dechlorination prior to filtration or the addition of bacterial inhibitors to the feed water (Ridgway and Flemming, 1996).

## Fouling Prevention and Reversal

### Surface Modifications

To prevent fouling, advances in membrane technology allow modifications of the active membrane surface. Operations such as hydrolization and postelectrolyte complexing modify the surface chemistry and decrease adsorptive fouling. Low temperature plasma and ultraviolet radiation are applications that aid in annealing and smoothing of surface features (Asfardjani et al., 1991).

To detect modifications in membrane surfaces which are beneficial to operation, surface characterization techniques such as atomic force microscopy (AFM) and scanning electron microscopy (SEM) can be used. Pore diameter distributions and permeability can be measured in UF membranes with AFM (Clark and Lucas, 1998). Combined with contact angle measurements to measure hydrophobicity, a better understanding of possible solution/surface interactions can be obtained (AWWA Membrane Committee, 1998).

### Pretreatment

Pretreatment of the feed water source with acids and sequestering agents can reduce precipitant fouling of the membrane (Wiesner and Aptel, 1996). Sulfuric or hydrochloric acid, for example, is added to lower the pH of the feed water below the Langlier saturation index for carbonate species. Sequestering agents, like sodium hexametaphosphate (SHMP) and EDTA, bind to metal ions to prevent precipitation of oxides and sulfates. Recent developments in commercial RO antiscalants offer improved performance without the need for acidification (Boffardi, 1997).

Prior to filtration of water, several pretreatment steps can be taken to increase the efficiency of the UF membrane process. Previous research presented by Henneghan and Clark (1991) and by Laîné et al. (1990) evaluated the use of PAC and coagulation prior to UF to enhance UF filter performance for DOC, turbidity and bacteria removal. PAC addition prior to filtration can remove DOC that is responsible for flux decline. When combined with coagulation, the reduced turbidity significantly improves flux over time. Iron oxide particles can also be used to remove NOM to control surface adsorption (Chang and Benjamin, 1996).

As in the case of this project, multistage filtration can further remove fouling precursors. By placing the PAC/UF process in series with NF/RO, operation is improved by normalizing variations usually associated with conventional treatment. As shown in Figure 3-5, conventional treatment with multi-media anthracite/sand filters have variable turbidity removal. The effluent turbidity is consistently less than 0.3 NTU which is below the current Illinois requirements of 0.5 NTU.

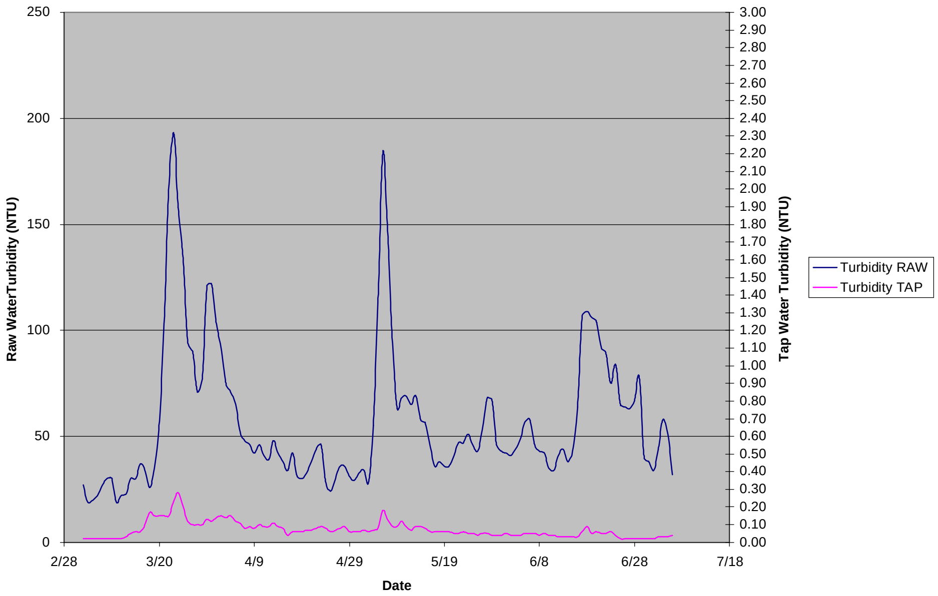

**Figure 3-5**
**Tap and Raw Water Turbidity after Conventional Treatment (SWTP - Decatur, Illinois)**

Full and pilot scale UF treatment plants consistently produce waters at or below 0.1 NTU regardless of feed water turbidity (Aquasource, Inc., 1992). UF is not designed to prevent precipitive fouling, but UF can remove NOM to reduce adsorptive fouling. UF feed waters for RO offer less particulate and adsorptive fouling than multi-media filters.

### Backwash Cleaning

Backwash cleaning of UF membranes occurs when the build up of reversible cake layers create a reduction in flux or increase in pressure beyond maximum operable levels. Prior to backwash, clean permeate is dosed with a disinfectant to minimize bacterial growth. The process of filtration is then reversed so that the dosed permeate flows opposite of normal operating conditions and forces the gel or cake layer from the membrane surface. The scoured material is then carried out of the system to sewer drains or storage for further treatment. Steps within the backflush procedure assure that the full module is flushed to minimize irreversible fouling and particle detention within the system.

### Chemical Cleaning

After flux declines below acceptable values, it is necessary to invoke chemical cleaning to rejuvenate the membrane to original production rates. Depending on the type of scaling or adsorption that has occurred, combinations of cleaners can be applied to reverse the fouling effects. Regular cleaning on a daily to weekly schedule is necessary during the life of an element. Acidic and descalant cleaning solutions remove most forms of inorganic scaling. Caustic solutions and chealating agents can be used for the removal of adsorbed organic compounds. Should biological growth occur, biocidal, detergent, enzymatic, and additional cleaners can be employed to treat the filtration unit (Ridgway and Flemming, 1996). Membrane manufacturers will recommend cleaning mixtures and application to avoid damage to their particular membranes (Taylor et al., 1996).

### *Post-treatment*

Due to the nature of permeate water quality, post-treatment is sometimes necessary to stabilize and/or disinfect the product water. Post-treatment considerations for surface water sources are often less than for subsurface waters because of the lack of hydrogen sulfide that can freely pass through membrane treatment. Dissolved hydrogen sulfide leads to the creation of unpleasant sulfide gas odor at the faucet (Montgomery, 1985). While UF and some NF processes may not be affected, lower pH RO permeates are naturally corrosive with low alkalinity and hardness. As with all municipal water supplies, disinfectants must be used to prevent bacteriological contamination according to the SDWA and disinfectants/disinfection by product rule (D/DBPR).

To solve problems associated with aggressive waters, additional methods such as blending and reconstitution can be implemented. Blending RO permeate with conventionally treated tap water can be used to recover alkalinity and hardness while diluting other compounds. When using only RO permeate, reconstitution of alkalinity and calcium hardness to saturation levels is normally accomplished with soda ash (sodium bicarbonate), quick lime, or aeration steps. In any event, the finished water quality parameters must meet state-specified standard values to optimize corrosion control under Section 141.82(g) and Section 141.87(d) of the Federal Lead and Copper Rule (61 FR 16348) (Pontius, 1996). In regards to this rule, the Illinois Environmental Protection Agency has set minimum conditions for pH, alkalinity, and hardness to prevent lead and copper leaching within the distribution system (Smith, 1998). These minimum conditions are set such that Decatur keeps a positive Langlier saturation index.

## Contaminant Issues

### *Nitrates*

#### Background

Organisms use nitrogen in both its reduced and oxidized states, forming a complex biogeochemical cycle which includes the atmosphere. Nitrification involves the oxidation of ammonia, $NH_4^+$, or nitrogen gas, $N_2$, first to nitrite, $NO_2^-$, and then nitrate, $NO_3^-$, via *nitrosomonas* and *nitrobacter* bacteria respectively. Denitrification breaks down nitrate to ammonia and nitrogen under anoxic conditions to complete the cycle (Brock et al., 1994).

Sources of nitrate in the environment are generally non-point sources such as by products of plant and animal wastes or natural deposits. Fertilizers, septic tanks and sewage represent anthropogenic sources of nitrate (Pontius, 1998). Application of anhydrous ammonia as a crop fertilizer results in nitrate via nitrification. Within the soil, nitrate is the most mobile nitrogen source in the watershed (Starrett et al., 1995). Regardless of source, inorganic nitrate compounds are highly soluble in water with little or no affinity for adsorption or precipitation.

When ingested, nitrate in the bloodstream is converted to nitrite. If not reduced to other metabolites by enzymes, nitrite binds directly to blood hemoglobin to prevent oxygen transfer causing methaemoglobinemia, cyanosis, or, as more commonly known, “blue baby syndrome”. The symptoms of cyanosis are noticeable by a bluish or brown tint of the skin created by methaemoglobulin present in the bloodstream. Eliminating the source of nitrate and/or nitrite can easily eliminate mild cases of “blue baby syndrome”. Severe cases may cause death without timely medical treatment with intravenous methylene blue or exchange transfusion. Infants from birth to six months of age are extremely susceptible to methaemoglobinemia because they lack the necessary enzymes to eliminate nitrites from their bloodstream. In children older than 6 months, adults, and pregnant women, no acute toxicological effects can be noted from normal levels of nitrate ingestion.

Nitrate and nitrite alone are not carcinogenic in nature and are also found in common foodstuffs. If exposed to elevated levels of both nitrite and amines, the possible formation of nitrosamines from the combination could create a small risk of cancer (Mara and Clapham, 1997). It should be observed that fruits and vegetables are consumed on a daily basis that contain naturally occurring levels of nitrites and nitrates. As reported by Professor Robert Cassens, the meat packing industry uses nitrite salts to preserve meats from spoiling at levels set by the Food and Drug Administration (1998). The presence of nitrate in nature is unavoidable.

## Regulations

In the case of drinking water, the USEPA regulates and enforces maximum nitrate concentrations through the safe drinking water act (SDWA). The maximum contaminant level (MCL) for nitrate is 10 mg $NO_3^-$ as N/L or 45 mg $NO_3^-$/L. The maximum contaminant level goal (MCLG) is the same as the MCL for this contaminant. When Illinois drinking water concentrations of nitrate exceed 8.5 mg $NO_3^-$ as N/L, public notification of elevated nitrate levels is required by the IEPA along with the distribution of nitrate-free bottled water for use by children younger than 6 months and health-impaired individuals. Bottled water must be distributed until four (4) consecutive weekly tests for nitrate measure below the 8.5 mg $NO_3^-$ as N/L set point. Violations above 10 mg $NO_3^-$ as N/L on a frequent basis require compliance plans to be submitted to the IEPA for meeting the MCL.

During the 1997-1998 fiscal year, Decatur Water Production distributed over 4500 gallons (17,000 L) of nitrate free water (Lamarr, 1998). Surveys completed by AWWA and by the USEPA indicate that over 23% of primary MCL excursions were caused by nitrate in 1985, with up to 1,130 public and 250,000 private potable water sources in noncompliance in 1990 (Kapoor and Viraraghavan, 1997). Future protection of source waters through state agencies may reduce nitrate pollution by implementing watershed management techniques.

## Available Removal Technologies

Physiochemical treatment methods to remove or reduce nitrate can be implemented. Anionic ion exchange (IX), biological denitrification, and RO represent the three recommended technologies for full scale water treatment. As described by Kapoor and Viaraghavan (1997), other removal methods include, and are not limited to, chemical denitrification, catalytical reduction, and electrodialysis.

Chemical denitrification and catalytic reduction both utilize elemental metals to strip oxygen from the nitrogenous compound. Catalytical reduction is a relatively new process that requires further study and may be cost effective for small treatment plants. Results from chemical denitrification indicate that ferrous hydroxide or aluminum can be oxidized by nitrate to release ammonia and nitrogen gas. Electrodialysis uses electrical charges to attract and separate the contaminant across a semipermeable charged membrane (Schoeman and Thompson, 1996).

Mainstream use of ion exchange, biological denitrification and RO is beginning to find acceptance and support from state agencies and other groups. Ion exchange uses cationic and anionic resins to remove dissolved solids from the waste stream. Anionic resins exchange chloride ions with nitrate ions. Once breakthrough of the nitrate occurs within an IX column, regeneration with sodium chloride brine is completed to remove the absorbed nitrate to produce a nitrate-laden brine waste. Studies in Tulare, California by EPRI are currently evaluating the use of an additional electrochemical process to reduce the nitrate present in the waste (Lamarre, 1998). The overall recovery for IX can be as high as 97%.

Biological denitrification of potable water was derived from experience with wastewater denitrification. The use of bacteria under anaerobic conditions leads to the degradation of nitrate to ammonia and nitrogen gas. Because most fresh water sources are low in carbon content, carbon substrates are added to encourage bacterial growth. The substrate can be a simple as corn syrup or ethanol (Lamarre, 1998). After contacting the water with the denitrifying columns, the water is decontaminated further with filters to remove the substrate and bacteria.

RO is designated by the USEPA as the best available treatment technology (BAT) for the removal of nitrates. RO represents a non-intrusive method to remove nitrate from water sources. Selection of the membrane material is important for nitrate removal. Newer membrane products featuring PA-TFC technology allow for the formation of high rejection membranes (99.4% salt rejection) for low-pressure operation, 200-300 psi (1379-2068 kPa). These membranes promise to offer the best overall nitrate rejections in this pilot study. Softening NF and energy conserving membranes have lower rejections of nitrate. Table 3-1 compares all three technologies.

**Table 3-1**
**Summary of Ion Exchange, Biological Denitrification, and Reverse Osmosis for Nitrate Removal from Drinking Water (modified from Kapoor and Viraraghavan, 1997)**

| Aspect | Ion Exchange | Biological Denitrification | Reverse Osmosis |
|---|---|---|---|
| Status of the process | Proven on full scale | Proven on full scale | Proven on full scale but not for nitrate removal in specific |
| Start-up period | Minutes | Up to three (3) weeks | Minutes |
| Ability to automate | Good | Partial | Good |
| Sensitivity of operation to temperature | None | Lower limit of 2-6°C (36-43°F) | None |
| Generated waste | Brine regenerate | Possible biomass | High TDS retentate |
| Posttreatment | Stabilization of corrosive product water | Filtration of microorganisms and substrate | Blending: virtually none<br>Straight: pH adjustment; recorbonation |
| Operation | Stable | Stable | Stable |
| Sensitivity to presence of: | Sulfates, NOM, and chloride | Dissolved oxygen | NOM and TDS |
| Nitrate removal | 99% (EPRI) | 85% (University of Colorado) | Dependent on membrane type, pressure, and feed solution.<br>High Rejection RO 90-98.5% (This study - University of Illinois) |

## *Pesticides—Atrazine*

### Background

The application of pesticides is a great concern because concentrations in the range of parts per billion in contaminated waters can induce chronic toxicological and carcinogenic effects. The contamination of sub-surface and surface water sources by atrazine, herbicides, and pesticides is common in agriculture. In order to protect crops from devastating pests, over 300 million kilograms of pesticides per year were applied on Midwestern corn and soybean crops in the late 1980's (Schotter et al., 1994).

With Illinois containing the largest area of drained cropland in the United States, pesticide contamination is a concern (Fausey et al., 1995).

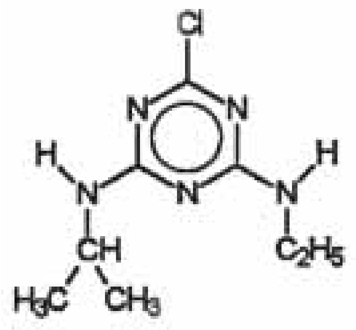

**Figure 3-6**
**Chemical Structure for Atrazine**

Cyanazine, simazine and atrazine are a group of pesticides known as triazines. Atrazine remains the most extensively used pesticide. After application, about 1% of atrazine leaches into watershed drainage (Laroche et al., 1996). Many watchdog groups and regulatory agencies cite high levels of multiple pesticide contaminants within several Illinois water supplies, including Danville and Decatur.

Atrazine, as shown in Figure 3-6, is an organic compound that exists as a colorless powder. As a herbicide, the substance is sprayed onto crops to control grassy weeds and other vegetation. For combined fertilizer/pesticide application, the triazine herbicide can be applied with nitrogen in the early to late spring (Shell, Inc. 1990). In the soil, atrazine can be metabolized to form didealkyl-atrazine (Figure 3-7) by bacteria (Adams and Watson, 1996).

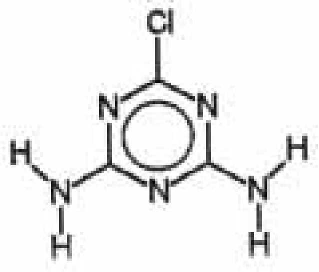

**Figure 3-7**
**Chemical Structure for Didealkyl-Atrazine**

While the health effects of atrazine are not as well known, the USEPA Office of Pesticide Programs recognizes atrazine as a possible carcinogen. The SOC Contaminants - Drinking Water Standards state that atrazine may cause an increased risk in mammary gland tumors (1998). The didealkyl-atrazine metabolite is important because of the two amine groups present. As mentioned before, the amine groups could combine with nitrite to form a nitrosamine compound. Other research has shown that atrazine causes chromosomal breaks in Hamster cells which could be relevant to human carcinomas (Newman, 1995).

Other pesticides and herbicides pose unknown threats to public health. Atrazine was chosen as the compound of study because of its known presence in the raw water. Other pesticide levels in Lake Decatur are normally below water quality limits throughout the year (Smith, 1998).

### Regulations

Beginning in 1986, the USEPA placed Atrazine under regulation as part of the SDWA. The current MCL for atrazine is exceeded if concentrations are above 3 ppb for four consecutive quarters. After application of pesticides and anhydrous ammonia in the spring months, rainfall will wash the pesticides through the watershed. For a period of three to four weeks, surface water treatment plants affected by agricultural run off may measure atrazine levels above 3 ppb in their raw water. However, without further pesticide application, pesticide levels are near or below detection limits, for the remainder of the year.

### Available Removal Technologies

The USEPA currently recommends activated carbon as the best available treatment (BAT) technology for the removal of atrazine. Conventional lime and multi-media filtration treatment processes cannot remove atrazine from source waters. With PAC injected into the early stages of the treatment process, long contact times can be achieved to effectively remove atrazine (Adham et al., 1994). The atrazine levels are normally below 0.3 ppb but can go up as high as 5.0 to 7.0 ppb in early to late spring after pesticides are applied to the fields. In combination with PAC, the UF process can be used to filter the atrazine contaminated PAC and maintain better membrane performance (Adham, 1993). Decatur, Illinois has not exceeded the MCL for atrazine and continues dosing PAC at levels of 10-12 mg/L. The atrazine regulation will be re-evaluated in 1999.

A much larger form of activated carbon, granulated activated carbon (GAC), achieves similar removals of atrazine using contact columns. With regular regeneration at periods of one or two months, GAC columns can adsorb organics and pesticides similar to PAC.

Ozone and ultraviolet (UV) radiation can breakdown atrazine by oxidation. However, the oxidation of triazines by UV and ozone produce byproducts which are undesirable (Adams and Watson, 1996).

With or without PAC, membrane technologies can also remove pesticides and herbicides. Under normal circumstances, NF and RO membranes with a low molecular weight cut off (<200 Daltons) can remove pesticides to a great extent. Some membrane elements may slough off certain pesticides after a few days of operation leading to negative rejection; where concentrations of pesticide in the permeate can be higher than the feed (Buckley and Hurt, 1996). Cellulose Acetate RO elements have shown poor removals of organic compounds in previous tests. In other studies, TFC type RO elements have shown that the pesticide alachlor can be removed below detectable limits without absorption onto the membrane (Buckley and Hurt, 1996).

## *Regulations Affecting Membrane Processes*

### Water Quality Production

Membrane filtration can be implemented to remove more than nitrates and atrazine. The secondary benefits from removing NOM, heavy metals, and pathological agents using UF and/or RO aids in meeting guidelines set by the SDWA. Since the promulgation of the SDWA in 1974, the USEPA enforces and regulates drinking water. Further amendments to the SDWA were completed in 1986 and in 1996 to strengthen pollution prevention and regulatory issues by providing funds for treatment plant modernization (Harms, 1998).

Due to an increased need for better water quality, recent regulations may promote the use of membrane filtration as the “best available technology” for certain contaminants. The following represent important regulations that could further impact the use and implementation of membrane filtration: microbial/disinfection by-products (M/DBP) agreement in principle, disinfectants/disinfection by-products rule (D/DBPR), the interim and long term enhanced surface water treatment rule (IESWTR and LTESWTR), and the information collection rule (ICR). While agreements and regulations under the ICR and M/DBP are currently in action, future proposals and final rulings on the D/DBPR, IESWTR, and LTESWTR legislation will determine the future of membrane systems in water processes.

### The Information Collection Rule (ICR)

The Information Collection Rule requires the evaluation of NF and RO membranes and GAC columns with bench and pilot scale tests. The ICR became effective in July 1997. To a great extent, the ICR implements the collection of data relating to the current removals of DBPs and microbial agents. Combined with results from the RO and GAC technologies, removal efficiencies for all water treatment processes can be compared. The results from the various data collections will then be used as part of the Stage II - D/DBPR and the LTESWTR.

In order to implement the ICR, seven classifications were used to select numerous water treatment plants to obtain treatment information. Populations above 100,000 served by surface water and 50,000 served by sub-surface water are included in the ICR studies. Monitoring of DBPs and *Cryptosporidium* in the treatment process was facilitated at some sites. Based on the individual water quality parameters, other sites required either NF/RO or GAC bench or pilot studies to remove DBPs and other contaminants (USEPA, 1996).

### The Microbial/Disinfection By-Products (M/DBP) Agreement in Principle

The M/DBP agreement represents a legally binding document that establishes deadlines for the promulgation of both the IESWTR and the Stage I - D/DBPR. The main points of the agreement included consideration of DBP MCLs, enhanced coagulation, turbidity, and multiple barriers for *Cryptosporidium*.

The M/DPBP agreement recommended that the MCLs for DBPs remain at the levels proposed under the proposed stage I - D/DBPR. These levels are 0.080 mg TTHMs/L, 0.060 mg HAA5/L and 0.010 mg bromate/L.

From the turbidity stand point, treatment systems must now meet turbidity measurements below 0.3 NTU more than 95% of the time. This decrease in final turbidity from 0.5 NTU will further limit the possibility of bacteria and virus contamination.

Removal of both turbidity and DBPs with membranes could also be implemented as part of the multiple barrier approach for water treatment under the D/DBP rules. As part of proposals for the LTESWTR, the M/DBP document also suggests that data collection, technology capabilities, and cost considerations are accounted for in establishing the final LTESWTR.

### The Disinfectants/Disinfection By-Product Rule (D/DBPR)

The D/DBPR is divided into two stages for the reduction of DBPs in product water. The first stage should set the MCLs for total trihalomethanes (TTHMs), five haloacetic acids (HAA5), and bromate as recommended in the M/DBP agreement. Stage II will lower the MCLs for the aforementioned contaminants by fifty percent. Upon initiation of the D/DBPR, the first stage will be promulgated in November, 1998. The D/DBPR second stage promulgation will not occur until after the year 2000. Under both stages, regulated disinfection by-products will include chloroform, bromoform, bromodichloromethane, dibromochloromethane, dichloroacetic acid, trichloroacetic acid, chlorite, bromate, chloral hydrate, chlorine chloramines, and chlorine dioxide (USEPA, 1996). Most of these compounds are of concern because of risks associated with cancer development.

### The Interim and Long Term Enhanced Surface Water Treatment Rule (IESWTR and LTESWTR)

Both IESWTR and LTESWTR will focus on microbial decontamination by turbidity removal and disinfection. With lower DBP expectations, the evaluation of contact times with a few disinfectants (ozone, chlorine dioxide) will be reevaluated to form new criteria for disinfection credits. Information provided by the ICR will aid in the determination of disinfection abilities from the water treatment plants that participated in the study. Finished water quality of 0.3 NTU or better will finally apply to 95% of monthly measurements with a maximum of 1 NTU at any time throughout the month.

Membranes can be used as a non-intrusive method for disinfection. Without imperfections or damages to the membrane surface, uncompromised membranes are able to reach 99.20% to 100.00% removal of pathogens. With greater than 1 log removal, this solution is comparable to chemical disinfection. Imperfections incurred during the membrane manufacture can compromise these removal rates. However, if combined with other forms of filtration, multiple barriers could then be present that would overcome the effect of surface imperfections (Buckley and Hurt, 1996).

### State Alternative Technology Approval Protocol (SATAP)

As part of state guidelines, alternative treatment technologies may require additional information and case studies before approval is granted to design and construct full-scale facilities. As an aid to state EPA branches, the association of state drinking water administrators (ASDWA) and the USEPA released the state alternative technology approval protocol (SATAP). This protocol allows states to identify problems in conventional water treatment while determining available alternatives that are compatible with existing facilities. Using information exchanges on technology proven in other states, approval for such technologies can be more reliably based on pilot studies or laboratory data.

## Membrane Residual Disposal

As with most treatment processes that remove unwanted contaminants, membranes generate a concentrated waste that must be disposed of according to applicable local, state, and federal laws. Membrane systems, in particular, produce a waste that is typically higher in suspended and dissolved solids depending on the rate of recovery and solids rejection. For UF processes, waste production can occur during filtration when solids are being deconcentrated or with filter backwash. The multiple stage NF/RO process waste is continuous and forms a retentate with TDS almost a magnitude higher than in the feed water. Additional waste treatment may be required before the release of retentate wastes. For a given membrane process with 80%

recovery, over 2 MGD (7,600 $m^3$/d) of retentate may be produced for every 8 MGD (30,400 $m^3$/d) of permeate water produced.

On another note, the residuals from membrane cleanings with caustic and acidic solutions may also require additional permits for waste storage and disposal. While volumes of cleaning wastes can be minimized, they oftentimes contain enzymes, phosphorous, organic acids, and caustics which may not be desirable in a municipal waste stream.

### Applicable Legislation

Permitting issues for membrane retentate span a number of applicable federal laws. Pontius et al. (1996) recognized the following as possible federal laws that may apply to the disposal of retentate waters:

- Clean Air Act (CAA)
- Clean Water Act (CWA)
- Coastal Zone Management Act (CZMA)
- Comprehensive Environmental Response, Compensation, and Liability Act (CERCLA)
- Hazardous Materials Transportation Act (HMTA)
- Occupational Safety and Health Act (OSHA)
- Resource Conservation and Recovery Act (RCRA)
- Safe Drinking Water Act (SDWA)
- Solid Waste Disposal Act (SoWDA)
- Toxic Substances Control Act (ToSCA)

While not all of these laws may be applicable, the ones that have the most influence over retentate disposal in the Midwest include the CWA, CERCLA, OSHA, SDWA, ToSCA, and RCRA. The CWA requires an NPDES permit for direct or indirect discharge of pollutants to navigable waters from publically owned treatment works (POTW). CERCLA may not be enforceable unless the retentate or cleaning water contains known hazardous substances (i.e., atrazine) above reportable quantities. OSHA ensures that workers are not exposed to hazards during employment. The safe drinking water act (SDWA) disallows the practice of underground injection if wellheads are subject to contamination as regulated by the underground injection control (UIC) rule. The UIC regulations are implemented by state agencies and normally classify wastes for injection to certain wells based on contaminate concentrations. Under the toxic substances control act, the sale of treatment residuals may require premanufacturing notification and other actions to be undertaken. Finally, RCRA uses the cradle-to-grave approach for the generation, storage, transport, treatment, and disposal of hazardous wastes (Subtitle C) and non-hazardous wastes (Subtitle D).

Future regulation by the USEPA will include the filter backwash water rule (FBWR). While proposals are not due until August of 1999, this rule will regulate the recycle of filter backwash water within public treatment facilities. UF processes may fall under this rule because of the backwash operations used to clean the membrane. Separate treatment and disposal of backwash waste may be a requirement.

### Disposal Methods

The method of disposal for membrane residuals will determine which regulations must be met. Several options can be considered for the disposal of membrane wastes. These include (Pontius et al., 1996):

- Discharge into Surface Water
- Discharge to Wastewater Treatment Plant
- Land Application
  - Irrigation
  - Infiltration/Percolation
- Underground Well Injection
- Ponds or Lagoons

Due to the high salt concentration within RO water, the method of disposal should be chosen to minimize the impact on the environment. Also, if at all possible, receiving waters should not be polluted further by concentrate disposal. The quantity of retentate will determine the cost associated with each method and the waste disposal. From previous designs, Tables 3-2 and 3-3 describe how disposal methods were employed for different sized treatment plants for MF/UF and RO treatment plants.

**Table 3-2**
**RO Concentration Disposal Technique Rated by Production,**
***Water Processes Membrane Treatment* (1996)**

| **Disposal** | **<0.3 MGD** | **0.3-1 MGD** | **1-3 MGD** | **>3 MGD** | **Total** |
|---|---|---|---|---|---|
| Surface | 34 | 12 | 9 | 11 | 66 |
| Land Application* | 14 | 2 | 0 | 1 | 17 |
| Sewer | 18 | 8 | 3 | 3 | 32 |
| Deep Well Injection | 3 | 1 | 5 | 5 | 14 |
| Evaporation Pond | 8 | 0 | 0 | 0 | 8 |
| Total | 77 | 23 | 17 | 20 | 137 |

*None planned in future

**Table 3-3**
**MF and UF Disposal Methods of Backwash and Chemical Residuals Based from 107 Survey Responses (Adham et al., 1996)**

| **Disposal** | **Backwash (63)** | **Chemical (44)** | **Total (107)** |
|---|---|---|---|
| Recycled to Source | 37% (23) | 28% (12) | 33% (35) |
| Sewers | 25% (16) | 70% (31) | 44.3% (47) |
| Reservoirs | 10% (6) | 2% (1) | 6.6% (7) |
| Rivers | 18% (11) | 0% (0) | 10.4% (11) |
| Irrigation | 10% (6) | 0% (0) | 5.7% (6) |
| Total | 100% (62) | 100% (44) | 100% (106) |

* ( ) = Number

Surface discharge of treatment wastes is the most common application of retentate discharge and usually can be done in cases where land is inexpensive. Ponds, canals, and lagoons can hold designed volumes of retentate for retention or detention purposes. Biological treatment could possibly be used to degrade contaminants before release back into streams or lakes. When considering UF retentate treatment, lagoons could be used prior to discharge to provide sedimentation of suspended solids. Reservoirs and rivers near treatment plants could also be considered for UF backwash water.

Land application could be accomplished with low TDS concentrations (1000-2000 mg/L). While future treatment plants were not planning land application of RO wastes, marketing of low TDS RO retentate as "grey" water for non-potable use could provide cheap or free bulk water for farmers to irrigate crops. Atrazine residuals would be fairly low (<1 ppb) with high nitrates (60-100 mg/L) in the retentate water (700-1100 mg/L TDS) for eight to ten months out of the year if low pressure RO were to be installed in Decatur, Illinois. Winter months and an abundance of rainfall would short-circuit this disposal method.

When considering wasting to sewers, wastewater treatment plant capacity may influence decisions. For a 10-15 MGD (38,000 to 57,000 $m^3$/d) RO treatment plant, an additional 2-3 MGD (7,600 to 11,400 $m^3$/d) waste capacity may be difficult for waste facilities already near design capacity. However, wasting a portion of high nitrate waste into the sewer may be beneficial to wastewater treatment plants (WWTP) during summer months to serve as an oxygen source for aerobic treatment when higher water temperatures limit oxygen saturation (Brownell, 1998). Adjustments to the wastewater treatment plant may be required if large quantities of retentate are to be dealt with.

If other alternatives fail to provide a way to dispose of membrane concentrates, deep well injection could work if contamination of confined aquifers is avoided. Additional pumps, well casings, and permits would be necessary to force water into the substrata.

## Facilities Management Through Remote Operation

Besides investigating the membrane treatment of Lake Decatur water, computerized operation of membrane filtration at remote distances was another aspect of this project. During the 1980's, concern with the acquisition and maintenance of facility features, combined with the emergence of computerized graphical technologies, led to the development of an application termed facility management (FM). In the infant stages, FM appeared as a specialized application of computer-aided design (CAD) drawings in combination with database management systems (DBMS). Automated mapping and on-line data were later integrated with DBMS to automate the graphical display of information in a spatial environment (Hackman and Koch, 1998).

With the application of computers and automation hardware/software, facilities management (FM) through geological information systems (GIS) is a reality. Municipalities are able to utilize the capabilities of GIS to 1) support the planning and delivery of maintenance to road, sewer, water, and cable facilities; 2) update, display, and analyze facilities data; and 3) plan for facilities expansion (Aronoff, 1995). Water and wastewater professionals can also take advantage of GIS's faculties for FM, particularly in the areas of operation and maintenance, planning and engineering, and finance and administration (ESRI). The extent of information that can included ranges from watershed management, wellhead protection, treatment works, and distribution systems. A summary of current GIS technologies applicable to water and wastewater treatment includes:

- Automated Mapping
- Automated Meter Reading
- Business Systems
- Connection Review/Approval
- Disaster Preparedness, Emergency Response
- Laboratory Information Management System (LIMS)
- Maintenance Management Systems (MMS)
- Maps/Records Management
- Planning/Forecasting
- Public Information, Presentation Mapping
- Spill/Wastewater Control Systems
- Supervisory Control and Data Acquisition (SCADA)
- System Modeling

Due to the nature of the UF and RO processes, the use of FM for SCADA and other management related piloting issues could be accomplished. Newer versions of LabView software from National Instruments include the capacity of HTML and JAVA language programming for active internet access (1998). Data transfer rates between two computers can be increased by using faster 58.8 kbaud modems or T1 (ethernet) type computer connections. Data files, written by automation software and transported via ftp or kermit protocols, can be written in numerous formats including Microsoft Excel Spreadsheets. Standardization of file formatting would allow data from multiple sites to be easily compiled. Video streams can also serve as a way not only to view the pilot at work, but also to take images to estimate parameter values from gauges and instrumentation without the need for a data file.

A few difficulties in a fully automated system do arise in the ability to calibrate sensors and refill feed tanks. In order to assure proper operation of sensors, it is necessary to have operators calibrate transducers, replace sample tubing, and conduct maintenance on pumps, piping, and electrical systems. Feed tanks require filling over time to keep the system within operation specifications. System shutdowns and restarts of any portion of a UF or RO pilot usually require the presence of an operator to prevent damage to the membrane or system due to lack of feed from clogged filters, stuck valves, or the need for additional cleaning. However, with the remote operation capabilities, changes in weather, feed water quality, and system performance can permit modifications in real time to meet future demands.

## Full Scale Membrane Facilities

While there are a few studies that consider cost of full scale systems, over 140 membrane treatment plants across the United States are in operation. The data gathered from those treatment plants in operation have been used to generate basic equations for cost estimating. While research with bench and pilot scale systems may allow for an estimation of operational characteristics, it is difficult to extrapolate full scale costs from such tests. The novelty of membrane processes has yet to wear off in the United States because of its limited use in water treatment.

### *Cost of UF Implementation—Model*

#### Cost Model Background

With the installation of full scale UF systems, cost information becomes more reliable. Pickering and Wiesner (1993) have studied 7 UF and MF low pressure membrane processes (0.1 to 6 MGD (378.5 to 22,710 $m^3/d$)) to estimate costs for larger treatment plants. Surveys compiled by Adham, Jacangelo, and Laîné (1996) from UF and MF treatment plants from 1993-1995 showed the economies of scale for increased UF production. The following equations represent parts of both models; Pickering and Wiesner look at direct costs and cost per 1000 gallons ($C_x$, $Y_Y$) and Adham et al. view cost as dollars per 1000 gallons ($Y_x$). Either can be converted to direct or product cost by using the engineered flow rate $Q_{eng}$ for comparison.

#### Capital Costs

The following equations use the total amount of membrane area and flow to determine capital cost. Membrane costs represent the purchase of elements from membrane manufacturers based on flux and flow designs. Non-membrane costs represent the pipes, valves, instrumentation, pumps, and buildings required to make the elements functional. Equations 3.21 and 3.22 represent the cost of water production in 1000 gallon increments.

Membrane Area—$A_{mem}$:

$$A_{mem} = \frac{Q_{eng} \cdot t_{tot}}{J_s \cdot t_p - J_{bw} \cdot t_{bw}} \qquad (3.15)$$

Where:

$Q_{eng}$ = Engineered Flow Rate
$t_{tot}$ = Total Time per Filtration Cycle
$t_p$ = Time for Permeate Production
$t_{bw}$ = Time for Backwash Procedure
$J_s$ = Solute Flux
$J_{bw}$ = Backwash Flux

Number of Elements—$n_{ele}$:

$$n_{ele} = INT\left(\frac{A_{mem}}{A_{ele}} + 0.5\right) \qquad (3.16)$$

Where:

$A_{ele}$ = Area of Element

Membrane Capital Cost—$C_M$ (Pickering and Wiesner), $C_{MS}$ (Adham et al):

$$C_M = C_{ele} \cdot n_{ele} \qquad (3.17)$$

Where:

$C_{ele}$ = Cost per Element

or:

$$C_{MS} = 0.78 \cdot Q_{eng}^{-0.38} \qquad (3.18)$$

Non-Membrane Capital Cost—$C_P$ (Pickering and Wiesner), $C_{PC}$ (Adham et al):

$$C_P = 150{,}037.56 \cdot (n_{ele})^{0.74} \qquad (3.19)$$

or:

$$C_{PC} = 1.29 \cdot Q_{eng}^{-0.40} \qquad (3.20)$$

Total Capital Cost—$Y_T$ (Pickering and Wiesner) ,$Y_{TC}$ (Adham et al):

$$Y_T = \left[ \frac{(C_M + C_P)}{Q_{eng}} \cdot \left( \frac{A}{P} \right) \right] \tag{3.21}$$

Where:

$\frac{A}{P}$ = amortization factor

or:

$$Y_{TC} = \left[ \frac{(C_{MS} + C_{PC})}{Q_{eng}} \right] \cdot \left( \frac{A}{P} \right) \tag{3.22}$$

With a given present worth for a treatment plant, equation 3.23 is used to calculate annual payments over the design life for a plant with interest and inflation adjustments.

$$\left( \frac{A}{P} \right) = \frac{i \cdot (1+i)^{DL}}{(i+1)^{DL} - 1} \tag{3.23}$$

Where:

$DL$ = Design Life
$i$ = interest rate

### Energy Cost

The cost to run the UF system pumps is representative of the energy required *(E)* for the filtration cycle during production and backflushing. Recirculation pumps are used to create cross flow velocities that prevent cake layer formation. Backflush pumps may be required to pump backwash residuals away from the filtration plant to a reservoir. Costs for additional energy use for lighting and HVAC systems are included in O&M costs rather than the energy cost category.

Feed Water—$E_f$:

$$E_f = \frac{P_f \cdot Q_f}{\eta_f} \tag{3.24}$$

Where:

$P_f$ = Feed Pressure at Pump
$Q_f$ = Flow at Feed Pump
$\eta_f$ = Overall Feed Pump Efficiency

Retentate—$E_r$:

$$E_r = \frac{P_r \cdot Q_r}{\eta_r} \tag{3.25}$$

Where:

$P_r$ = Pressure at Retentate Pump
$Q_r$ = Flow at Retentate Pump
$\eta_r$ = Overall Retentate Pump Efficiency

Backflush—$E_{bw}$:

$$E_{bw} = \frac{P_{bw} \cdot Q_{bw}}{\eta_{bw}} \tag{3.26}$$

Where:

$P_{bw}$ = Backwash Pressure at Pump
$Q_{bw}$ = Flow at Backwash Pump
$\eta_{bw}$ = Overall Backwash Pump Efficiency

Surface Scour - Recirculation Pump—$E_{ss}$:

$$E_{ss} = \frac{\Delta P_{ss} \cdot Q_{ss}}{\eta_{ss}} \tag{3.27}$$

Where:

$P_{ss}$ = Pressure Difference at Recirculation Pump
$Q_{ss}$ = Flow at Pump
$\eta_{ss}$ = Overall Recirculation Pump Efficiency

Total Energy Cost—$C_E$:

$$C_E = C_{kWh} \cdot (E_f + E_r + E_{bw} + E_{ss}) \tag{3.28}$$

Where:

$C_{kWh}$ = Cost of Energy in kWh

### Operations and Maintenance Cost—$Y_{O\&M}$

The use of the facilities requires further energy, chemical additions, HVAC, and maintenance and parts replacement. The following equation (3.29) represents operational and maintenance costs, $Y_{O\&M}$.

$$Y_{O\&M} = 0.43 \cdot Q_{eng}^{-0.15} \tag{3.29}$$

### Disposal Costs

The disposal of UF residuals may impose additional cost based on the method chosen for disposal. Part of the estimate for the Plant Capital Cost, equation 3.22, considers the use of a wash water recovery system. The Operation and Maintenance Cost, equation 3.29, includes residual disposal although no specific method was mentioned (Adham et al., 1996).

### Total Treatment Cost

Equation 3.30, below in the Figure 3-8, summarizes the cost of full scale UF based on four modern UF plants (Adham et al., 1996). As depicted in Figure 3-8, the total costs are broken down into three main components: membrane system, non-membrane system, and operations and maintenance for an entire UF processing plant. Amortization of plant costs was calculated at 7% interest and 20 year design life.

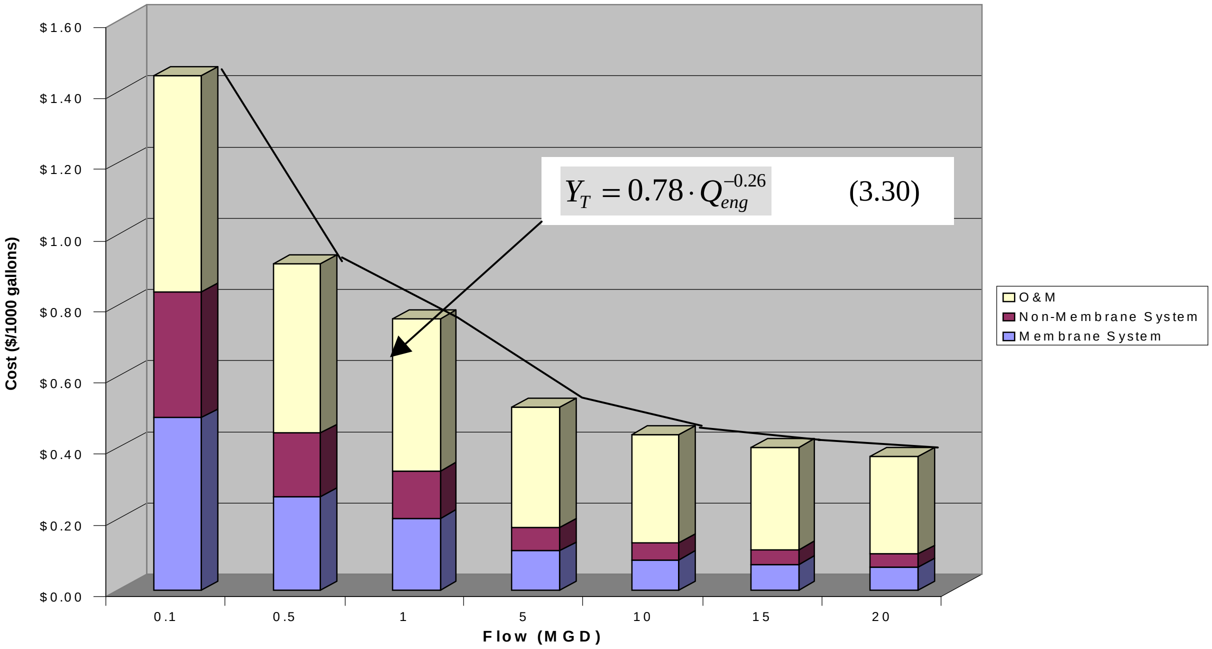

**Figure 3-8**
**Predicted Cost with UF Capacity—I = 7%, DL = 20 years and based on 1993-95 prices (Adham et al., 1996)**
**Note: $/1000 gal x 0.264 = $\$/m^3$**

## *Cost of NF/RO Implementation—Model*

### Manufacturer's Computer Models

Membrane manufacturers have developed simulation software for design of their products. Two separate computer programs were obtained for evaluation during the course of research to identify membrane products and determine possible operating costs for separate conditions.

The ROSA software from Dow Chemical, Inc. represents membranes produced by Filmtec, Inc. Osmonics-Desal, Inc. produces the Winflows software to evaluate their RO membrane products. As part of their analysis, auto design features allow a user to input flows, recoveries, and specific water quality parameters that allow for an appraisal of system performance. Multiple staged arrays can be graphically created along with the addition of pre- and post-treatment. Dosages of chemicals and operating costs are part of the ROSA output. Winflows offers more ability to change and update system setups through a virtual PID diagram. The effects of water temperature, quality, recovery, and pretreatment on specific manufacturer NF/RO elements are part of the final analysis that supplies possible operating data and finished water quality parameters. As a result, final outputs can then be compared with pilot testing to calibrate the results with the experimental data. While these programs supply statistics based on controlled situations and simplified analyses, they offer insight as to what may or may not be possible for NF/RO membranes at pilot and near full scale (5-10 MGD (18,925-37,850 $m^3/d$)).

## Cost Model Background

To determine an estimated cost for a full scale RO system, necessary parameters include design flow, pressure, and number of elements. Similar to UF, RO systems require capital investment (membrane and non-membrane), O&M, and energy costs. Energy costs for RO are higher than UF primarily because of an increase in energy consumption. As pressure and flow increase, costs will increase proportionally.

Unlike the UF cost model, few models exist for calculating NF/RO costs. Case studies of similar full scale treatment plants offer the best insight to future costs. Combined with a practical working knowledge of the reverse osmosis process, curve fitting can be achieved to estimate expenditures. The following equations can be used to estimate full-scale RO treatment costs.

## Capital Costs

To calculate overall system capital requirements, the required number of elements must be determined by selecting a design flow. Elements staged in arrays can be placed in parallel with one another accordingly. For ease of description, a 6-3-2 array would appear as follows in Figure 3-9 (Each pressure vessel holds 6-8 elements).

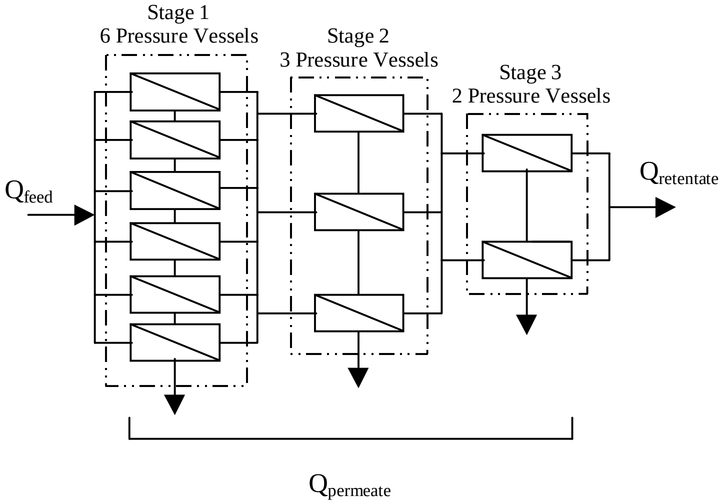

**Figure 3-9**
**Schematic of a 6-3-2 Array Membrane System**

Membrane Costs:

The flow rate per element, $Q_{ele}$, is based on equation 3.31.

$$Q_{ele} = A_{ele-RO} \cdot J_v \tag{3.31}$$

Where:

$A_{ele\text{-}RO}$ = Area per Element

The number of elements per array, $n_{arr}$, required (Equation 3.32).

$$n_{arr} = PV_{arr} \cdot n_{ele/PV} \tag{3.32}$$

Where:

$PV_{arr}$ = Pressure Vessels per Array
$n_{ele/PV}$ = Number of Elements per Array

The number of arrays required for an engineered system flow (Equation 3.33).

$$N_{arr} = INT\left(\frac{Q_{eng}}{n_{arr} \cdot Q_{ele}}\right) \tag{3.33}$$

Capital cost of all RO membrane elements and vessels (Equation 3.34).

$$C_M = N_{arr}[(c_{ele} \cdot n_{arr}) + (c_{PV} \cdot n_{PV})] \tag{3.34}$$

Where:

$c_{ele}$ = Price per Element
$c_{PV}$ = Price per Pressure Vessel

Non-membrane Cost:

The non-membrane costs, $C_{NM}$, are based on work by Gumeran et al. (1984) as presented by Wiesner and Sethi (1996). The following formula represents the necessary equipment, construction, and instrumentation required for RO operation.

$$C_{NM} = 96.75 \cdot (Q \cdot h)^{0.39} + 17.64 \cdot (Q)^{0.42} + 8.15 \cdot (Q)^{0.66} + 628.09 \cdot (Q)^{0.32} + 52.16 \cdot (Q)^{0.39} + 48.84 \cdot (Q)^{0.44} + 4.14 \cdot (Q)^{0.71} \tag{3.35}$$

Where:

$h$ = Pressure
$Q$ = Design Feed Flow

Total Capital Costs:

The total capital cost of water production per 1000 gallon is as follows.

$$Y_T = \left[ \frac{(C_M + C_{NM})}{Q_{eng}} \cdot \left( \frac{A}{P} \right) \right] \tag{3.36}$$

Where:

$$\left( \frac{A}{P} \right) = \frac{i \cdot (1+i)^{DL}}{(i+1)^{DL} - 1} \tag{3.37}$$

## Energy Costs

Energy consumption for NF/RO processes often creates a high operational cost that places RO membranes out of contention with other treatment technologies. In the past, RO feed pressures for brackish water treatment required pressures from 800 to 1000 psi (5516 to 6895 kPa)and higher. As mentioned before, recent developments have allowed for lower pressure NF and RO membranes (100-250 psi (689-1,724 kPa)) that can readily filter freshwater for potable use. Depending on pump efficiency and local utility rates, energy costs are minimized with lower pressures and lower electrical rates ($/kWh). Equation 3.38 evaluates the energy cost, $C_e$, from pumping.

$$C_e = c_{kWh} \cdot \frac{P_f \cdot Q_f}{\eta_p} \tag{3.38}$$

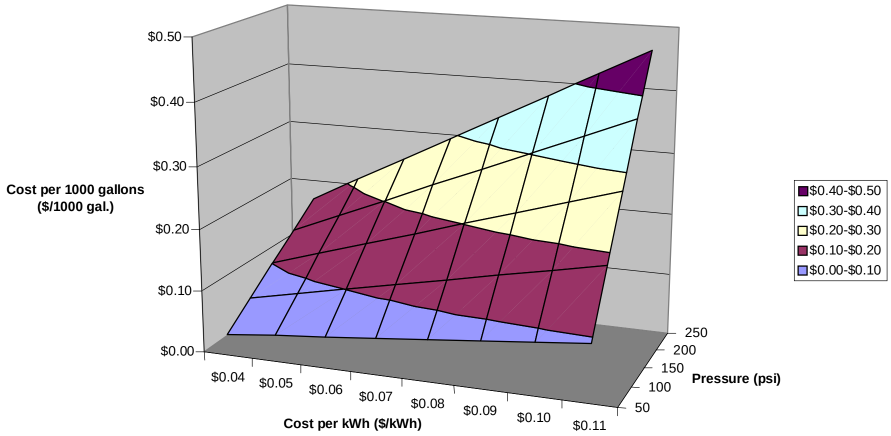

**Figure 3-10**
**Energy Cost at 75% Pump Efficiency**
**Note: $/1000 gallons x 0.264 = $/m$^3$**

## Disposal Costs

The disposal of RO retentate may require additional costs based on the method chosen for disposal. Case studies on NF waste injection wells in Florida have indicated costs for disposal in the range of $0.10 - $0.24 per 1000 gallons ($0.026 - $0.063 per cubic meter (m$^3$)) (Bergman, 1996). Depending on future design, costs for permitting and handling of wastes may be similar to conventional treatment with the exception of higher TDS and TOC concentrations (Wiesner and Sethi, 1996).

## Operations and Maintenance

Operations and maintenance requires further energy, chemical additions, HVAC, and membrane replacement. During a study of NF softening plants in Florida, Bergman identified five categories of O&M costs (1996). Figure 3-11 shows how these categories were appropriated between three 12 MGD (45,420 m$^3$/d) NF plants.

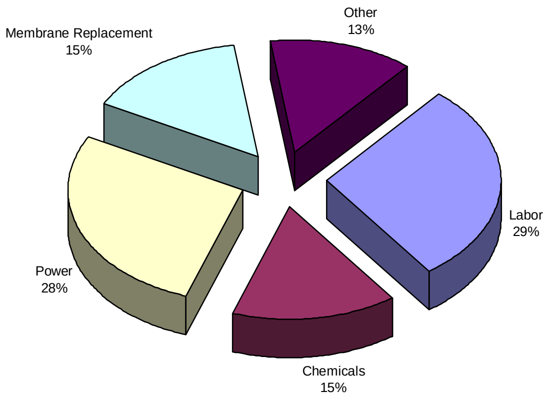

**Figure 3-11**
**O&M Appropriation**

Previous estimates of power consumption (see equation 3.38) could be included as part of the O&M expenditures. Table 3-4 includes estimates for O&M costs excluding energy costs (Bergman, 1996).

**Table 3-4**
**O&M Costs for NF/RO Treatment in Florida ($/1000 gal.) Bergman, 1996**

| Permeate Production | Labor | Chemicals | Other | Membrane Replacement | Total |
|---|---|---|---|---|---|
| 12 MGD | $0.12 | $0.10 | $0.02 | $0.09 | $0.33 |
| 12 MGD | $0.18 | $0.06 | $0.10 | $0.06 | $0.40 |
| 12 MGD | $0.15 | $0.08 | $0.09 | $0.08 | $0.40 |
| 6 MGD | $0.14 | $0.13 | $0.04 | $0.04 | $0.35 |
| 6 MGD | $0.17 | $0.19 | $0.06 | $0.10 | $0.52 |
| 4 MGD | $0.39 | $0.05 | $0.05 | $0.07 | $0.56 |

## Total Costs

By summarizing all individual component costs of the RO treatment process, full scale estimates are feasible. Final cost estimates will be affected by membrane selection, flow, and pressure. Overall, RO treatment costs have ranged from $0.55 to $1.00 per 1000 gallons (0.145 to 0.264/$m^3$) depending on system characteristics (Bergman, 1996; Taylor and Jacobs, 1996; and Furukawa and Burton, 1997).

# 4

# MATERIALS AND METHODS

## Materials

### *Raw Water Source:* Lake Decatur, *Illinois*

Lake Decatur is a shallow man made lake used for recreation and as a source of potable water. The Upper Sangamon River Watershed consists of 925 square miles (2,396 square kilometers) of land used mainly for crops like corn. This watershed drains into Lake Decatur via the Sangamon River and its tributaries (Figure 4-1). With a surface area of 2,800 acres (1,134 hectares), the lake level is maintained via a computer controlled dam which feeds water to the lower Sangamon River.

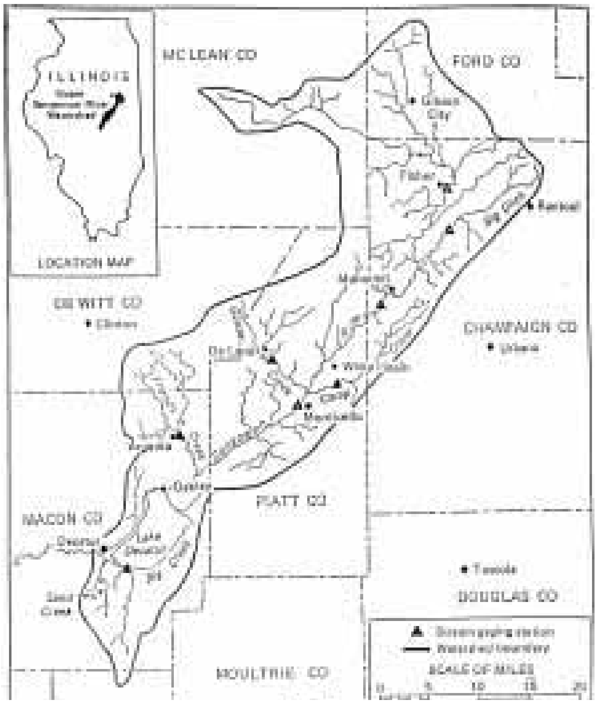

**Figure 4-1**
**Lake Decatur Watershed Annual Report—Decatur (1997)**

Within the watershed, farmers apply anhydrous ammonia, herbicides, and pesticides, which are eventually washed downstream through storm runoff, tiling, and subsurface groundwater. Rainstorm events produce a surface runoff that can dramatically influence the raw water quality of Lake Decatur. As a result of precipitation, high flow rates from the Sangamon produce a very turbid water (150-300 NTU) in combination with increases in TOC, nitrates, and (periodically) atrazine. The soft rainwater causes a decline in total hardness, an increase in ammonia concentrations (if nitrogen is recently applied) (May 1998), and a concern for the ability of the treatment plant to meet tap water turbidity standards.

Following a storm event, a portion of the rainwater percolates and carries agricultural and commercial additives into the ground. Once in the soil, biological nitrification of anhydrous ammonia produces nitrates to aid plant synthesis and growth. Tiling drains the farmfields by collecting the excess water which can contain nitrate. Slowly, this water is channeled downstream through the Decatur watershed.

Between periods of precipitation, shallow aquifers feeding into the Sangamon River sustain the lake level. The water contained in the aquifer has a higher hardness, due to the nature of the subsurface morphology, and are shallow enough to be influenced by surface supplies.

### Tap Water: City of Decatur—South Water Treatment Plant (SWTP)

The SWTP pumps raw water from an intake screen 5 feet (1.5 meters) from the bottom of the lake. As it reaches the pump house, additives such as potassium permanganate ($KMnO_4$), ammonia, PAC, and polymer coagulants can be added. The current use of PAC at doses of 5-15 mg/L lessens atrazine and TOC concentrations throughout treatment. The following table (Table 4-1) represents the characteristics of the influent to the plant.

**Table 4-1**
**Raw Water Analysis Results (Annual Report—Decatur, 1997)**

| Monthly Averages | Turbidity (NTU) | pH | Total Hardness (mg/L $CaCO_3$) | $Ca^{+2}$ Hardness (mg/L $CaCO_3$) | Non-Carbonate Hardness (mg/L $CaCO_3$) |
|---|---|---|---|---|---|
| Minimum | 14 | 7.6 | 220 | 104 | 38 |
| Average | 31 | 8.1 | 240 | 130 | 56 |
| Maximum | 76 | 8.4 | 263 | 167 | 80 |

Once pumped to the plant, the pretreated water is then lime softened in the primary basins. The softened water is then collected at the top of the basin through V-shaped weirs where it is gravity fed into the secondary tank. As lime precipitates, settling occurs in the bottom of primary basin after going through the solids contact zone. As sediment and floc settle out, the excess sludge is discharged to sludge lagoons off of the plant site. This also occurs in the secondary tank.

In the secondary tank, alum ($Al_2(SO_4)_3$) is added to continue coagulation of remaining particles and dissolved solids. The secondary basin acts similarly to the primary basins except it maintains a longer retention time to remove a majority of coagulated turbidity prior to filtration. As the clarified water is collected, it is again gravity fed to the filters.

After clarification, the influent to the multi-media sand/anthracite filters can be treated with polymer filter aids. The filters are backwashed manually with an air scour followed by a water flush until the majority of solids have disappeared. The backwash water is then stored in a reservoir and slowly added back into the raw water feed to the treatment plant. Times between backwash can be affected by water temperature, turbidity, and backwash duration. During periods of high turbidity and/or cold water temperatures, filter effluent turbidity increases due to decreased particle removal efficiency. This decrease in water treatment capability has yet to violate existing standards for turbidity and bacterial contamination.

**Table 4-2**
**Tap Water Analysis Results (Annual Report—Decatur, 1997)**

| Monthly Averages | Turbidity (NTU) | pH | Total Hardness (mg/L $CaCO_3$) | $Ca^{+2}$ Hardness (mg/L $CaCO_3$) | Non-Carbonate Hardness (mg/L $CaCO_3$) |
|---|---|---|---|---|---|
| Minimum | 0.04 | 9.5 | 110 | 47 | 55 |
| Average | 0.15 | 9.7 | 132 | 64 | 72 |
| Maximum | 0.54 | 9.9 | 154 | 102 | 92 |

As a last measure to avoid biological hazards and promote public health, chlorine or chlorine dioxide and fluoride is injected prior to water storage in the clearwell. The clearwell allows for the appropriate disinfection contact times before the water is stored in reservoirs and then pumped into the distribution system. Concentrations of fluoride and free chlorine are maintained at 1.0-1.2 mg/L. Table 4-2 provides tap water analysis results for 1997.

Total treatment time is approximately four hours. The water from the plant and distribution system is monitored on a quarterly basis by the USEPA for SOCs, VOCs, THMs, inorganic chemicals, HAAs, and *Giardia* cysts and *Cryptosporidium* oocysts. The South Water Treatment Plant consistently produces water of high quality for all monitored contaminants except nitrates (Figure 4-2). Due to the inability of treatment to remove the nitrates, the USEPA has required Decatur to have enacted a compliance plan by April 2001 to meet the MCL of 10 mg/L $NO_3$ as N.

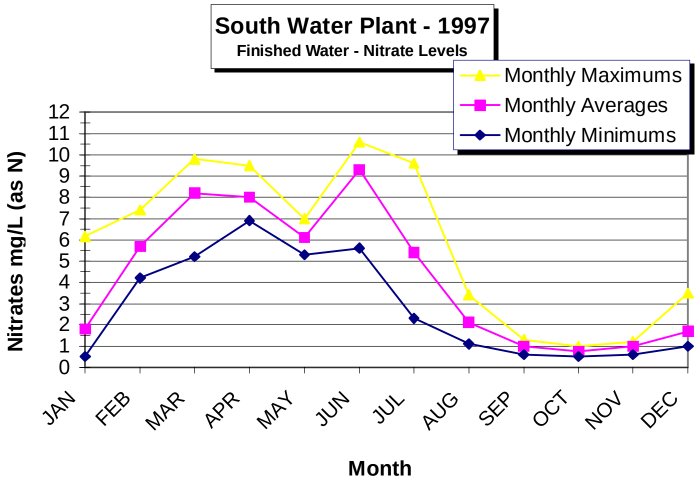

**Figure 4-2**
**SWTP Tap Water Nitrate Levels—1997 Annual Report—Decatur, Illinois**

Water quality results for the first six months of 1998 are shown on Figures 4-3 to 4-8.

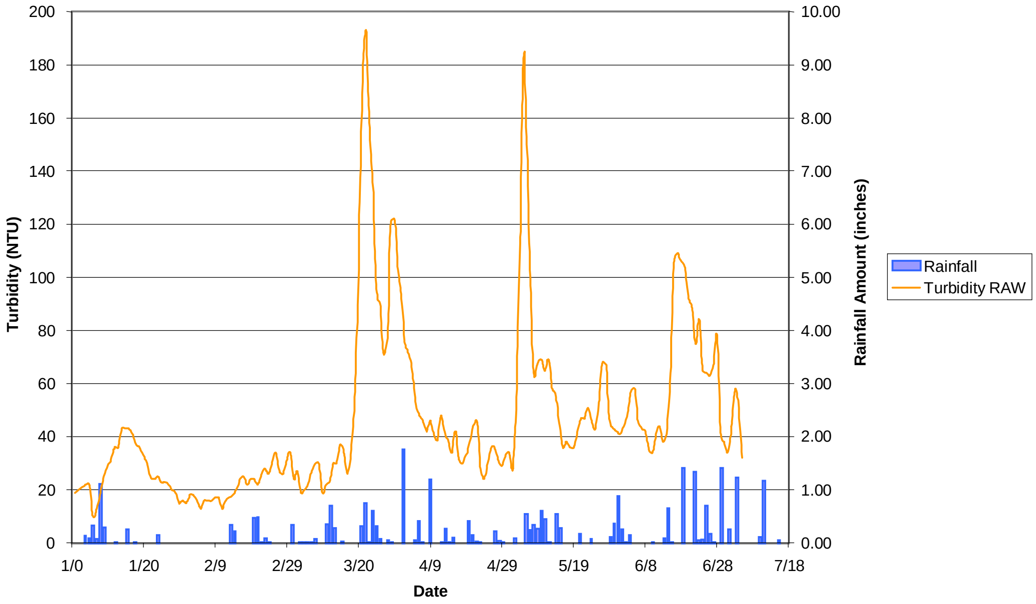

**Figure 4-3**
**Raw Water Turbidity and Rainfall for Lake Decatur, Illinois—1998**

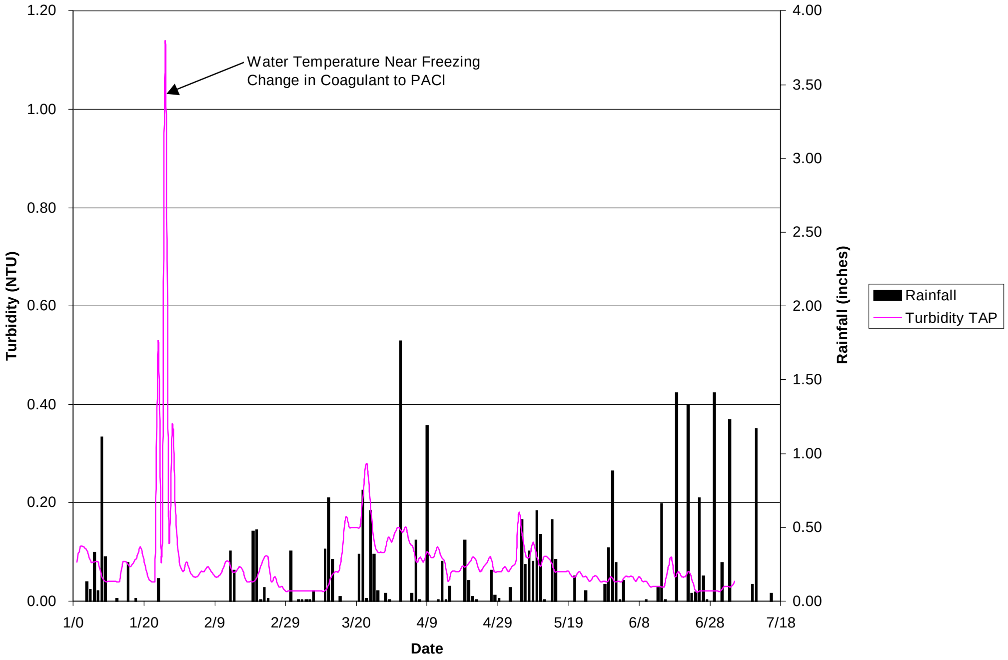

**Figure 4-4**
**Product Water Turbidity and Rainfall for SWTP, Decatur, Illinois—1998**

**Figure 4-5**
**Measured Water Temperatures for SWTP, Decatur, Illinois—1998**

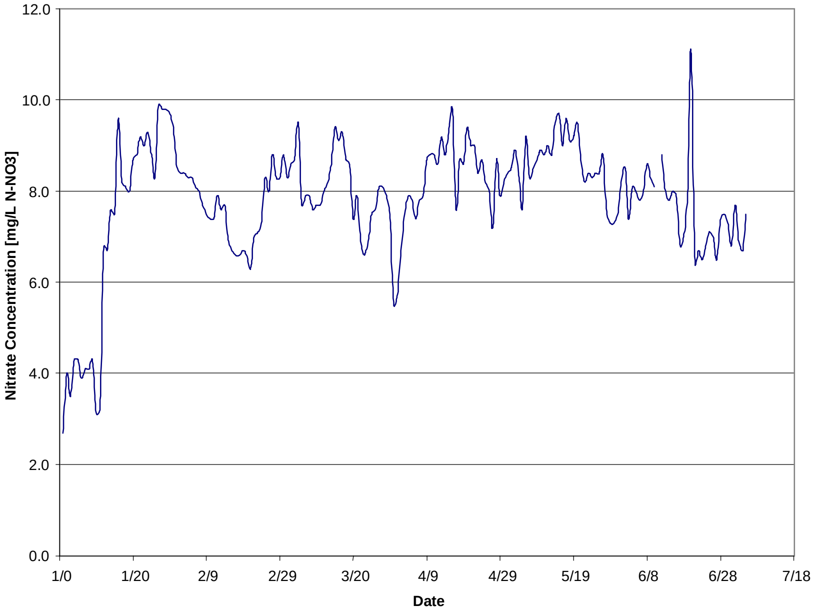

**Figure 4-6**
**Product Water Nitrate Levels for SWTP, Decatur, Illinois—1998**

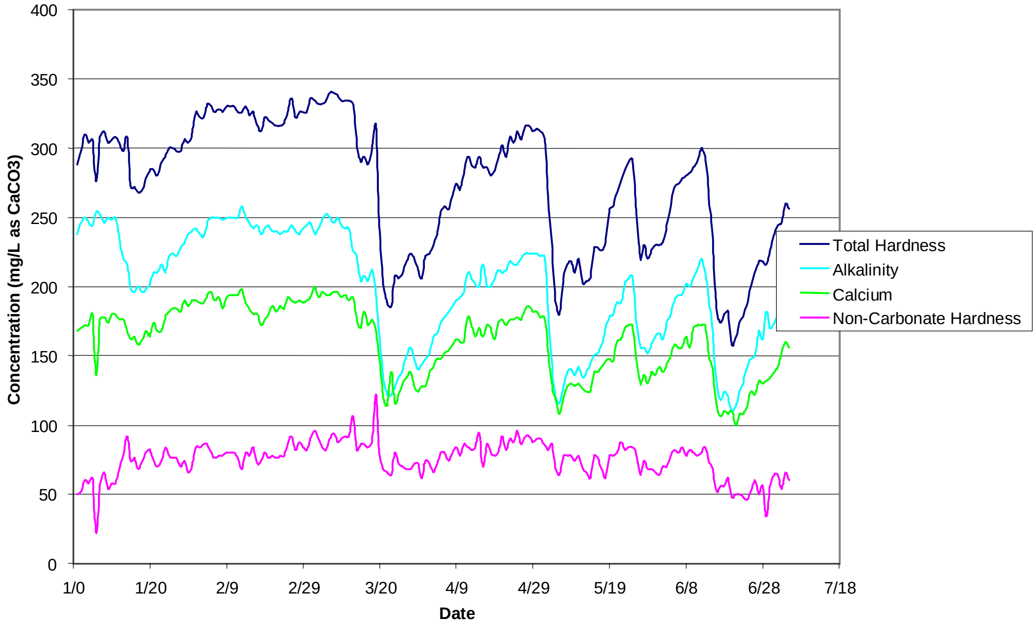

**Figure 4-7**
**Raw Water Dissolved Solids Concentrations for Lake Decatur, Illinois—1998**

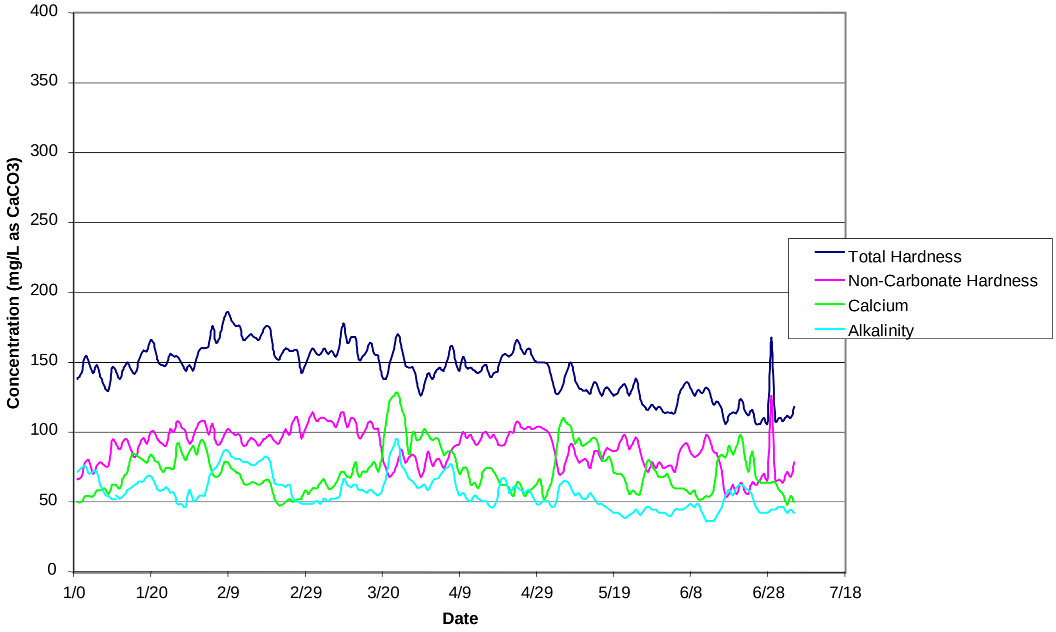

**Figure 4-8**
**Product Water Dissolved Solids Concentrations for SWTP, Decatur, Illinois—1998**

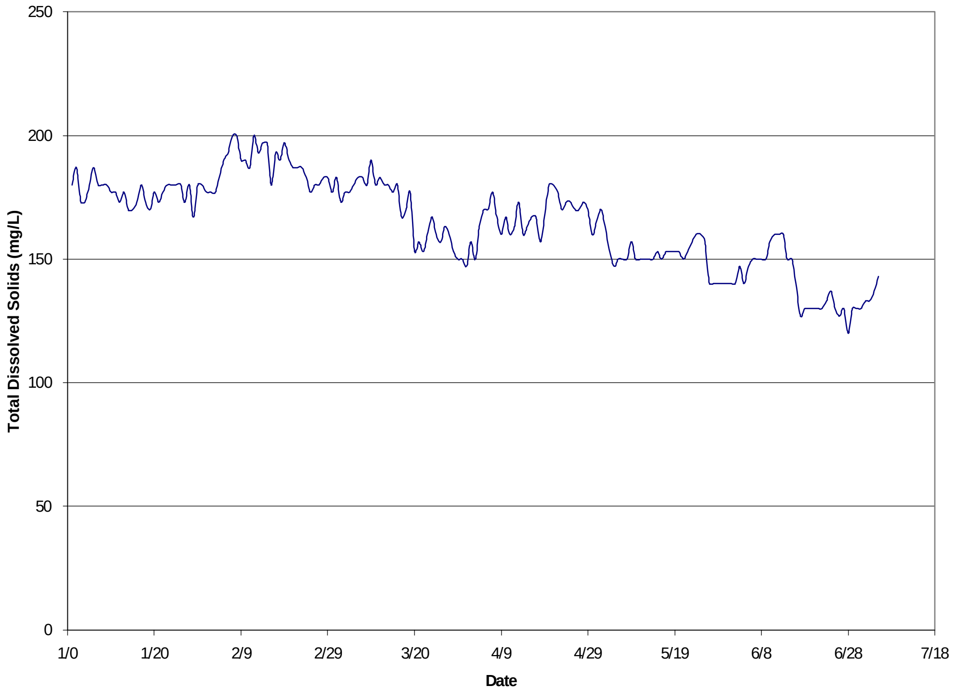

**Figure 4-9**
**Product Water Total Dissolved Solids for SWTP, Decatur, Illinois—1998**

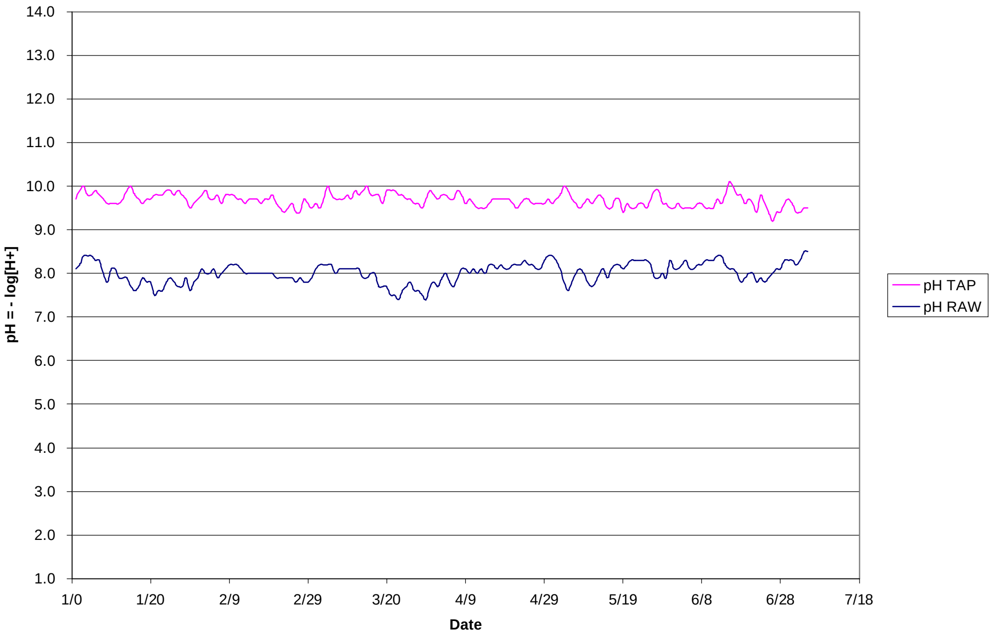

**Figure 4-10**
**Measured pH for SWTP, Decatur, Illinois—1998**

## *COMFOWARD System*

In order to meet the project requirements, the first task of the project was to put together a pilot plant capable of using membranes in a computer automated process. The second task of the project was to evaluate the system operation. This portion of the report will focus on the construction of a system that allowed Computer Operated Membrane Filtration Of Water At Remote Distances (COMFOWARD). Figures 4-11 and 4-12 show the system setup.

**Figure 4-11**
**COMFOWARD System in Decatur, Illinois**

### Water Supply and Pretreatment

#### Raw Water Pump

Raw water was pumped from a lake intake 1500 ft (457 meters) from the plant at a rate of 20 GPM (76 LPM) with a submerged pump near the inlet of the SWTP. Existing supply lines in the treatment plant were unsuitable for use because of the addition of potassium permanganate, PAC, and other coagulant aids. Unused chemical treatment lines consisting of 2 in. diameter PVC material, were attached to a pump at the raw water inlet of the SWTP. Within fifteen minutes of entering the pump, the untreated raw water arrived in the basement of the treatment plant where various methods of treatment were employed.

#### 200 Micron Pre-filter

To remove large particles and other debris prior to the UF module, a 200 micron bag filter and stainless steel housing from Rosedale, Inc (Part number EPDM #4-12-1 1/2P-

200SENB) was used. As head loss occurred from particle build up, the filter bags were removed from the housing and washed using hot tap water and hand soap.

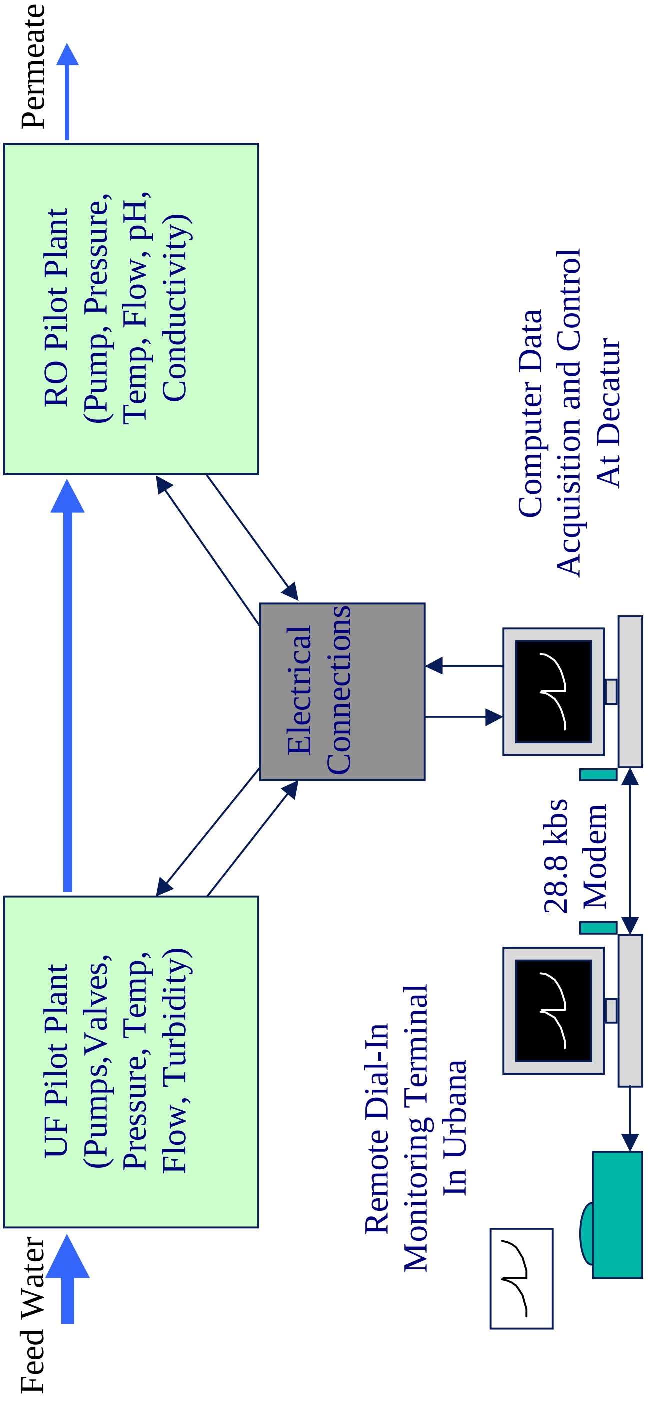

**Figure 4-12**
**Diagram of COMFOWARD Setup**

### Storage Tank

A 550 gallon (2080 L) polypropylene tank was used to store the raw water for pretreatment and further filtration. The raw water feed inlet was located opposite the raw water feed outlet. A drain valve was included to allow the raw water to drain off when the UF pilot recirculated retentate into the feed tank. As the build up of sediment occurred, the tank was periodically scrubbed with tap water.

### Mixer

As chemicals and additives were added into the feed tank, an A310 propeller (8 in. diameter) (20 cm) mounted on the top porthole of the tank was used to thoroughly mix the tank (Ducoste, 1996). Mixing speeds were adjusted according to pretreatment additive. Higher speeds of 160 to 180 rpm were used for PAC addition while lower speeds, 40-60 rpm, were used during coagulant addition. The lower speeds allowed for coagulation to occur without particle breakup, while the higher mixer speeds prevented settling of the PAC mixture.

### Pretreatment Additives

#### Pumps

Chemical feed pumps were required to add solutions of PAC, PACl, sodium metabisulfite, sodium hypochlorite, and sulfuric acid within the pilot plant. Multiple Masterflex Peristaltic Pumps with matching Gear Pump Controllers were used in combination with Tygon food and Norprene tubing (Cole-Parmer Instrument Company, Chicago, Illinois). The compatibility of fluid with the tubing material was checked using compatibility tables provided by Masterflex. The pump speed was varied from 2-200 rpm to control the addition of PAC, PACl, sodium hypochlorite, and sulfuric acid. The sodium metabisulfite pump ran at 2 rpm to provide a 1 mL/min flow. These pumps were plugged into 120 VAC receptacles controlled via the computer.

#### Powdered Activated Carbon—PAC

For all tests, the SWTP supplied a Watercarb PAC slurry. The PAC slurry was stored in a 100 gallon (376 L) container and continuously mixed using a 1800 rpm mixer (Labmaster mixer, LIGHTNIN, Rochester, New York). Data on the Watercarb PAC is given in Table 4-3.

**Table 4-3**
**Acticarb, Inc.—Watercarb PAC Specifications (Acticarb Bulletin No. 788)**

| | |
|---|---|
| Phenol Value | 18 ± 2 ppm |
| Modified Phenol Value | 1.82-2.28 g/L |
| Density | 28-32 lbs./$\text{ft}^3$<br>0.45-0.51 g/mL |
| Minimum Iodine Number | 550 Mg/g |
| % Maximum Moisture | 5% |
| % Maximum Acid Insoluble Ash | 1.5% |
| Fineness, US Sieve Series | |
| % Minimum through 100 Mesh | 99% |
| % Minimum through 200 Mesh | 95% |
| % Minimum through 325 Mesh | 90% |

## Polyaluminum Chloride

Polyaluminum chloride (PACl) was used as a coagulant to promote the formation of larger floc and improve UF membrane performance (Table 4-4). From a bulk storage tank in the basement of the SWTP, the PACl solution was used by Decatur's water plant as an aid to alum coagulation during times of high turbidity and lower than normal temperatures.

**Table 4-4**
**Westchlor® FA 900S—Basic Polyaluminum Hydroxychlorosulfate (Westwood Chemical Corp., 1997)**

| | |
|---|---|
| Aluminum | 5.80 ± 0.20 % |
| Chloride (Cl) | 10.3 ± 1.0 % |
| Basicity | 70 ± 5 % |
| Specific Gravity (25°C/25°C) | 1.240 ± 0.010 |
| pH 1% w/v aq. at 25°C | 4.3 ± 0.5 |
| Color | Translucent white to yellow |
| Turbidity | <25 NTU |
| Freezing Point | 12° ± 1°F (-11°C + 0.6°C) |
| Lot Assessment | |
| Product Lot # | 38697B |
| Aluminum Assay | 5.74 |
| Specific Gravity (25°C/25°C) | 1.2446 |

## Sodium Metabisulfite

In order to prevent damage to the membranes from free chlorine in the supplied tap water, sodium metabisulfite ($Na_2S_2O_5$—Catalog Number S244-500, NIST Traceable, ACS certified, Fisher Scientific Company) was used to reduce free chlorine to chloride. The following reaction is the presumed pathway for dechlorination.

$$S_2O_5^{-2} + 3OCl^- + 2e^- \rightarrow 2SO_4^{-2} + 3Cl^-$$

**Sulfuric Acid**
To maintain the proper pH for the membranes, concentrated 37N sulfuric acid ($H_2SO_4$) from Fisher Scientific Company (Catalog number MK-2876-212) was pumped into the treatment system prior to RO. Along with maintaining water quality within operational specifications, the acid addition aids in preventing calcium scaling on the membrane surfaces. Scaling of the membranes decreases performance and could cause irreversible damage.

## UF Pilot Plant

Two different membrane filtration techniques were assessed during the project. The first filtration technique, ultrafiltration, was used to remove particulate matter and compounds of molecular weight 100,000 Daltons and greater. The UF system was fully automated for data acquisition and system control. Jim Goodrich at the USEPA, Cincinnati, Ohio, helped secure a pre-existing system. However, when it arrived at the University of Illinois at Urbana—Champaign, the donated pilot was inoperable and needed major reconstruction for it to be fully autonomous.

### Membrane

**Specifications**
For the experiments with ultrafiltration, two hollow fiber DN100-M1A35 modules were donated by Lyonnaise de Eaux, Le Pecq, France. The modules are manufactured by Aquasource, Toulouse, France. During the test runs from December 1997 to May 1998, the F0241 DN100-M1A35 Aquasource Ultrafiltration module was used. The second UF module was kept in storage as a backup. The module was secured on the feed and retentate sides by using two 4 in. diameter compression fittings supplied by McMaster-Carr. The permeate connection was held in place by a Victaulic hydraulic compression fitting. To assure that air was purged from the membrane, the air purge lug, located at the top of the module, was removed prior to start-up of the system until the membrane housing was filled with permeate water. Table 4-5 describes the manufacturer's specifications for the modules used.

**Table 4-5**
**Aquasource DN100-M1A35 Ultrafiltration Module Specifications (Aquasource, Inc., 1996)**

| | | |
|---|---|---|
| Molecular Weight Cut-off (MWCO) | 100,000 Daltons | |
| Total Module Surface Area | 7.2 $m^2$ | |
| Approximate Number of Fibers per Module | 1120 fibers | |
| Fiber Length | 1.1 m | |
| Inside Fiber Diameter | 0.93 mm | |
| Maximum Applicable Pressure (Inside → Outside) | 29 psi | |
| Maximum Backwash Pressure (Outside → Inside) | 36 psi | |
| Maximum Free Chlorine Dose | 10 mg/L | |
| Operational pH range | 4.0 – 8.5 | |
| **M1A35 Module Inspection Test Data** | **F0235** | **F0241** |
| Approximate Maximum Permeate Flow, LPM (GPM) | 41.7 (11.0) | 38.4 (10.1) |
| Approximate Maximum Clean Water Flux (L/$m^2$-hr-bar) | 250 | 230 |

### Storage

When not in use, the membrane modules were filled with a storage solution as per manufacturer's specifications (Table 4-6). During periods of short term storage (4-5 days), the membranes were first backwashed with chlorinated permeate water. The system was then filled with tap water (pH = 8.0). With the system fully flushed and filled, the system was then closed to prevent drying.

**Table 4-6**
**Long Term Storage Solution for UF Modules (Aquasource, Inc., 1996)**

| | |
|---|---|
| Distilled water | 50% |
| Glycerol | 50% |
| Sodium metabisulfite<br>$Na_2S_2O_5$ | 1% in solution |

### Membrane Preparation and Clean Water Flux

After installation in the UF pilot, the storage solution was washed from the module following the manufacturer's directions on cleaning. The initial installation and cleaning was conducted using untreated tap water (approximate pH of 9.8). While a decrease in performance may have occurred because the recommended pH was exceeded, the membrane exposure during cleaning and filtration was limited to 24 hours. (Table 4-5 suggests that the feed water pH be maintained below 8.5).

Once purged of the storage solution, clean water flux tests were conducted to verify system performance with computerized data logging. Initial tests of the clean water flux indicated an insignificant performance loss due to the violation of the recommended pH. Data points were collected while running the system in the constant flux mode of operation. Normalized to 20°C (68°F), the data were averaged to determine the clean water flux.

### Chemical Cleaning Agents

During tests with raw water, the UF membrane required cleaning to restore clean water flux to original levels. Pressure increased very rapidly over the course of a few days which necessitated cleaning. The manufacturer's cleaning methods were followed. The system was cleaned four times during its operation with chemicals provided by Lyonnaise des Eaux. The cleaning procedure indicates the types of chemicals required. The first and third cleanings used the Ultrasil 53 detergent. The second cleaning required the use of Ultrasil 53 (detergent) and Ultrasil 49 (antiscalant). The fourth cleaning utilized all three cleaners, Ultrasil 53, Ultrasil 49, and Ferisol (iron and maganese removal), to fully regenerate the membrane. Hot tap water, 30°C (86°F), with a controlled pH of 8.0 was used to dissolve the appropriate amounts of chemical cleaner in the pretreatment feed tank. The UF pilot plant was then configured and manually run according to directions.

During backwash operation, Clorox brand regular bleach was used. This serves as a source of free chlorine (sodium hypochlorite) which is believed to clean the membrane of bacterial contamination. The Clorox was stored within the frame of the pilot in a 10 gallon (38 L) Nalgene container with cover.

### Data Acquisition and Control System

With the Aquasource Ultrafiltration DN100 module, the pilot plant design was structured to provide flexibility and reliable results. The system was configured in several operating modes for permeate production with the ability to backwash or clean the membrane once certain conditions were met. A process and indicator design (PID) chart, Figure 4-13, was used to layout the various components of the filtration process.

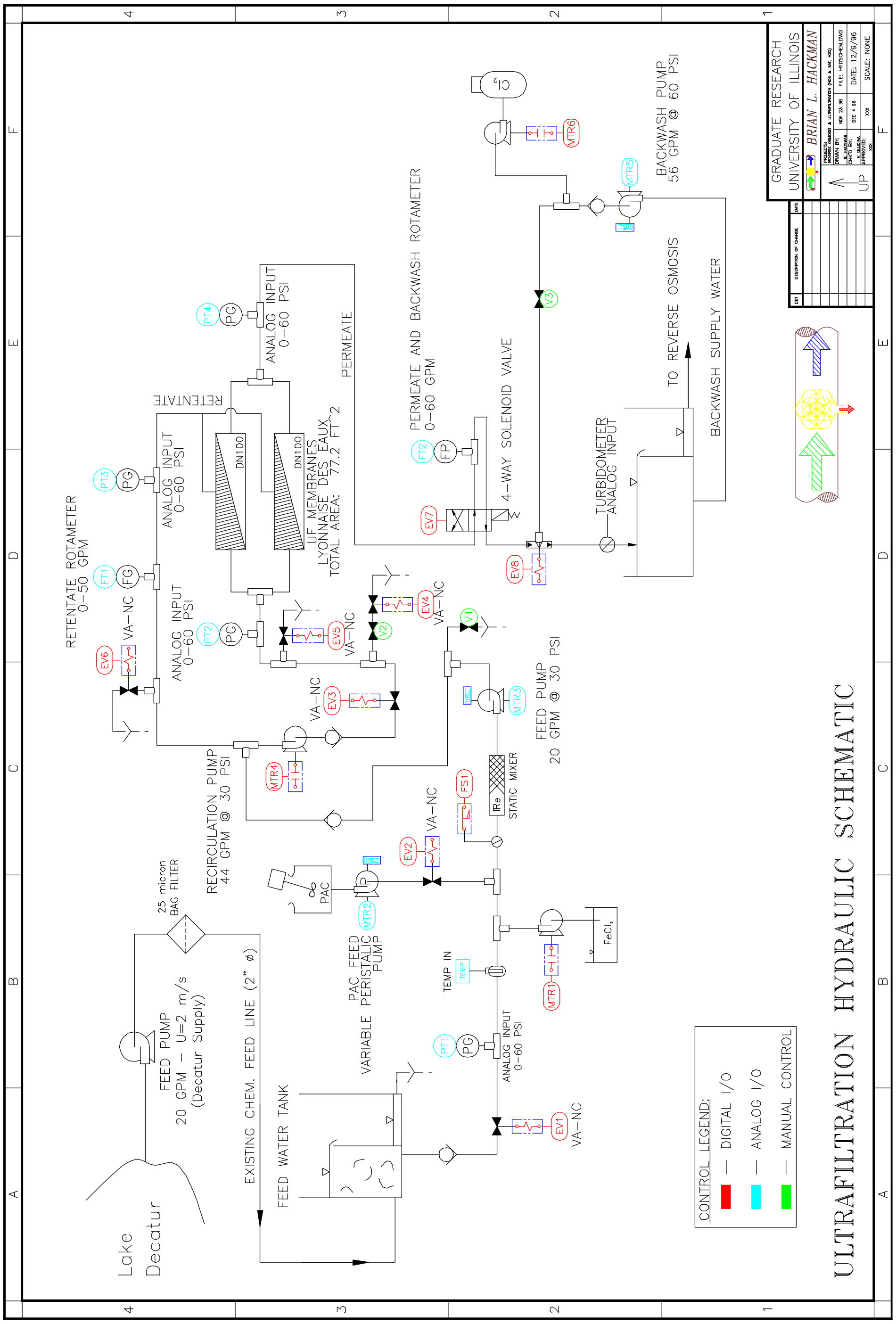

**Figure 4-13**
**UF Pilot PID Diagram**

System components were selected following design. The process was controlled by a 7200/120 PowerPC Macintosh computer (Apple Computer, Inc., Cupertino, California) as described in the project proposal. The automation of the system was made possible using LabView Graphical Programming for Instrumentation, Version 4-1 for Macintosh from National Instruments, Inc.—Austin, Texas. This software was adapted from previous studies and membrane pilot tests at the University of Illinois (Lee, 1992; Wetterau, 1994; Marriot, 1996; Jack, 1997). The software uses graphical components to develop a logical process diagram, which could control the appropriate valves and pumps based on selected methods for operation.

In combination with the software, hardware cards, also available through National Instruments, were used for the data acquisition and control (DAC). For this project, two PCI cards were purchased (Table 4-7). With the DAC cards, various voltage signals, both analog and digital, could be sent or received according to the electronic portion of the system being referenced. Most of the mechanical and electrical components were purchased with the ability to tie into the computer.

**Table 4-7**
**National Instruments PCI Card Description (National Instruments, 1998)**

| Card Description | Connection | Analog | Digital |
|---|---|---|---|
| PCI-1200 for MacOS<br>Part #777097-01<br>Low-Cost Multifunction I/O | Inputs | 8 SE, 4DE<br>(12-bit resolution) | 24 I/O<br>3 – 8 bit port sets<br>TTL Logic |
| | Outputs | 2<br>(12-bit resolution) | |
| PCI-MIO-16XE-50 for MacOS<br>Part #777099-01<br>E Series Multifunction I/O | Inputs | 16 SE, 8 DE<br>(16-bit resolution) | 8 I/O<br>1 – 8-bit port set<br>TTL Logic |
| | Outputs | 2<br>(12-bit resolution) | |

## Mechanical and Electrical System

### Electrical Supply

For all components to work, a standardized electrical use system was applied. The COMFOWARD plant was wired for 208VAC-Y three-phase with transformers to 120VAC single phase, 24VDC, and 5VDC. At the University of Illinois and WMRC, there were no problems in meeting this plan. However, in Decatur, a 15kVA transformer was required to transform the 480VAC-Δ three phase voltage down to the 208VAC-Y three-phase pilot wiring (Square-D, Part #Y48D28T15N).

With regards to protecting the electrical supply and operators, additional consideration was made to prevent shock and electrical damage according to national electrical code. To prevent accidental power down of the computer, an uninteruptable power supply (UPS) was incorporated. Components to the system were protected by surge protected plug-in strips and, in some cases, with a ground fault protected outlet. All internal transformers and equipment were protected from overamperage with bayonet style fuses or in-line fuse blocks. System components consisted of NEMA 3X or better enclosures to prevent any water sprays from entering electrical connections. Splash guards and conduit protected open electrical equipment and loose wires. Finally, main switch disconnects were placed in easy-to-find locations near the pilot while preventing accidental exposure to active circuits.

**Pumps**

As recommended by Karl Glucina (Lyonnaise des Eaux—CIRSEE), the system pumps were sized for the DN100 module. Price quotes were then requested using the previous requirements from various pump manufacturers as mentioned in the Thomas Register (http://www.thomas.register.com). Centrifugal type pumps, with variable frequency drives (VFDs), were selected over other pumps, such as progressing cavity, gear, or peristaltic pumps, because of their ease of use, low maintenance, ability to handle fluids containing abrasives at low pH, and lower cost. The VFD's allowed the computer system to control pump speed with a 0-10VDC control voltage (Figure 4-14). Tuskin Corporation distributed the feed and backwash pumps manufactured by the Multi-Duti Pump, Inc., while the recirculation pump was ordered through Grainger Catalogs (Table 4-8).

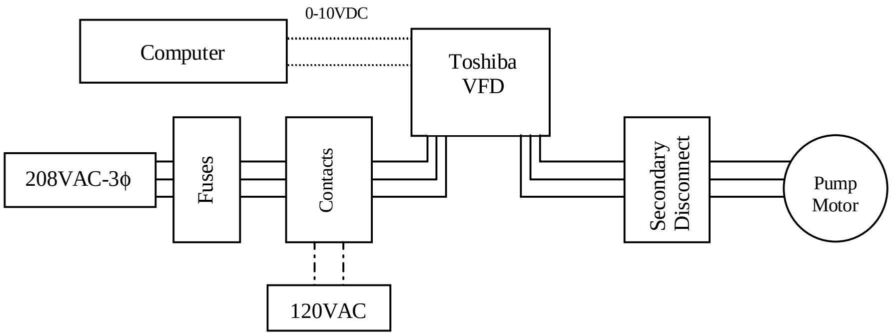

**Figure 4-14**
**Variable Speed Pump Electrical Connections**

**Table 4-8**
**Pump Specifications**

| | Feed Pump | Recirculation Pump | Backwash Pump |
|---|---|---|---|
| Distributor | Tuskin | Grainger | Tuskin |
| Maximum Flow (GPM) | 20 | 12 | 54 |
| Generated Pressure (psi) | 30 | 10 | 36 |
| Voltage | 208VAC | 120VAC | 208VAC |
| Phase | 3 | 1 | 3 |
| Variable Speed | Yes | No | Yes |
| Type | Centrifugal | Centrifugal | Centrifugal |

### Valves

Valves within the plant controlled the direction of flow for the various stages of operation. Three main types of valves were used during operation. Check valves were used to prevent flow reversal. The check valves of one and two inch diameters were made of PVC with viton seals(Hayward, Inc—Part Numbers TC10100ST (1 in. diameter) and TC10200ST (2 in. diameter)). Manual valves were used to operate parts of the system as well as to purge air within the pipes. Manual 1 in. and 2 in. diameter ball valves were made of PVC with viton seals (Hayward, Inc. Part Numbers TB10100ST (1 in. diameter) and TB10200ST (2 in. diameter)). Electrically actuated ball valves were used for automated control of the UF system during its various phases of operation. The electrical actuator (Hayward, Inc.—Part Number EAU29) fit onto specially designed valves similar to the manual valves (Hayward, Inc.—Part Numbers HCTB1100SACTV (1 in. diameter) and HCTB1200SACTV (2 in. diameter)). The actuator required two switched 120*VAC* inputs to activate the valve closure and opening. Figure 4-15 represents the circuitry involved to open/close a valve from the computer.

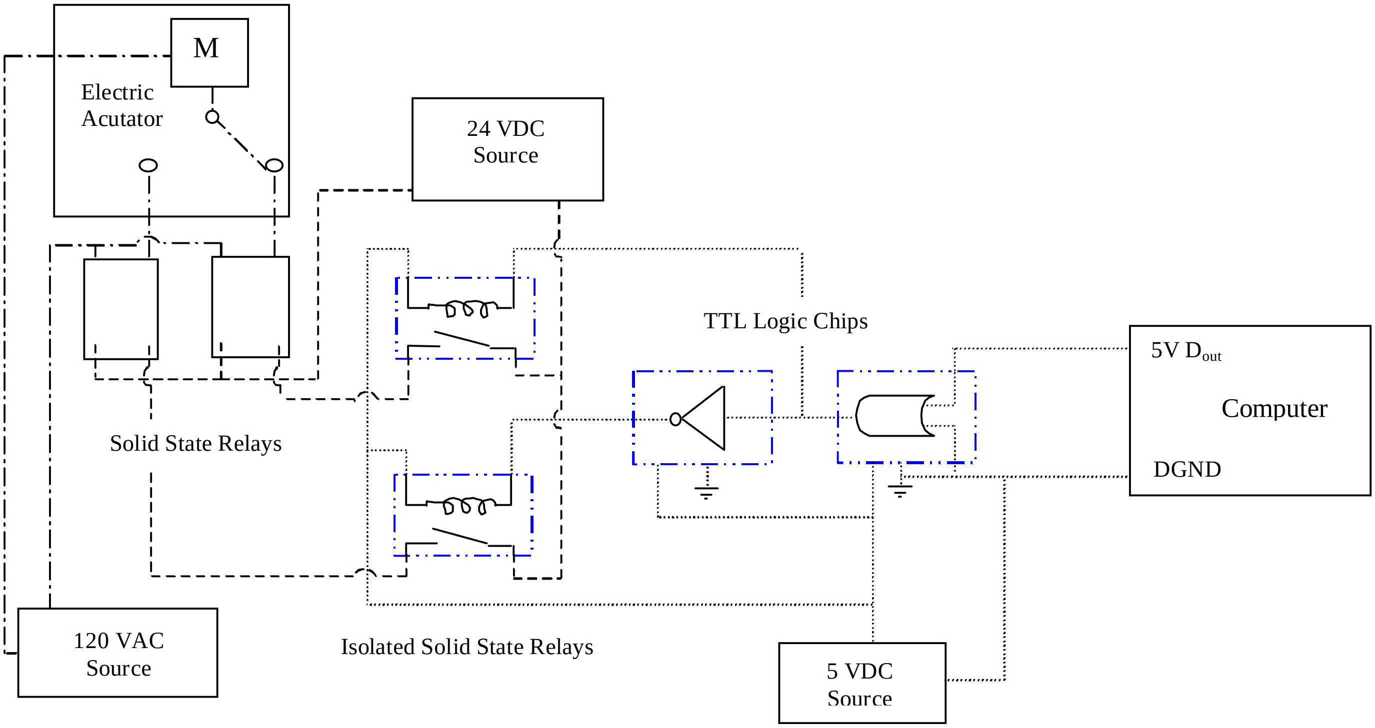

**Figure 4-15**
**Electrical Valve Control Diagram**

While other valves with electric solenoid, hydraulic, or pneumatic controls could have been used, the electrically actuated ball valves worked without additional pumps or compressors. The electrically actuated valves rotate the ball 90° to the full open and closed positions. This type of action was picked because of its ability to reduce shock loading on the membranes by rapid closures usually associated with solenoid type valves. Unfortunately, these valves remain in their last configuration during power loss and do not return to a normally open/closed setting without further action.

### Pressure Gauges and Transducers

After pump and valve components were chosen, pressure gauges and transducers were selected. Fortunately, four glycerol filled pressure gauges (0-100 psi (0-689 kPa)) from Ametek remained intact from the donated USEPA pilot. These served as a visual reference for pressures developed in the pilot.

To allow the computer to measure pressures within the system, industrial type pressure transducers were purchased from Omega Engineering, Inc. (Part Numbers PX181-060G5V (0-60 psi (0-414 kPa)) and PX181-500G5V (0-500 psi (0-3447 kPa)). These particular pressure transducers work with a 24VDC source to energize a semiconductor; as pressures changes, the piezoresistance within this transducer's circuitry changes. The difference in resistance results in a 1-5VDC output that is proportional to the pressure applied (Omega, 1995).

While in place, the pressure gauges were used to calibrate the transducers. Calibration consisted of running the coinciding pumps at various speeds within closed loops while monitoring the voltages from the transducers using the Transducer Monitoring Programs written with LabView. At each monitoring point in the system, the gauge and transducers were located on the same vertical plane directly across from one another on a 4-way PVC cross. For a range of pressures, a table of voltages could be developed and graphed within a Microsoft Excel Spreadsheet. Trendlines were used to fit the data and generate an equation to translate the voltages to pressure. Generated equations from the spreadsheet could then be placed within the COMFOWARD programming. This procedure was similar for all other transducers.

**Flow meters**

For monitoring flows throughout the system, a series of rotameters and low vortex shedding transducers were installed. Flow was measured manually with variable area rotameters (Cole-Parmer Part Numbers H-32472-01 (2-20 GPM) and H-32473-01 (6-60 GPM)). The manufacturer's installation instructions (Blue White Industries, 1994) were followed to prevent errors in measurement. These gauges were simple, low cost, and easy to maintain (Palluzi, 1992).

Difficulties arose in trying to find a reliable, accurate, and inexpensive ways to monitor flux. The feed water contained particles and chemicals that could have had a serious effect on some meters. Thus, paddles wheel and other volumetric type sensors were ruled out. Differential pressure devices, a common and proven measurement method, were not selected because particles can clog the concentric orifice causing errors in measurement (Palluzi, 1992).

Vortex shedding flow meters appeared to be the best choice for this project because of their insensitivity to variable fluid characteristics, reliable and accurate electronics, and easy calibration. By measuring the vibration frequency produced by water forced pass a blocking body, a series of circuits within the vortex flow meter transforms this frequency into a 4-20mA output. With high tolerance 499±1.00 ohm resistors, the 4-20mA output could be read as a 2-10VDC signal to the computer. Again, the instructions for installation were followed to avoid errors (Cole-Parmer, 1995). Calibration of the flow meters consisted of running the pumps at various flows with manual measurement. As for the pressure transducers, voltage and flow curves could be produced in the form empirical equations to be placed into the COMFOWARD program.

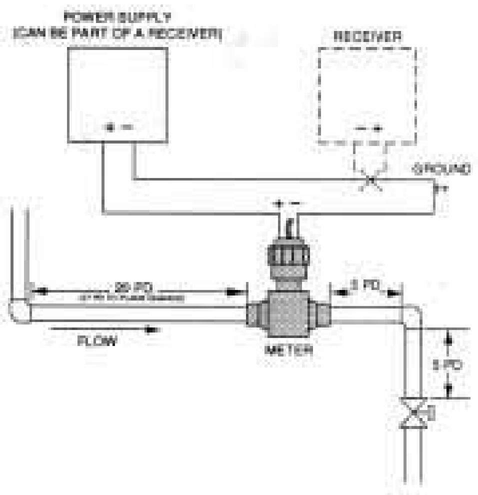

**Figure 4-16**
**Flow Transducer Set-up (Cole-Parmer, 1995)**

### pH

Two controllers allowed for the measurement of pH at the inlet to the UF and the outlet of the RO system. The two controllers (Cole-Parmer Part number 57000-00) were individually wired to self-cleaning flat electrodes with automatic temperature correction (Cole-Parmer Part number 27003-00). A 4-20mA output was included with the controller, but manual monitoring proved to be more reliable because of electrical interferences. Regardless of operation, changes in pH during test runs were small (±0.2) and did not necessitate continuous monitoring. Calibration of the sensor was conducted according to the controller's operating manual (Cole-Parmer, 1995) with NIST traceable calibration standards from Fisher Scientific at pH 4, 7, and 10.

### Conductivity

Conductivity sensors were wired like the pH meters and were employed as an indication of total dissolved solids present within the UF feed and RO permeate waters. Controllers (Cole-Parmer Part Number 19300-10) came with connections for a probe and a 0-100 mV output. Two different style probes were used due to the difference in dissolved solids (DS) content being measured. The UF feed water probe (Cole-Parmer Part number MN-19500-20) contained two Platinum bands to provide a cell constant of 1.0. The RO permeate probe (Cole-Parmer Part number MN-19500-22) utilized three Platinum bands for a cell constant of 0.1 required for lower concentrations of DS. Calibration of the controllers was conducted according to the operating manual (Cole-Parmer, 1992) with NIST traceable calibration standards from Cole-Parmer at 46.7 microS and 445 microS.

### Temperature

Temperature monitoring was accomplished with a Type-K thermocouple and transmitter (Omega Engineering, Inc., Part numbers CF-000-K-2-60-1 (probe) and TX53-K1 (transmitter)). The probe was electrically isolated by wrapping the tip with a thin, sealed layer of electrical tape. The electrical tape delayed temperature sensing by a few seconds, but allowed for accurate monitoring by computer. Temperature calibration took place with the use of a NIST traceable thermometer marked in 1°C increments. The probe and thermometer were submersed in hot and cold tap water contained in a closed styrofoam vessel.

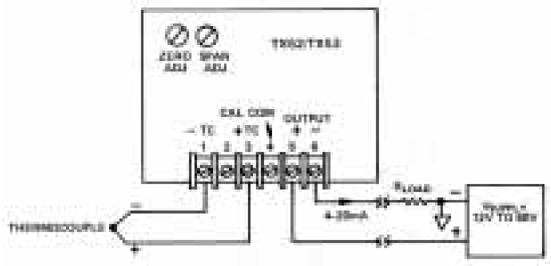

**Figure 4-17**
**Temperature Transmitter (Omega, 1992)**

### Turbidity

Throughout testing, turbidity in the UF permeate was monitored using a Hach Ratio/XR turbidimeter fitted with a flow-through Hach turbidity cell. The turbidimeter allowed for the connection of a 0-5 V output to the computer monitoring system. The turbidimeter was calibrated for use with the COMFOWARD system using the Hach gel standards.

**Piping—**All piping within in the system was consistent with either Schedule 80 or Schedule 40 PVC standards. Two types of connections (threaded and glued) were necessary to join the plumbing components. All threaded connections were wrapped in teflon tape to prevent leakage. For glued joints, the ends of each male and female connection were sanded, cleaned with standard PVC cleaner, doped with PVC joint strengthener, and glued with PVC joint glue. Once prepared, the ends were then placed into one another and given a quarter turn to establish a near leak free connection. All PVC cleaners and glues were approved for safe use with potable water systems.

With the pilot plumbing in place, water pressure was applied above 40 psi (276 kPa) and any leaks were repaired. Using a miniature butane flame torch and generic glue sticks, a seal could be formed within the leak.

**Permeate Tank—**A 33 gallon (125 L) plastic garbage container stored the UF permeate water for backwashing purposes. A PVC connection was made at the bottom of the tank for the backwash pump to draw backwash water. The tank was placed above the pump to prime the centrifugal pump. To prevent air from being drawn into the system, a 3 in. (7.6 cm) long piece of tubing was used as a flow disrupter at the tank. Prior to the backwashing, free chlorine was added to this tank.

## RO Pilot Plant

The reverse osmosis pilot plant built by Osmonics, Inc. was provided by the Waste Management and Research Center of the Illinois Department of Natural Resources. While few modifications were necessary, the pilot required a few simple components to allow the computer to monitor and control the RO process. It should be noted that this pilot required little or no automation due to the simplicity of operation.

### System Characteristics

With its two centrifugal pumps, the RO pilot could produce pressures of up to 800 psi (5500 kPa) providing a total flow of 4 GPM. The pumps consisted of multi-stage, stainless steel, centrifugal impellers (Tonkaflo™, Osmonics, Inc.). Prior to the pumps, a 15 micron filter, with air purge, removed any particulate matter that could damage the pump's impellers. Stainless steel (316SS) ½ in. diameter pipes and valves were used for all of the high-pressure flows. Plastic tubing, similar to that used in carbonated beverage dispensers, transported the water from the pilot. Two pressure vessels, one from Osmonics, Inc and one from Code Line Advanced Systems, Inc., allowed for the operation of 2 in. and 2.5 in. diameter by 40 in. length spiral wound elements respectively. For monitoring purposes, two 0-500 psi (0-3500 kPa) pressure transducers from Omega Engineering, Inc. were installed. Figure 4-18 illustrates the RO process.

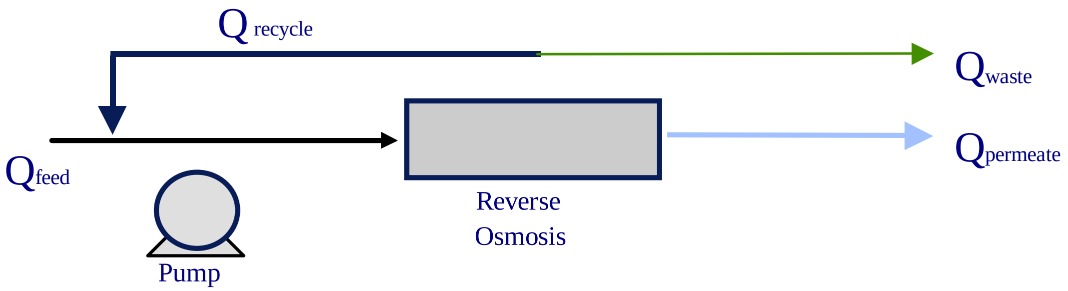

**Figure 4-18**
**RO System Diagram**

Gauges on the system allowed for operational data to be collected. Two pressure gauges were used for measurements at the feed and retentate sides of the membrane. Two sets of three rotometers quantified the amount of concentrate and permeate produced. One temperature gauge monitored the temperature of the permeate water.

A 33 gallon (125 L) HDPE plastic feed tank allowed for the storage of UF permeate or tap water leading into the RO pilot plant. The container was isolated from the free chlorine added for the UF backwash cycle through the use of a 1-1/2 in. diameter tube connected across the top of the UF permeate tank and the RO feed tank. The feed tank was covered to prevent particles from entering into the pumps. At the bottom of the container, a PVC connection was made for a 2 in. diameter PVC pipe drain into the RO feed pump.

To cool the permeate water to ambient temperatures, a heat exchanger was used with a tap water hook–up. After the heat exchanger, the warmed tap water was then used as feed water to the RO pilot. While not significant in the pilot study, heat extraction from the processes could serve as a form of energy recovery with full-scale treatment.

For automatic operation, a three-position two-pole switch made it possible to turn on the system by hand or computer. Any error in operation was detected through alarm circuitry, and the pilot was automatically shut down regardless of switch position. An alarm bypass button allowed for further operation beyond system error conditions.

## Membranes

### Specifications

In response to letters asking for support, three manufacturers we able to supply the project with a total of seven membrane elements. Hydranautics, Inc. provided the necessary pressure vessel to house the larger 2.5 in. diameter elements. For the purposes of testing, a variety of membranes providing high rejection to moderate softening were selected to show their effectiveness in nitrate removal. Table 4-8 gives a brief description of the membrane elements used in this study.

**Table 4-9**
**Reverse Osmosis Membrane 2040 and 2540 Test Elements**

| Manufacturer (Surface Area) | Element Type | Description |
|---|---|---|
| Fluid Systems, Inc.<br>(27 $ft^2$)<br>2.5 $m^2$ | TFC-HR | Thin Filmed Composite–High Rejection–RO |
| | TFC-S | Thin Filmed Composite–Softening–near NF |
| | TFC-ULP | Thin Filmed Composite–Ultra-low Pressure–RO |
| Hydranautics, Inc.<br>(28 $ft^2$)<br>2.6 $m^2$ | CPA2 | Polyamide–High Removal–RO |
| | ESPA | Polyamide–Energy Saving–RO |
| Osmonics, Inc.<br>(16 $ft^2$)<br>1.5 $m^2$ | HR-CA | Cellulose Acetate–High Rejection |
| | HR-PA | Polyamide–High Rejection |

Of the group, six of the membranes were composed of thin filmed composite polyamide material. The seventh was composed of a cellulose acetate blend (HR-CA).

### Storage

Proper storage of the elements prevented degradation of the membrane surfaces. The elements were shipped in a sealed bag and box. Each spiral-wound element was placed into a dark temperature control room held at 4°C (39°F) to be later transported to and from Decatur in the trunk of a car. Care was taken to avoid heat and direct sunlight which could have caused irreversible damage.

In Decatur, previously used elements were treated with a 5 mg/L sodium metabisulfite solution to preserve the membrane for future use. The element would then be removed from the pressure vessel, and the new one installed. Just before installation, the membrane housing would be filled with tap water and the new element placed inside. The membrane was soaked overnight to allow full wetting of the membrane. Tap water with a free chlorine concentration of 1 mg/L was flushed through the pilot during periods of short term storage (3-5 days) to prevent biological contamination.

After use, the spiral-wound units were returned to their original shipping containers and transferred to the temperature control room. At the University of Illinois, the element was immediately returned to the temperature control room.

#### Remote Monitoring

To remotely control and operate the system from Urbana, installation of a telephone line in Decatur was necessary. Remote monitoring made it possible to track changes and make adjustments in the pilot plant operation. While remote operation has been used in other research projects at the University of Illinois (Jack, 1996), additional programming within the course of this project allowed for the computer to notify the user of system failure by automatically dialing a preset phone number. Within the COMFOWARD programming in LabView, system errors would cause a script written for the third telecommunications software, MacComm Center, to be activated. Preset phone numbers would then be dialed and the research assistant, with the help of caller ID, could recognize that the system was shut down.

To accomplish dial-in access and control, three telecommunication software packages were employed. The first, Apple Remote Access (Apple Computer, Inc., Cupertino, California) allowed the remote computer in Decatur to act as a dial-in server. Timbuktu Pro for Macintosh Version 2.0 (Farollon Computing., Inc, Alameda, California) could then be used to connect the remote and local computers via modem, allowing direct access to the operation of either computer. The remote computer appeared as a virtual window on the local computer where all keyboard and mouse commands could be viewed and sent in real time. Two Hayes Accura 288 V.34 modems (Hayes Microcomputer Products, Atlanta, Georgia.) made the connection between the two computers possible.

A future update of LabView to Version 5.0 would allow the system to be readily operated through TCP/IP links over a phone line. With the TCP/IP links to internet servers, the newer software allows for homepages with the systems parameters automatically generated and controlled, with password protection via HTML and JAVA programming (National Instruments, 1998).

## Methods

### *COMFOWARD Operation*

#### UF Pilot—Permeate Production

Given the versatility of system parameters with the electronic automation, a number of production modes were possible for the UF pilot. The Graphical Display (Figure 4-19) allowed great flexibility in operation of the UF Pilot as described below.

**Operation Mode**
Three basic operation modes were used. With the Deconcentration Mode Selector in the

“Off” position, dead-end filtration took place with internal loop recirculation and a hollow fiber cross flow velocity near 1 m/s. The dead-end filtration mode concentrates solids within the recirculation loop until a backwash occurs after the appropriate end mode was reached. Dead-end filtration (Figure 4-20) was used with waters of lower turbidity (<30 NTU) when the system could handle incremental increases in solids concentrations without the build up of a cake layer causing rapid flux decline or pressure increase.

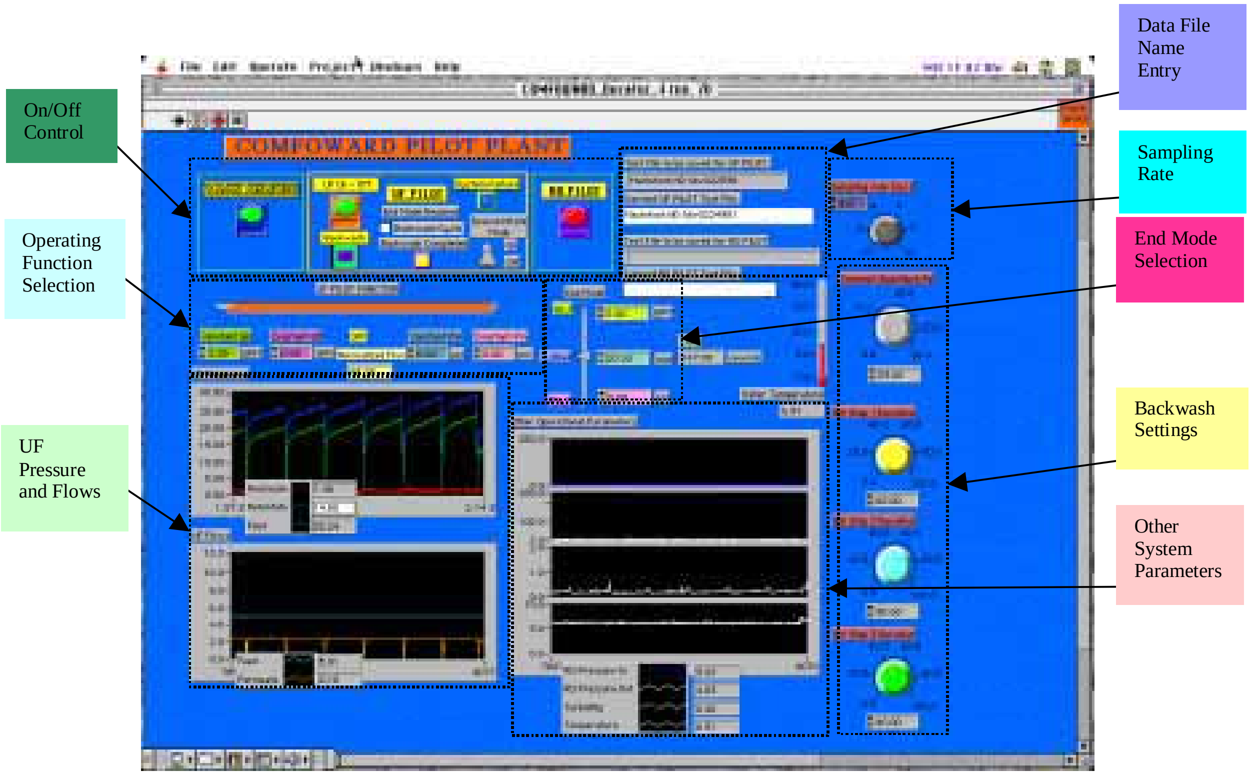

**Figure 4-19**
**Graphic Display Set-up for UF Pilot Operation**

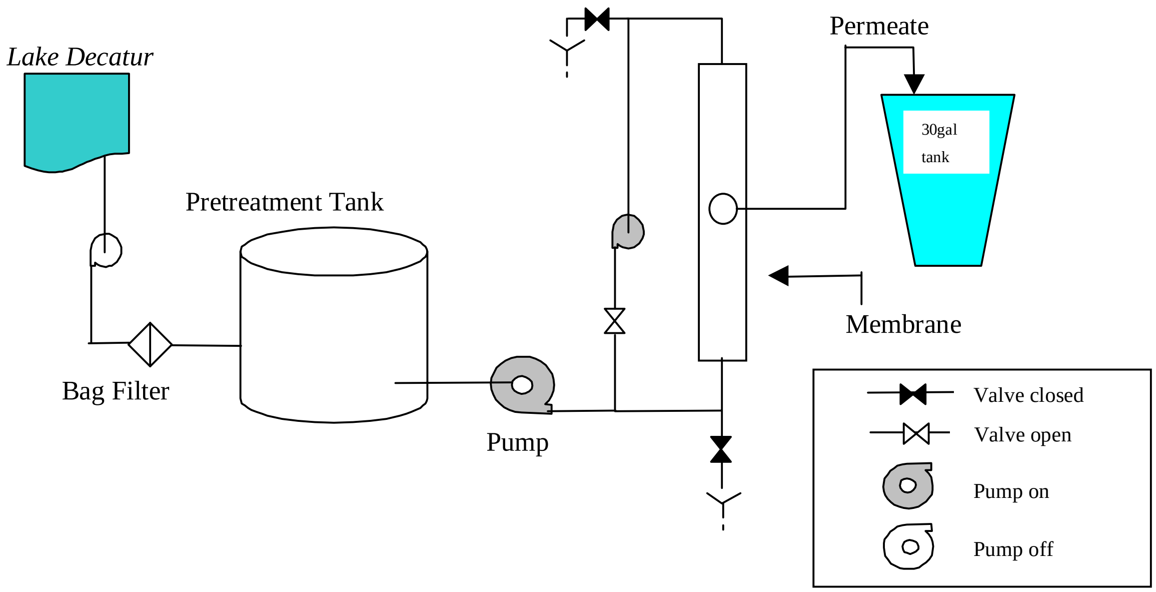

**Figure 4-20**
**Diagram of Dead-end Filtration**

With the deconcentration mode selector switch toggled to the “On” position, the other two modes of operation could be invoked. For waters with turbidities higher than 60 NTU, the build up of solids within the system became too great under dead-end conditions. The first method for deconcentration allowed the UF retentate to be directly disposed of without recirculation by opening a waste valve on the system. Recovery rates of 50% or better with this first stage UF pilot could be achieved under these operating conditions. When turbidities fell between 30 and 60 NTU, the retentate flow from the waste drain was diverted back into the feed tank. The drain on the pretreatment tank was allowed to bleed the system at a controlled rate allowing steady state conditions. While more particulate matter was held within the pilot, operation was still possible. Cross-flow velocities within the membrane for this set-up (Figure 4-21) were held at 1 m/s or better.

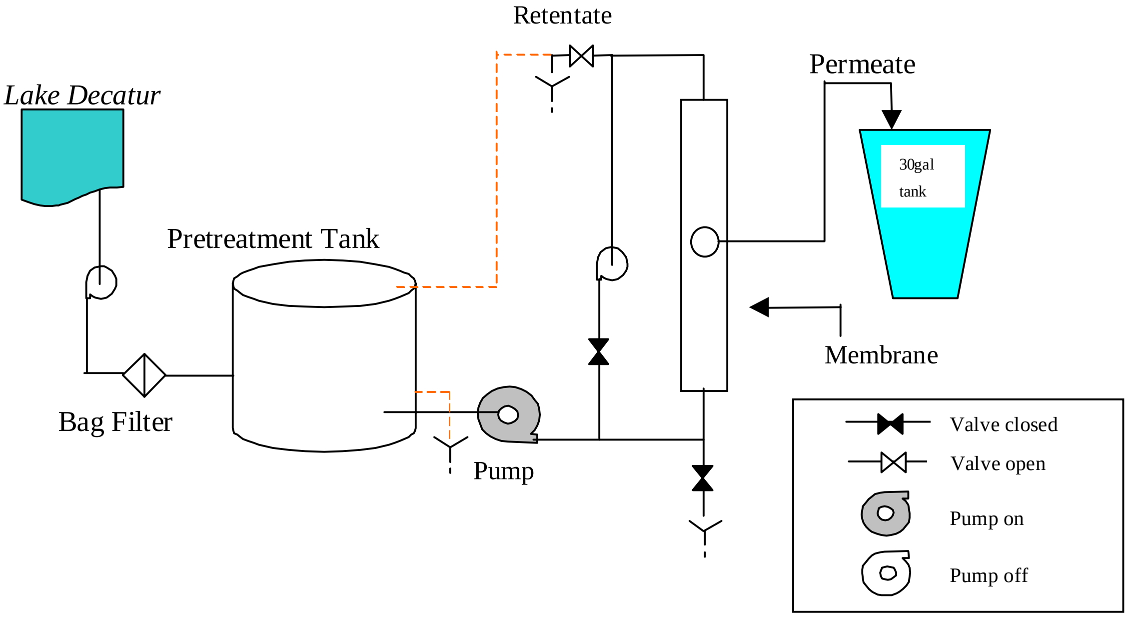

**Figure 4-21**
**Deconcentration Mode with Retentate Recovery**

### Operating Function

Within the Operating Function Selection Section, four different types of experimental functions could be selected. Two methodologies involved constant flux operation by controlling the total influent flow rate ($Q_I$) or the total permeate flow rate ($Q_p$) in units of GPM. In either mode, the COMFOWARD software would monitor the flow rates from the respective flow transducer. If the flow rate did not match within a given tolerance range, the pump speed could be increased or decreased to match the set flow rate.

The other two selections involved the use of pressure as a control strategy for the UF pilot. Here a set pressure for either the feed into the membrane ($P_f$) or the transmembrane pressure ($P_t$) could be selected and entered into the graphical control display. With a set pressure, the LabView programming monitored the pressure transducers and made adjustments to the pump speed in a fashion similar to the previous flow options.

### End Mode Selection

As the software continuously monitored and controlled the system, a selected end mode would be reached allowing for backwash or other actions to be taken. Three different end conditions were made available. Although not used for the tests with the UF pilot, a minimum permeate flow ($Q_p$) could be set as the end for a particular cycle. In this case, flux would not be allowed to reach a minimum rate.

The time end mode offered the ability to backwash at regular intervals. Given that pressures and flows applied to the membrane were dependent upon the interval between backwashes ($t_i$), this mode allowed for comparison between cycle times with the raw water.

In order to prevent the membrane from reaching the maximum pressure condition, a maximum feed pressure ($P_f$) could be set to control backwashing events. This mode was used to operate the pilot during its initial test runs prior to the use of coagulant aids.

**Emergency Shut Down Considerations**

Should an emergency condition be detected by the system, such as maximum feed pressure ($P_f$ = 29 psi (200 kPa)), the system would shut down within seconds of detection. Within the feedback loop of the program which monitors pressures, flows, and time, provisions could be set to indicate that the system was not performing properly. If the emergency conditions were met prior to the end-mode setting, then the system would close all valves and halt all pumps. Towards the end of experiments, software programming made it possible to contact the operator via the phone to indicate the system had shut down. The system could be restarted over the remote monitoring system, although in most cases, a site visit was required to either replace the pre-filter or a pretreatment solution.

**Data Logging**

As long as the system remained operational, the UF pilot software wrote transducer data directly into files setup for analysis with spreadsheets. Filenames could be entered into the graphical display that would indicate the water source and date of testing. Rates of data collection were controlled by setting a sampling rate for which each transducer would be read and logged. Within all particular files, all pressures (feed, retentate, permeate), flows (feed and permeate), time, temperature, effluent turbidity were written in a format compatible with Microsoft Excel spreadsheets.

### UF Backwash

**Setup**

Backwashing of the UF module was required to remove the cake or gel layer formed during the production of particle-free permeate water. This process cleans or rejuvenates the filter by forcing clean, chlorinated water back through the hollow fibers. The backwash pressure was normally set at 34 psi (234 kPa) to avoid damage to the membrane.

With the pressure set, the automated backwash procedure could then be varied by selecting the intervals of time to wash the membrane. During an actual backwash run, the system was monitored for pressures, and adjustments in the pump speed could be made to match the pressure set point. If pressures went above or below set operating conditions, the UF system would automatically cease and the operator would be notified.

**Backwash Procedure**

The following steps and Figures 4-22 to 4-27 describe in detail the automated backwash procedure used with the COMFOWARD system in combination with the Aquasource DN100 UF Module from Lyonnaise des Eaux. Based on membrane performance and water quality, backwashes were conducted at various times from every 15 minutes to 1 hour of permeate production. The figures only incorporate the valves and pumps required for backwash and do not represent the entire system diagram.

**Step 1:** Chlorine Addition

Regular bleach (Clorox, 5.25% Sodium Hypochlorite, NaOCl) is added to the 30 gallon (114 L) tank to reach a free chlorine concentration of 5-7 mg/L in the permeate/backwash water as measured by Hach DPD method for free chlorine. Losses in free chlorine to chloramines and other byproducts are in the range of 0.1 to 0.2 mg/L as indicated by the Hach DPD method for total chlorine. The addition of NaOCl is regulated by turning on and off a variable speed Masterflex pump and drive (Norflex tubing) for a set period of time using the COMFOWARD/LabView programming. All valves are closed at the start of the backwash cycle.

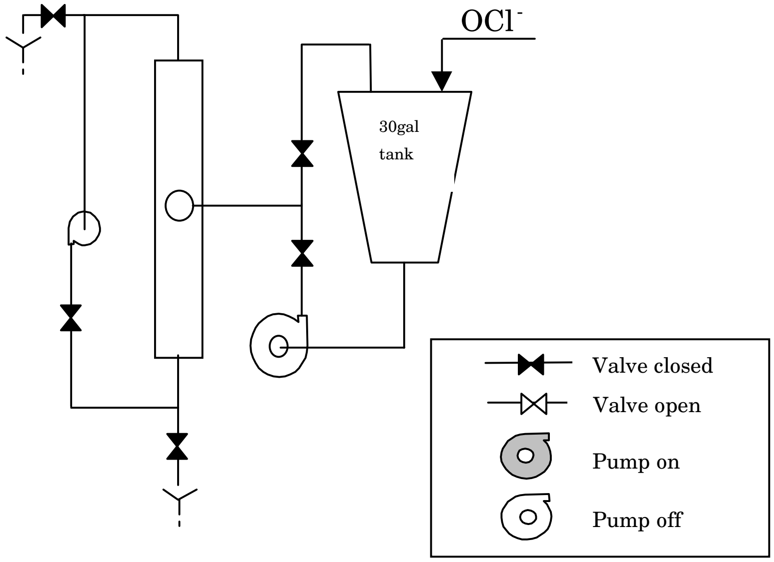

Figure 4-22
Chlorine Addition

**Step 2:** Permeate Loop Air Purge and Wash

The chlorinated permeate is then flushed through the permeate loop using the backwash pump. Water is not allowed to permeate through the membrane outside in.

This step lasts for 20 seconds at a flow rate of 15 gpm and assures that the permeate side of the system remains free of bacteria.

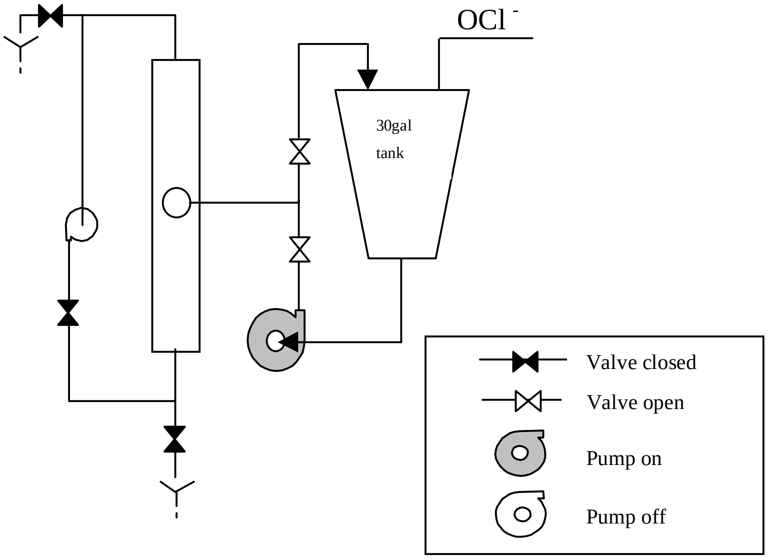

**Figure 4-23**
**Air Purge and Wash**

**Step 3:** Upper-half Module Backwash
After flushing the permeate loop, backwashing commences on the upper-half of the UF module with the permeate/disinfectant. During this step, the automation program activates the upper drain valve and adjusts the pump speed to match the pressure preselected from the COMFOWARD front panel display on the computer. Backwash pressures for the system were at 33-34 psi (228-234 kPa). This avoids the maximum backwash pressure of 36 psi (248 kPa). The length of run for this step varies for a given backwash time (Table 4-10).

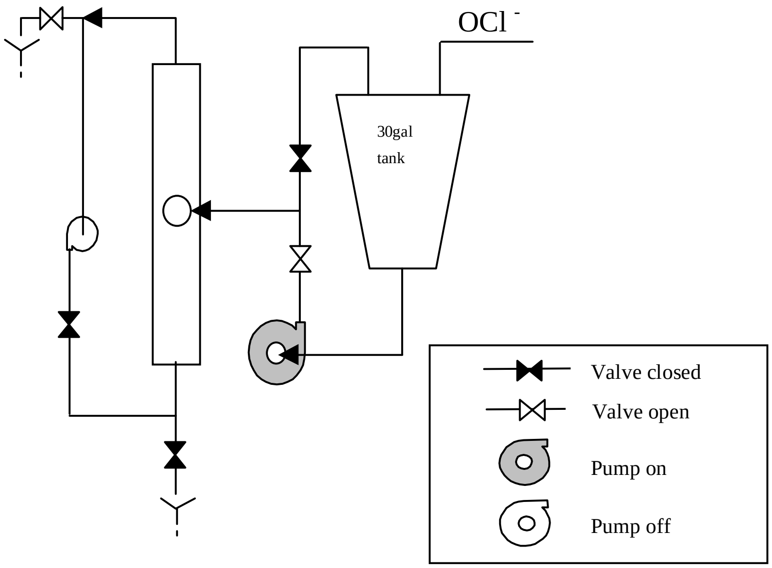

**Figure 4-24**
**Module Wash—Upper Half**

**Table 4-10**
**Upper-half Module Backwash**

| Total Backwash Time | Time—Step 3 |
|---|---|
| 1 minute | 15 seconds |
| 2 minutes | 20-30 seconds |

**Step 4:** Lower-half Module Backwash
With one-half of the membrane backwashed, the system switches valves to clean the other half of the membrane. The automation program switches valves after time has run out for the first step and continues this step until its time expires. Pressure is held constant at 33-34 psi (228-234 kPa). The length of run also varies for this step, and is longer than Step 3 because of the actual volume of pipe is larger in this section (Table 4-11).

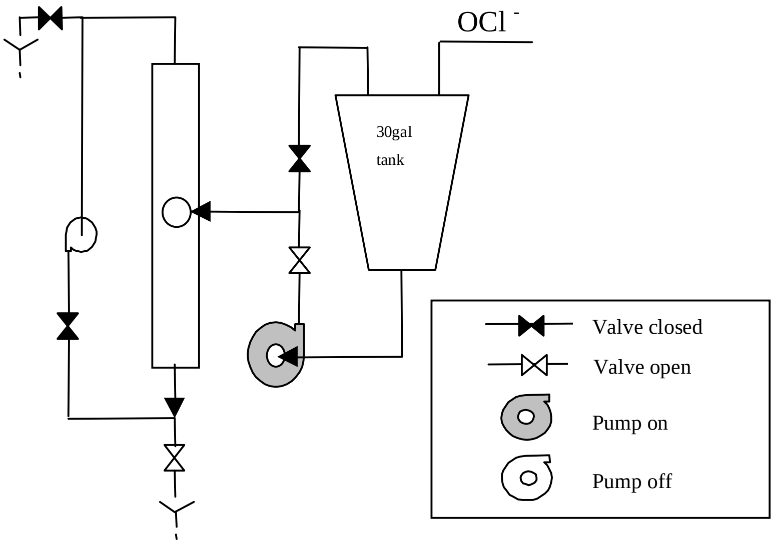

**Figure 4-25**
**Module Wash—Lower Half**

**Table 4-11**
**Lower-half Module Backwash**

| Total Backwash Time | Time—Step 4 |
|---|---|
| 1 minute | 20 seconds |
| 2 minutes | 30-40 seconds |

**Step 5:** Full Module Backwash and Surface Scour
To include a full system cleaning of the recirculation loop, a surface scour of the membrane was implemented at 0.7 – 0.8 m/s during backwash. Pressure was again held constant at 33-34 psi (228-234 kPa) for this step. The length of run varies for this step and allows for a longer running time than Step 4, to remove any particles that have built up in the recirculation loop regardless of production mode (Table 4-12).

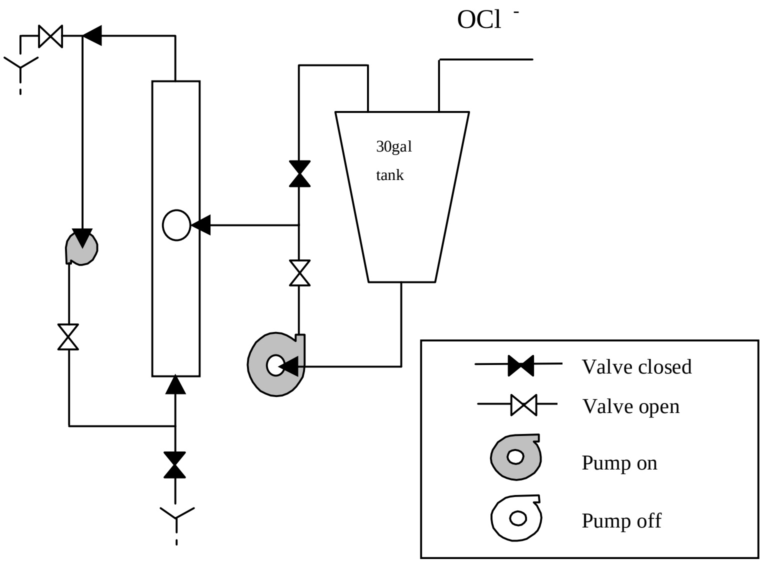

**Figure 4-26**
**Full Module Wash and Scour**

**Table 4-12**
**Full Module Backwash and Scour**

| Total Backwash Time | Time—Step 4 |
|---|---|
| 1 minute | 25 seconds |
| 2 minutes | 50-70 seconds |

**Step 6:** Backwash Stop and Prepare for Permeate Production
With Step 5 concluded, all system valves close, and the system waits for 5 seconds. The waiting allows for all valves to reach the closed position and allows the system to settle and avoid pressure shocks. Shortly after the wait, the feed pump introduces the feed water as it flushes through the system to start production of permeate shortly thereafter.

Total water use during backwash is described in Table 4-13.

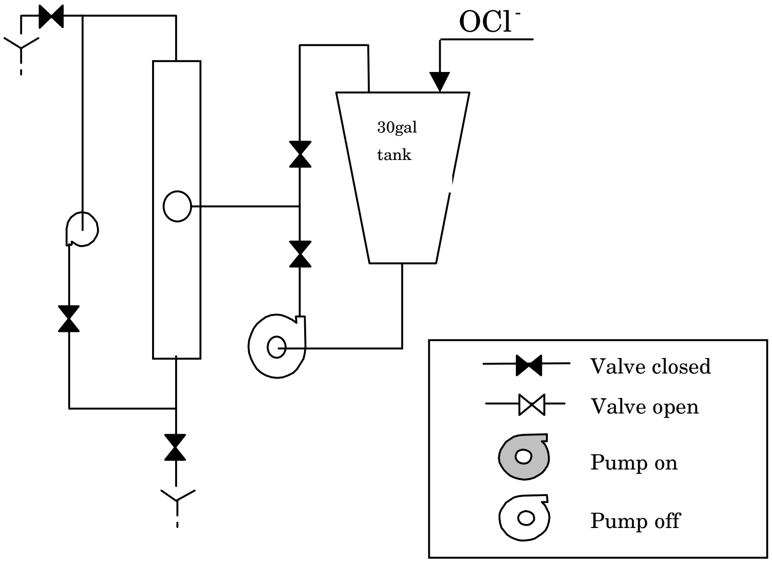

**Figure 4-27**
**Backwash Procedure Complete**

**Table 4-13**
**Backwash Water Volume Used**

| Backwash Time | Volume Used |
|---|---|
| 1 minute | 8 gallons (30 L) |
| 2 minutes | 15-16 gallons (57-61 L) |

## RO Pilot

To assure no damage occurs before any operation, the system was purged of air with the aid of a purge valve on the 15 micron prefilter housing. By operating a feed pump provided with the plant, water could be flushed through the system. Following the system purge, the concentrate and recycle valves were set to operate the system at one set pressure point. Pressure increases up to 200 psi (1379 kPa) on the membrane were slowly induced over a span of five to seven minutes. After reaching the set pressure, the computer was allowed to monitor and control the pilot.

## Water Quality Analysis

### Sample Collection and Handling

The numerous samples in this project required a standard handling procedure. For the majority of samples collected for analysis at the University of Illinois, 500 mL Nalgene bottles were used. These containers were durable and withstood transport to and from Decatur, Illinois; they also allowed minimal contamination of the sample prior to analysis. With the collection of each sample, both bottle and lid were rinsed and then filled full with the sample. Permanent markings on the side of the container indicated the type of sample, date, and time of sample. Once the grab sample was collected, it was transported back to University of Illinois in the trunk of an automobile and immediately placed into a temperature control room, held at 4°C (39°F). Samples for dissolved solids analysis were also collected in 500 mL Nalgene sample bottles. One set of specimen containers at the SWTP were used to collect samples and were cleaned after use.

To prepare the 500 mL aliquots for TOC tests, 40 mL glass sample vials were used to divide the sample as needed. All 40 mL vials and other glassware were cleaned through an elaborate process involving an enzymatic soap bath, milli-Q or ultrapure water rinse, sulfuric or hydrochloric acid wash, and drying in a 300 °C (570°F) oven; this minimized the risk of contamination from outside sources during total organic carbon (TOC) analysis. Reynolds aluminum foil was placed under caps to contain the individual samples.

Prior to TOC and UV analyses, raw water samples required pre-filtration to remove particulate matter. For this a Millipore two-part vacuum filter holder held individual 47 mm diameter filters. Supported nylon 0.45 micron filters supplied by Micron Separations, Inc, Westboro, Massachusetts, were used to filter raw water samples. The 0.45 micron pore size allowed the removal of most suspended particles; the hydrophilic nylon material was well suited for filtering fluids containing organic constituents. The raw water was then collected in a clean 1000 mL side-arm flask. From the flask, 40 mL vials could be filled and stored for additional measurements. Between filtration runs, the filter holder was rinsed with soap water, milli-Q water, sulfuric or hydrochloric solution, and milli-Q water.

In order to identify specimens, a common marking system was adopted for all containers, lab notes, and analysis results. Each test sample was labeled as shown in Figure 4-16.

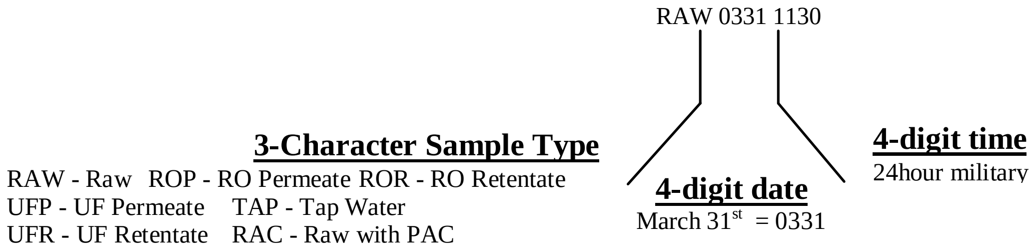

**Figure 4-28**
**Sample Label Format**

### Nitrate

With nitrate as the target compound, regular nitrate measurements were taken in the UF permeate, tap water, RO permeate, and RO retentate. The effectiveness of nitrate rejections for the various reverse osmosis membranes and operational models required consistent measurement in all tests. Using the Hach Nitrate HR method in conjunction with a Hach DR-2000 spectrophotometer, the measurements of $NO_3$-N could be compared to values measured by Decatur's SWTP to assure that readings were accurate. This method uses cadmium reduction to determine the amount of nitrate. Filtration assured us that the spectrophotometric method was not subject to interference from turbidity. De-ionized and distilled (DDI) water with a specific resistance near $10^{18}\Omega$ was used as the standard zero sample throughout all tests.

Just minutes after sample collection, the nitrate test was conducted. One Hach quartz vial was filled with 25 mL of DDI water while the remaining three matched vials were filled with 25 mL of filtered source water, RO permeate, and RO retentate. One full packet of Nitrate HR reagent was added per sample. After addition of the reagent to all samples, the vials were covered with aluminum foil and agitated for one minute by hand. The aluminum foil was used to avoid contamination of the sample by skin contact. Following agitation, the samples were allowed to sit for five minutes during which the vials were wiped clean with lint-free lab tissue. The DDI sample was then used to calibrate the zero concentration reading of the spectrophotometer. The samples were then placed into the spectrophotometer and measured for nitrates.

### Alkalinity

The measurement of total alkalinity was conducted according to **Standard Methods** Part 2320 Section B (1995). Measurements were reported as mg/L of $CaCO_3$.

### Total Hardness

To assess the concentration of cationic species present, total hardness tests were conducted according to **Standard Methods** Part 2340 Section C (1995). Measurements were reported as mg/L of $CaCO_3$.

### Calcium

As part of the total hardness, calcium measurements were completed using **Standard Methods** Part 2340 Section C (1995). Measurements were reported as mg/L of $CaCO_3$.

### Chlorine

On occasion, both the UF backwash and RO feed waters would be measured for free chlorine. Free chlorine, as previously described, was important for the biological decontamination of the UF system during backwash. The compound also has a negative affect on many reverse osmosis membrane materials. All chlorine measurements were conducted with a Hach visual DPD (N-Ndiethyl-p-phenylene-diamine) kit. Five drops from each of the #1 and #2 reagents from the kits were added to a 2 mL sample. With the development of a reddish-pink color in the sample, a comparison could then be made with a calibrated color strip indicating chlorine concentration.

### Turbidity

Additional raw and tap water turbidity measurements were obtained from the data logging charts at the SWTP. The measurements were taken on a continuous basis with Hach flow-through cell turbidometer routinely calibrated by Mike Loehr, SWTP Instrument Technician.

### Total and Dissolved Organic Carbon

Given the high organic content of soils within the watershed, Lake Decatur typically contains levels of total organic carbon (TOC) near 3-4 mg/L. During storm events when sediments are stirred up from the lake bottom, TOC levels can increase to 10 mg/L. With the higher carbon concentrations, disinfection by-products (DBPs) can also increase in the treatment plant effluent following chlorine addition. This would allow for the formation of trihalomethanes and haloacetetic acids which are considered carcinogenic (Water Production Staff, 1997). The use of PAC within the treatment process can remove limited amounts of organic carbon given the low PAC concentration (10-12 mg/L) and residence time of approximately 4 hours.

TOC analysis conducted on samples prior to and after filtration allows estimation of removal efficiency and fouling potential. Samples from Decatur were placed into pre-washed 40 mL sample vials for TOC analysis as previously mentioned. The nylon filtered samples contained only dissolved organic carbon (DOC). A Pheonix 8000 UV-Persulfate TOC analyzer manufactured by Dohrmann, Inc included an autosampler that allowed 70 samples per test. Three measurements per sample were taken in a consistent manner with the TOC measurement system. If one of the three measurements was outside of the standard deviation, the measurement was discarded as atypical data.

### UV Absorbance

To monitor changes in TOC make-up throughout runs, UV absorbance was measured at 254 nm using a Perkin-Elmer Lambda 3B UV/VIS spectrophotometer. One curvette was used repetitively for the sample containing DOC. Rinsing with DDI water between tests prevented erroneous readings. Before conducting measurements, the samples were allowed to reach room temperature.

### Atrazine

Atrazine can be detected in the raw and tap waters at Decatur, Illinois, especially after crops have been sprayed. However, the use of PAC at the SWTP may reduce maximum concentrations to 3 ppb.

Prior to use, both the testing kit and samples were allowed to warm to room temperature. The analysis for all tests was conducted in the IEPA certified lab at Decatur's SWTP with samples collected within the previous month. The RaPID Assay test kit from SDI, Inc. was used to measure Atrazine. The lab technique is based on an immuno-assay method.

The following procedure was followed to conduct the tests. Enzyme-atrazine conjugate and magnetic particle solutions were added to each kit standard and sample. Enzyme-coupled atrazine and atrazine present in the sample would then compete for the antibody active sites present on the magnetic particles. After an incubation period of 15 minutes, a magnetic tray would be attached to the test tube rack to retain the magnetic particles on the sides of the test tubes. With washing, the remaining solution could be displaced. A color solution was allowed to react with retained particles. In this type of test, a known concentration of enzyme-coupled particles directly competed with any atrazine present for magnetic particle sites. The amount of color development for a given sample, could then be correlated to the amount of atrazine present. With the end of the color reaction, a stop solution turned the sample to a yellow color proportional to atrazine concentration. The Hach DR-2000 spectrophotometer, fitted with a special 1 mL test tube holder and cap, was set at 450 nm to measure sample absorbance. A standard calibration curve for concentration vs. normal absorbance was developed from standards provided with the kit. The method allowed measurement of Atrazine concentrations between 0.1 to 5.0 ppb.

### Experimental Plan

An experimental plan was designed to observe the effectiveness of various treatment processes. To test the UF treatment of the raw surface water, several different pretreatment options were used through testing of the full pilot system. The first set of experiments used raw water without pretreatment. During these tests, UF flux declined significantly over a period of hours. Further experiments with the UF pilot included pretreatment with either PAC-adsorption or PACl-coagulation. With the addition of PAC, no observable increase in membrane performance was noted. With PACl addition, flux decline rates dramatically decreased and the pilot was able to run a period of days without chemical cleaning. Continuous operation of the UF pilot proved to be a challenge and could be improved upon in future testing.

Using the various NF/RO membrane elements with the RO pilot, tests were conducted with either UF permeate or SWTP product water. The UF permeate provided a clean source of water (≈ 0.1 NTU) with high DS concentrations. Pretreatment of the UF permeate with sulfuric acid was required to prevent scaling on the membrane.

The SWTP product water provided the RO pilot with high quality water (0.1 - 0.2 NTU) with reduced DS and containing polyphosphates and free chlorine. With the addition of sodium metabisulfite, the system could be run without the worry of free chlorine destroying the RO membrane surface. The pH was not adjusted to a negative Langlier Saturation Index during these short term tests because the polyphosphates prevented crystal formation. Throughout these tests, both the UF permeate and SWTP product water contained similar concentrations of nitrates.

In order to evaluate the effectiveness of nitrate and atrazine removal, short assays with the RO membrane elements were conducted. All seven membranes were subjected to both the UF permeate and SWTP product water in separate tests at three different pressures.

For each test, the membrane pilot(s) were operated for two hours prior to testing to establish steady state conditions. Once at the set operating pressure, samples were collected. Two or three samples were collected each hour for each pressure depending on feed water quality variations during testing. After sampling at an individual pressure, the pressure was either decreased or increased to the next operating pressure. Following another hour of operation, samples were again collected at the new set point. With each sample, data on pressure, flows, and other operating conditions were noted. Table 4-14 represents the short term experimental setup.

Since time was limited, short term tests for each membrane were needed to assess their individual effectiveness in nitrate/pesticide removal. The use of a single stage system limited the ability to simulate a full scale system, but allows for the estimation of rejection properties throughout a multi-stage plant with changes in influent quality. For testing with a single element test stand, the following information can be inferred (Bersillon and Thompson, Chapter 14 AWWA Water treatment Membrane Processes, 1996):

1. Rejection rates for a contaminant for each membrane element
2. Production of permeate and concentrate for further analysis
3. The determination of salt precipitants on the membrane elements at high recovery rates
4. Compatibility of RO membrane material with various combinations of water and pretreatment options

**Table 4-14**
**Short Term Analysis Plan**

<table>
<tr><th>Membrane Element→</th><th colspan="2">ESPA</th><th colspan="2">TFC-ULP</th><th>CPA2</th><th>TFC-HR</th><th>TFC-S</th><th>OSMO HR-CA</th><th>OSMO HR-PA</th></tr>
<tr><th colspan="2">Feed Water</th><th colspan="5">Pressure</th><th colspan="3">Pretreatment</th></tr>
<tr><td colspan="2">Tap</td><td colspan="2">100 psi<br>(689 kPa)</td><td colspan="2">150 psi<br>(1034 kPa)</td><td>200 psi<br>(1379 kPa)</td><td colspan="3">conventional lime softening, sodium metabisulfite</td></tr>
<tr><td colspan="2">RAW UF</td><td colspan="2">100 psi<br>(689 kPa)</td><td colspan="2">150 psi<br>(1034 kPa)</td><td>200 psi<br>(1379 kPa)</td><td colspan="3">PAC or PACl</td></tr>
</table>

With the completion of short term evaluations, two longer term tests were conducted at the maximum pressure for the RO membrane pilot (200-225 psi (1379-1551 kPa)). Using two comparable membrane elements from different manufacturers (CPA2 and TFC-HR), these elements would be subjected to variable water conditions for 28-30 days of operation. The CPA2 and TFC-HR were selected for the longer test runs because they exhibited the highest nitrate rejections. Testing on the first element, TFC-HR, was conducted below 50 % recovery while the second element, CPA2, was tested at slightly higher recoveries (55%). Daily grab samples were analyzed for nitrates, alkalinity, total hardness, calcium hardness, TOC, UV-absorbance (254 nm), and atrazine. To assure free chlorine was not present, the DPD method was conducted. The accidental exposure of the first element to free chlorine towards the end of the TFC-HR test showed the effect of chlorine on membrane performance.

# 5

# RESULTS AND INFERENCES

## UF Pilot Results

### *UF Installation*

After constructing the pilot, the UF module was installed on December 4, 1997. The installation went smoothly without damage to the UF module or the pilot system. With the entire system assembled for the first time, refinements were made to the COMFOWARD programming to assure normal operation. Start up included calibration of the sensors, system debugging, and purging air from the pilot plant.

Following the initial setup period, the system was cleaned with a solution of tap water (pH = 9.8, temperature = 33°C (91°F)) and 1% Ultrasil 59, an enzymatic cleaner. Thirty gallons of the cleaning solution were used according to the Aquasource instructions.

With the completion of the cleaning process, the pretreatment tank was scrubbed and refilled with tap water (pH = 9.8, temperature = 14°C (57°F)). The COMFOWARD system inputs were then set, and the system was allowed to run. The initial test to determine the intrinsic flux of the UF module with the tap water was conducted (Figure 5-1) . The intrinsic specific flux for the membrane was measured as:

$$J_{s,\, 20°c} = 235.25 \text{ LMH/bar} \qquad (5.1)$$

When compared to the manufacturer's specifications, this value was found to be acceptable (2% above manufacturer's value).

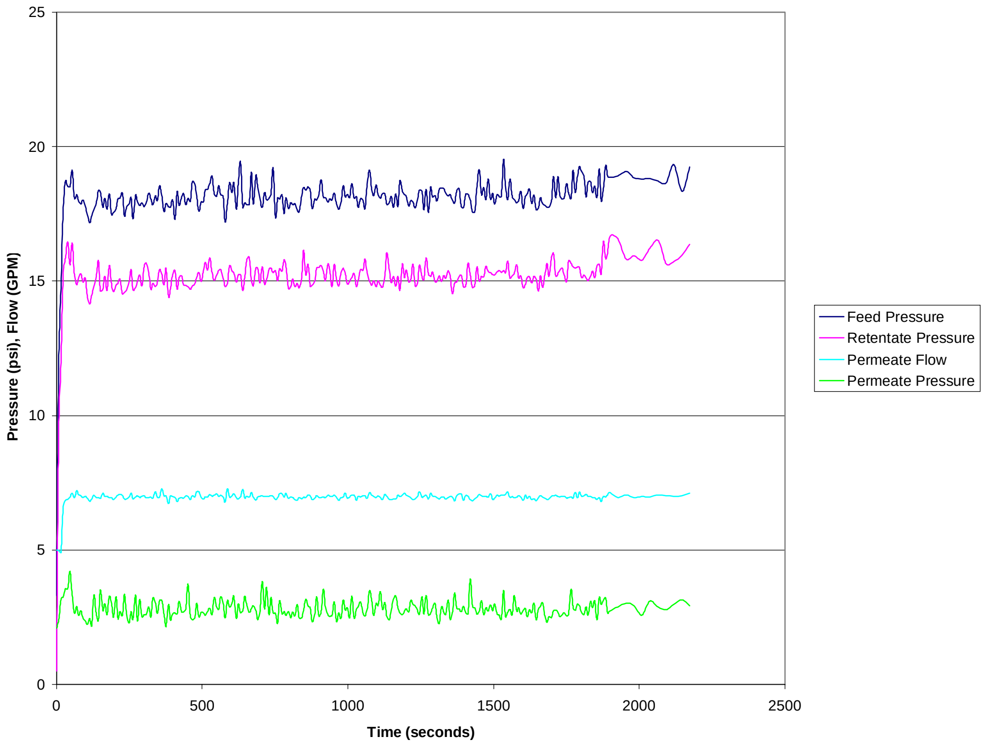

**Figure 5-1**
**Measured Clean Water Values for Determining Specific Flux**

Before testing the system with raw Lake Decatur water, a test with tap water was conducted to make adjustments to the backwash cycle. The tap water test used one minute backflush cycles every 1.5 hours of operation. After 11,000 L of flowthtrough volume (Figure 5-2 and Figure 5-3), an unexpected flux decline was noticed (Figure 5-3).

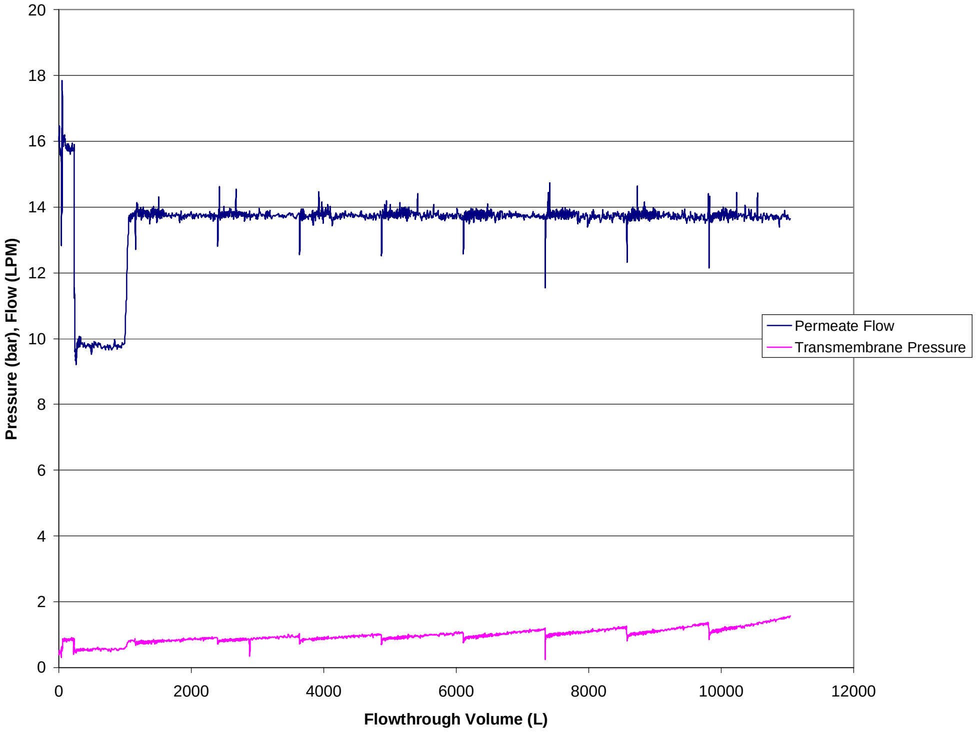

**Figure 5-2**
**Transmembrane Pressure and Flow Using Tap Water**

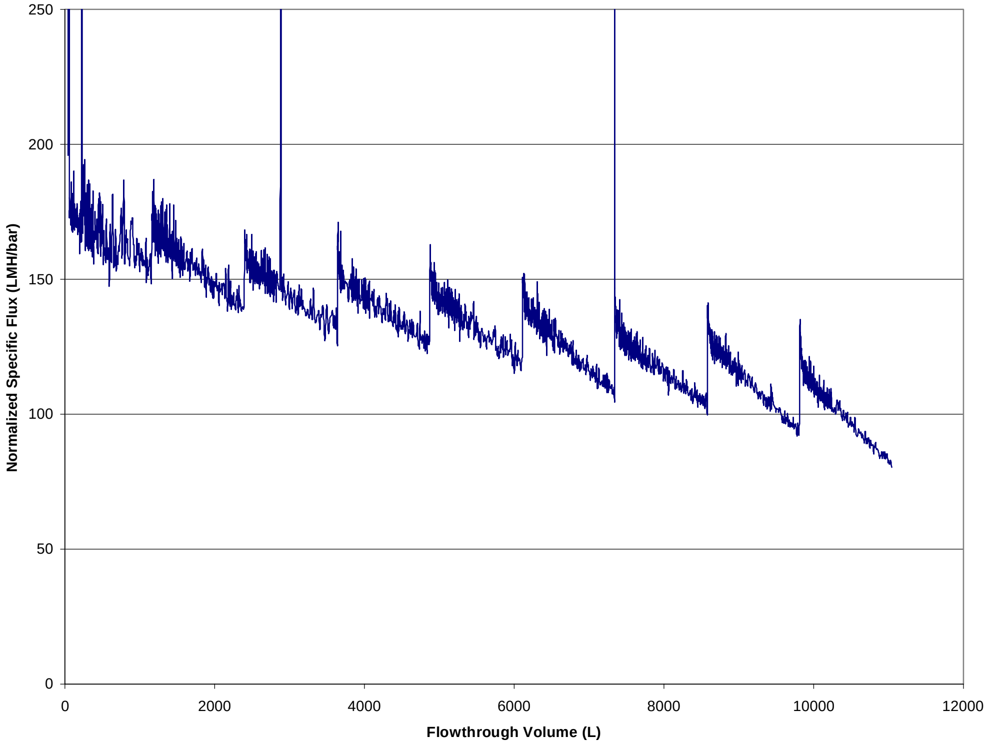

**Figure 5-3**
**Normalized Specific Flux at 20°C Using Tap Water**

One explanation for this flux decline is that by exposing the membrane to pH greater than 8.5 for more than 100 hours, irreversible fouling occurred (Glucina, 1998). Two other explanations for flux decline include compaction of the membrane and inadequate backwash. Due to the intrinsic properties of the membrane, clean water flux may decline before steady state conditions are met because the porous membrane surface compacts (Lee, 1998). The compaction creates resistance to flow that is indicated by losses in specific flux.

During the initial testing, the frequency and length of the backwash cycle may have also added to the initial flux decline. Even with tap water, some residual levels of TOC and particles may have led to the observed decline. To improve upon this, the backwash methods were modified as discussed in Chapter 4. Filtration of the Lake Decatur water would later prove to cause more significant losses in flux.

## UF Pilot Operation

The UF pilot plant was used as a source of water for the RO pilot. Testing with the UF unit proved to be a challenge because of the operator's inexperience and high turbidity in the feed water. Previous researchers at the University of Illinois had focused on UF pretreatment issues; however, these tests were conducted at the bench and pilot scale with small flows (≈300 mL/min) (Laîné, 1988; Hennegan, 1994; Marriot, 1996; Jack, 1997). The adjustments towards the end of March and early April, 1998, proved that the UF could be used to filter the raw water more reliably. Figures 5-4 through 5-7 show the operation of the UF pilot plant as a function of volume produced.

## Turbidity

Turbidity measurements (Figure 5-4) were continuously monitored using a flow-thru turbidity cell and turbidimeter with a 1-5V output to the COMFOWARD software/hardware. Throughout the project, this method produced sporadic results which did not always indicate the true performance of the UF module. The main reason for error in turbidity measurements came from air bubbles which entered the cell after backwash and during filtration. Frequent backwashing (every 15 minutes) did not improve the measurements during initial testing. Note that frequent backwashing during the broken fiber and low free chlorine tests (Figure 5-4) exhibited more intense and numerous high turbidity excursions. When backwash cycles were increased to one hour during normal operation (Figure 5-4), the turbidimeter measurements stabilized faster and produce more reliable results. This problem is to be expected with continuous turbidity monitoring.

### Broken Fiber Test

To see how a damaged fiber would affect permeate turbidity, a smaller 70-fiber element, with fibers identical to the DN100 module, was connected to the system for a small portion of the testing. One fiber out of 1000 was intentionally compromised by cutting it. With the compromised fiber in place, the UF permeate exhibited a minimum turbidity between 0.6 and 0.8 NTU. As indicated by a decrease in the minimum turbidity under the “Low Free Chlorine” section in Figure 5-4, the removal of the broken fiber immediately reduced steady state turbidimeter readings below 0.2 NTU. Since the smaller module’s permeate production was insignificant, the operation of the DN100 module was not affected.

A piece of the cut hollow fiber was extracted from a smaller module to see if surface visualization with AFM was feasible. The surface visualization of the wetted hollow proved to be too difficult (testing at the University of Illinois normally incorporated flat-sheet membranes). The flat sheet membranes used in bench scale studies are easier to handle and provide more reliable images using the AFM technique.

After the smaller module was removed for further studies, a spike of free chlorine (<50 mg/L) was added to the backwash permeate before the system was allowed to shut down. The DN100 membrane module was cleaned to remove any bacteriological contamination and to restore flux.

## Low Free Chlorine Test

After cleaning, an adjustment was made to the chlorine pump which proved to be detrimental to the UF permeate turbidity. Lower free chlorine concentrations (2-3 mg/L) resulted after the pump speed was accidentally decreased. With lower free chlorine concentrations in the backwash permeate, the effect on permeate turbidity could be detected. The amounts of free chlorine present still could have inhibited bacteria formation, but turbidity levels declined more slowly after restarting the pilot. Turbidity spikes of 1.2 to 1.8 NTU were more prevalent between 55000 and 64000 liters. Steady turbidity values were still reduced below 0.1 NTU during this test. Shortly after finding the chlorine pump out of adjustment and correcting it, turbidity measurements decreased (<0.04 NTU). The free chlorine concentrations were then measured at levels of 5-7 mg/L. After the problem was corrected, the system was chemically cleaned again to restore flux and reduce bacterial contamination.

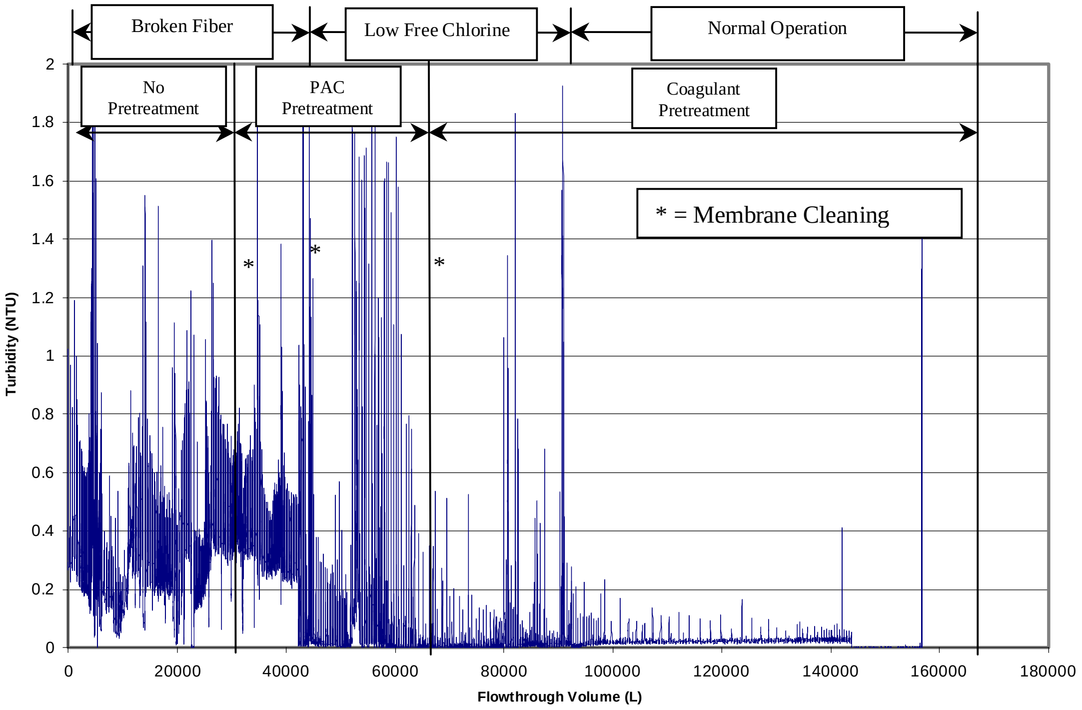

**Figure 5-4**
**Measured Turbidity of UF Permeate**

## Normal Operation

During normal operation, turbidity levels quickly returned below 0.1 NTU and no further problems were identified. Previous pilot studies and full scale plants, equipped with the Aquasource modules, have consistently produced permeate water with turbidities less than 0.1 NTU regardless of feed water conditions. After fine tuning the UF pilot to eliminate problems, it was necessary to show that this UF system could match or better the manufacturer's claims. With raw water turbidity ranging from 30 to 190 NTU, the DN100 module was able to produce at levels of 0.06 NTU for more than 95% of all measurements during normal operation.

## UF Without Pretreatment

With the new UF module in place, the effect of raw water filtration was observed (Figures 5-5 through 5-7). The UF system was set for constant flux mode. Backwashing occurred when the system reached a pressure of 28 psi (193 kPa). As operation proceeded, the backwash cycles became more frequent, decreasing from one hour down to 10 minutes. The length of the backwash cycle was set for 1.5 minutes at 34 psi (234 kPa). Specific flux declined from 150 LMH/bar to 40 LMH/bar within hours. Raw water turbidity ranged between 10 and 40 NTU. The feed water temperature averaged 3°C (37°F).

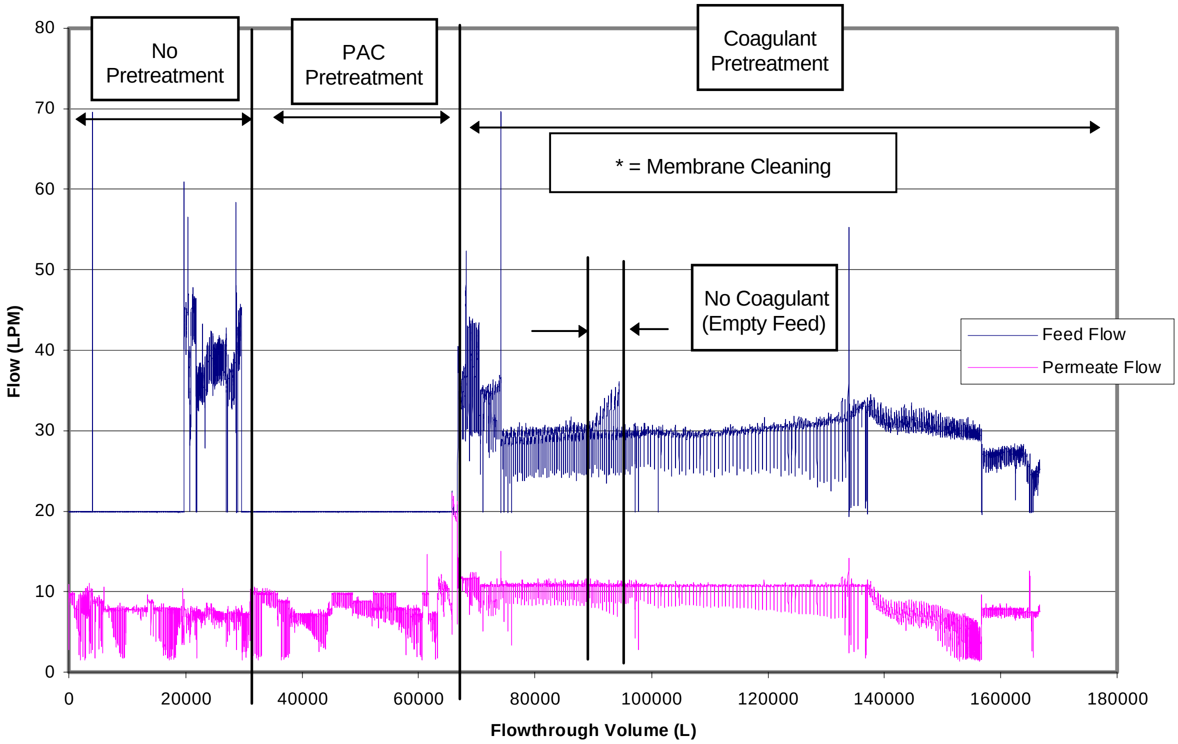

**Figure 5-5**
**Measured Feed and Permeate Flows from UF Pilot Using Raw Water**

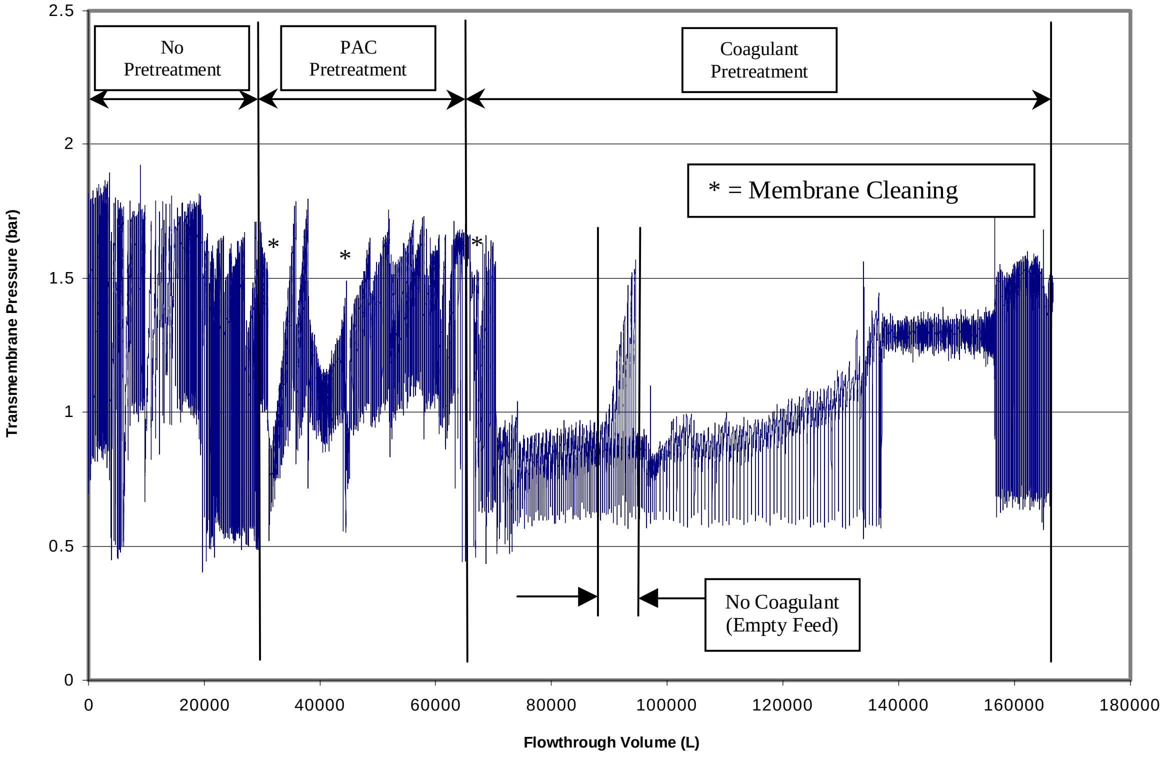

**Figure 5-6**
**Measured Transmembrane Pressure from UF Pilot Using Raw Water**

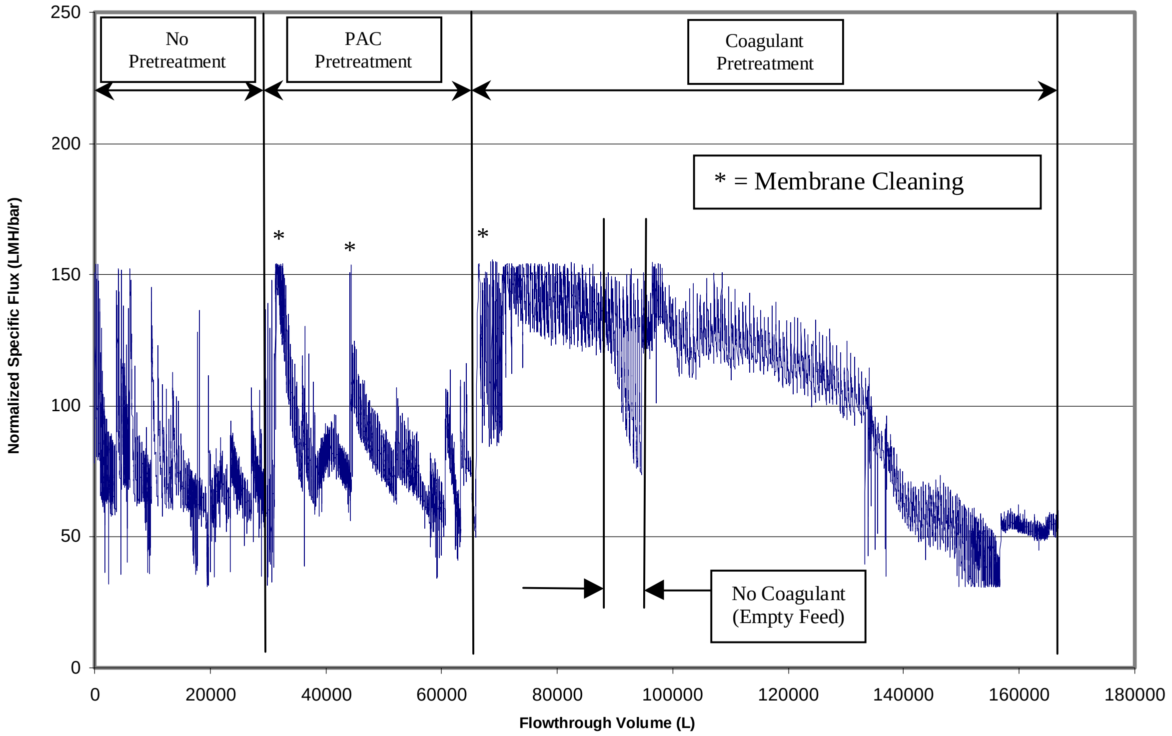

**Figure 5-7**
**Measured Normalized Specific Flux (20°C) from UF Pilot Using Raw Water**

After a shutdown of the UF pilot, the pilot was backwashed for 3 minutes. The combination of the long backwash and down time allowed some flux recovery. While the influent temperature was near freezing (Figure 4-5), the pilot system was contained in the basement of the SWTP at 17°C (63°F). This temperature difference may have slightly warmed the stored water enough to show a recovery of flux. However, further testing with the raw influent showed further losses in specific flux.

As a possible remedy to increase flux, cross-flow velocities in the system were brought to levels of 1 m/s or greater. By increasing the flow rate through the hollow fibers, the membrane surface is theoretically scoured by the high velocities, and cake formation is reduced. However, this change in cross-flow velocity did not significantly enhance filtration. This proved the need for pretreatment of the Lake Decatur water prior to ultrafiltration.

## UF with PAC Pretreatment

Removing NOM from surface waters prior to filtration has been shown to improve UF performance. Studies in both Decatur and Danville, Illinois, have implemented the UF-

PAC process with some success (Jack, 1997; Henneghan, 1994). Doses of 50-60 mg/L PAC were added into the pretreatment tank with a residence time of 1.6 hours. While this concentration of PAC would not be feasible for full-scale plant operation, the additional removal of NOM would prove if PAC could enhance UF performance.

During these tests, cross-flow velocities were held between 0.3 and 0.6 m/s using a recirculating pump. Membrane cleaning restored the raw water specific flux to 150 LMH/bar (10 LPM). After cleaning, the system was set for 15 minute backwash intervals (one minute backwashes). Dosing with the PAC was controlled by adjusting the peristaltic pump speed to add 60 mg/L PAC. The pretreatment tank mixer was set at 180 rpm to promote good mixing and prevent PAC sedimentation.

The COMFOWARD system showed a slight improvement in UF performance with PAC. Filtration runs lasted longer, but were still terminated within one day of operation. A rapid specific flux decline can be noted in Figure 5-7. During the PAC run, a second membrane cleaning recovered a majority (>90%) of the initial test flux. Flux decline after cleaning at 46,000 liters, was similar to filtration after 30,000 liters. The second PAC test was continued until the minimum specific flux was reached at 40 LMH/bar at 62,000L in Figure 5-7.

Even though operation was still impaired by low chlorine concentrations, filtration with PAC, in combination with more consistent operation, had slightly improved UF yield. Towards the end of the PAC test, Polyaluminum chloride (PACl) was added to the pretreatment tank to determine its affect on system performance. A dramatic increase in specific flux with PACl-coagulation was found, and this pretreatment showed the best overall improvement for operation of the UF pilot.

### UF with Coagulation

In the PACl-coagulation tests, 1 mL/min of the neat PACl solution was fed into the feed tank near the tank mixer. The A310 impeller in the pretreatment tank was set at 65 rpm for mixing the coagulant and allowing good floc formation. Pin-point floc could be observed in the pretreatment tank. The UF pilot parameters were set for two minute backwash cycles at one hour intervals and constant flux mode. Cross-flow velocities were also kept near 0.6 m/s for the test duration.

At a 30-40% module feed water recovery, based on Equation 3.7, the UF retentate was recirculated to the feed tank (Figure 4-21). A drain allowed for solids wasting at a rate of $Q_w$=4 LPM to reach a steady state-system recovery of 67%. The overall system recovery, $R_{s\text{-}UF}$, was calculated with Equation 3.7 by replacing the module feed flow, $Q_f$, with the system feed flow, $Q_s$. In recycle mode, the system feed flow was equal to the sum of the overall permeate produced (with backwash losses) and the waste from the pretreatment tank. The UF system feed flow can be calculated as follows:

$$Q_s = Q_p + Q_w \tag{5.2}$$

such that

$$R_{s-UF} = \frac{Q_p \cdot t_p - Q_b \cdot t_b}{Q_s \cdot t_p} \tag{5.3}$$

With good floc formation, the rate of irreversible fouling decreased. A combination of larger particle size (from floc formation) and settling of floc within the pretreatment tank could possibly explain the better performance. After backwash, flux recovery was near 100% of the initial test flux (from 65,000 liters to 95,000 liters, Figure 5-7). Flux decline during permeate production was also less significant than previous tests.

Without the PACl coagulation, as noted in Figure 5-7, the system operation was significantly affected. The PACl feed line broke during the test and stopped feeding the coagulant. Once the feed line was replaced, the system was reset and normal operation continued. Nominal flux losses were detected.

When specific flux decreased further, the system mode was changed to constant pressure. The UF module was run until flux decline reached the minimum of 36 LMH/bar. This pretreatment option had increased the overall volume of permeate produced by three times (100,000 liters) when compared to no pretreatment (30,000 liters) and PAC addition (34,000 liters). Table 5-1 summarizes the results from the UF pilot tests.

**Table 5-1**
**Summary of UF Tests**

| UF Test | Total Volume Filtered | Cleaning During Test | Mode of Operation | Backwash Interval (Backwash Cycle Time) |
|---|---|---|---|---|
| No Pretreatment | 30, 000 liters | No | Constant Flux | 15 minutes<br>(1 minute) |
| PAC Pretreatment | 34,000 liters | Yes (@12,000 liters) | Constant Flux | 15 minutes<br>(1 minute) |
| Coagulation Pretreatment | 100,000 liters | No | Constant Flux and Pressure | 1 hour<br>(2 minutes) |

## *UF Water Quality Testing*

### Total Organic Carbon

TOC removal with the PAC/UF process was not a specific goal of the project. However, the ability to remove TOC from RO feed waters may decrease fouling potential. TOC concentrations in the raw water ranged between 1.8 and 5.6 mg/L. Increases in TOC were consistent with increases in turbidity after precipitation. Figure 5-8 illustrates how the concentration of TOC varied over time.

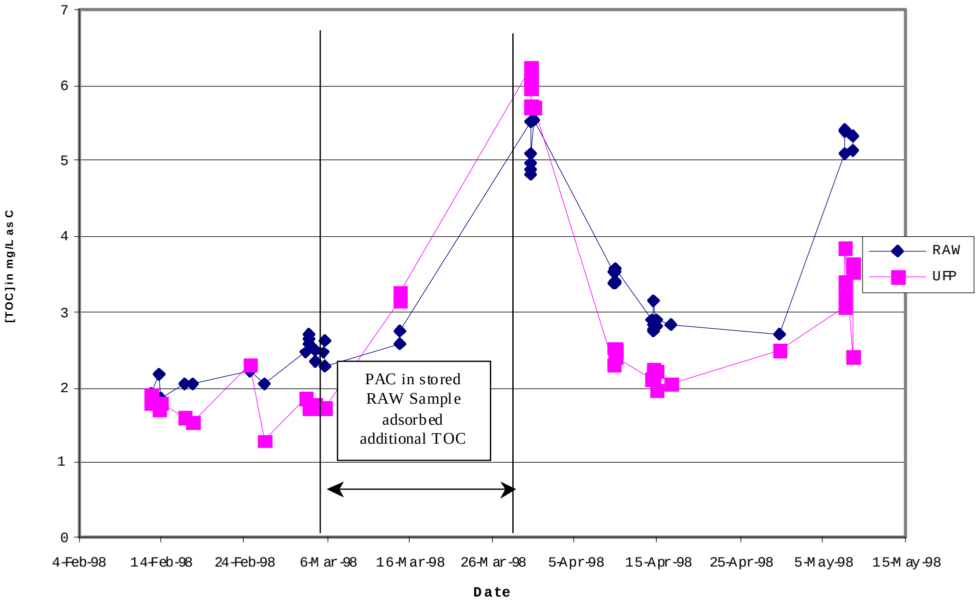

**Figure 5-8**
**TOC Concentration Between Raw and UF Permeate**

As seen in Figure 5-8, organic concentration to the influent water tracked the concentration in the UF effluent. Organic carbon removal (Figure 5-9) averaged 26.5%, with a range of 1% - 40% (excluding the section labeled "PAC in RAW Sample"). The raw water samples with lower TOC concentrations than the UF samples were contaminated with PAC present in the UF feed water. The additional TOC removal occurred because the PAC adsorption was not allowed to reach equilibrium within the UF pretreatment tank, but in the sample instead. Previous studies by Adham (1993) and Jack (1997) using similar Aquasource membranes showed TOC removals between 26.9% and 48% using 10 mg/L and 25 mg/L PAC respectively. The average here fell within these values using a higher dose of PAC.

The fact that removal efficiencies could have been improved with longer contact times leads to the belief that an additional 12% to 21% of TOC removal (Figure 5-9) could be achieved. This is evident in raw water samples contaminated with 60 mg/L PAC collected from the pretreatment tank. After reaching adsorption equilibrium, these results would then be more consistent with past experimentation.

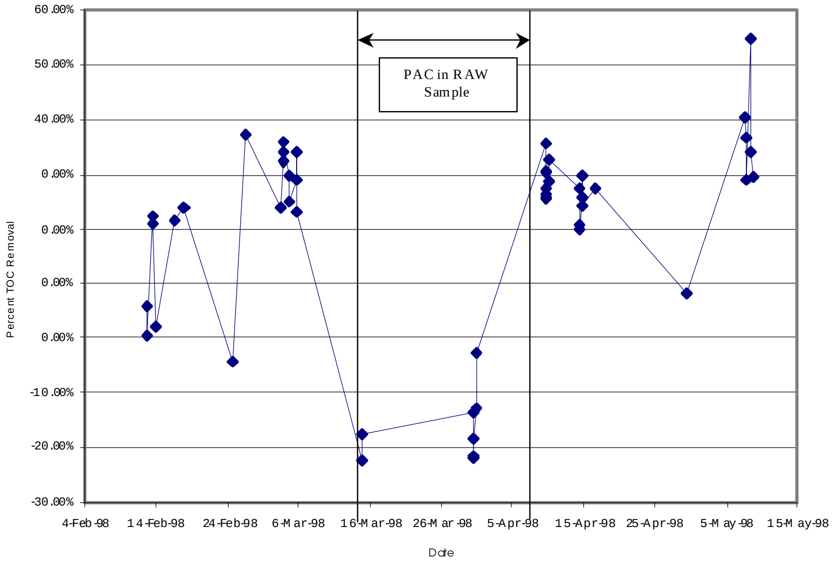

**Figure 5-9**
**Percent Removal of TOC from Raw Water by UF**

When comparing measured TOC values after both UF and conventional treatment, there appears to be no significant difference in TOC concentration (Figure 5-10). Tap water values, in some instances (March 3 and April 28), were lower than the UF permeate. Coagulation/sedimentation associated with lime softening and alum can be used to remove NOM (i.e. humic acids) (Montgomery, 1985). Thus, the removal of TOC with conventional treatment is not surprising.

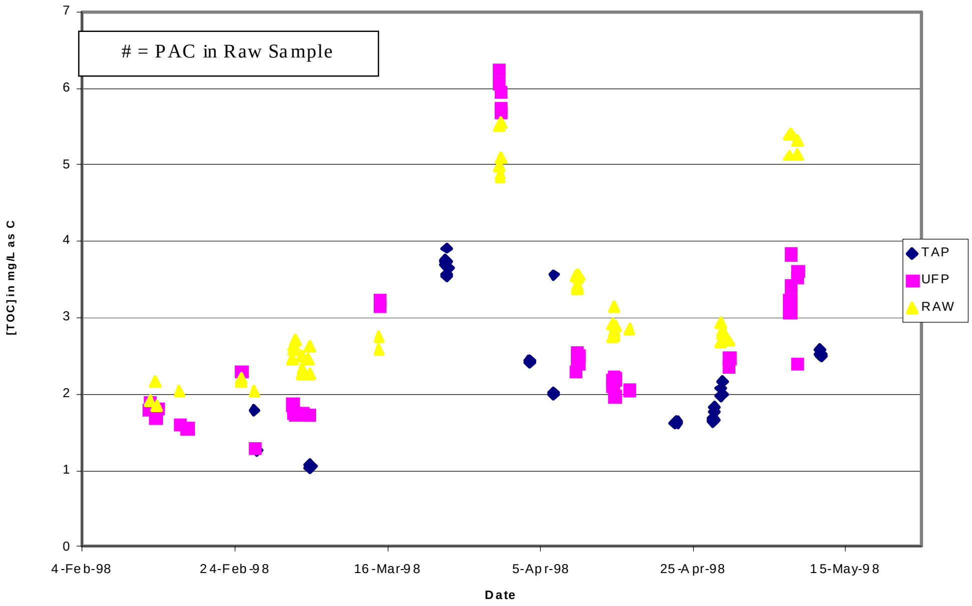

**Figure 5-10**
**TOC Concentration in Raw, UF Permeate, and Tap Waters**

## Atrazine

Since atrazine was in such low concentrations, the UF pilot testing could not provide conclusive information on atrazine removal. Atrazine levels were below 1 ppb for more than 90% of the testing period. In past years, measured concentrations of atrazine in the Sangamon River tend to peak in the beginning of May and slowly decrease to background levels by July (USGS, 1998). During the project, atrazine concentrations peaked on May 8, 1998 (Figure 5-11).

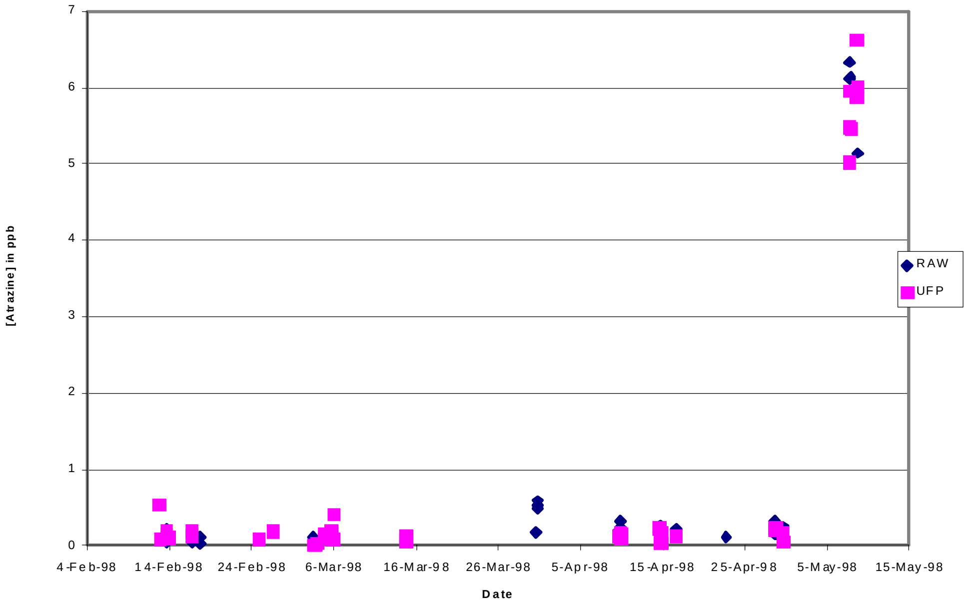

**Figure 5-11**
**Measured Atrazine Concentrations**

Without the addition of PAC, UF-atrazine removals were not detected over the length of experiments. The lack of removal is reasonable considering that atrazine (MW = 265 Daltons) is much smaller than the 100,000 Dalton MWCO of the UF module. Previous experiments with PAC-UF to remove atrazine have shown removals of up to 50% (Jack, 1997).

## UV Absorbance

UV absorbance tests were conducted at 254 nm-wavelength in part to monitor the system performance. As part of future regulations, specific UV absorbance (SUVA) measurements at 254 nm will indicate if a plant is required to practice enhanced coagulation. TOC and UV measurements were taken with filtered (0.45 micron) raw samples. SUVA measurements greater than 2 L/mg-m imply that enhanced coagulation is required. Raw water SUVA was consistently above 2 L/mg-m (Figure 5-12).

If a plant influent SUVA value is greater than the 2 L/mg-m limit, the treatment plant will be required to reduce the effluent SUVA value below 2 L/mg-m. SUVA measurements for the UF permeate and tap water were in great contrast with one another (Figure 5-13). Without PAC, SUVA values for the UF permeate were higher than the RAW measurements. TOC was removed with UF, as described previously, but UV absorbance was not reduced accordingly. The fraction of organic material contributing to UV absorbance may not have been removed, although this contrasts with expectations. Prior coagulation with PACl also did not improve the UF SUVA values. The additional removal of aromatic and smaller organic fractions from the water by conventional lime softening and alum coagulation at the SWTP may explain the decreased SUVA measurements. When compared with the UF permeate, the SWTP effluent met the SUVA standard (2 mg/L-m) more frequently than UF.

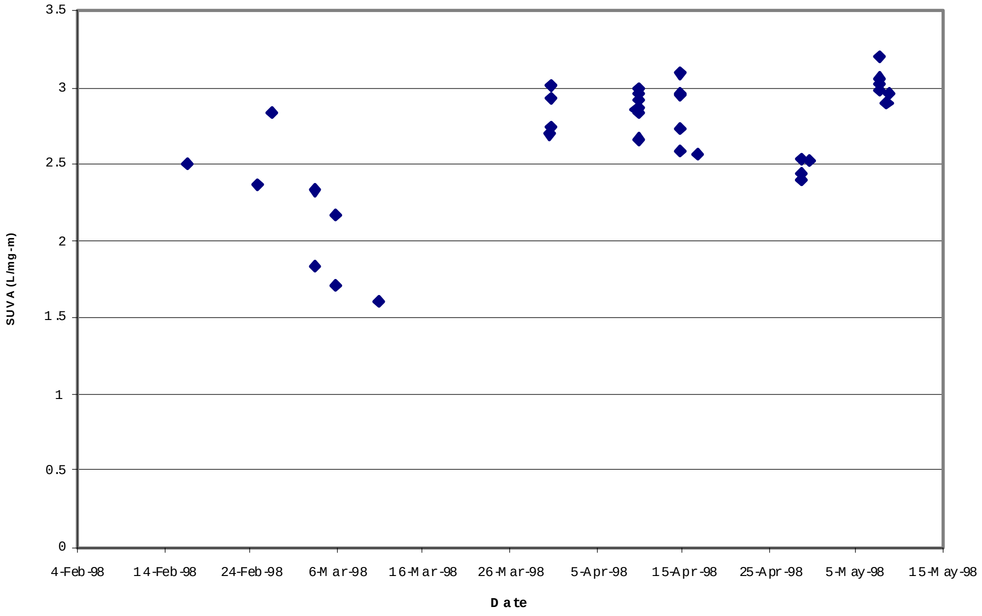

**Figure 5-12**
**Measured SUVA Values with Raw Water**

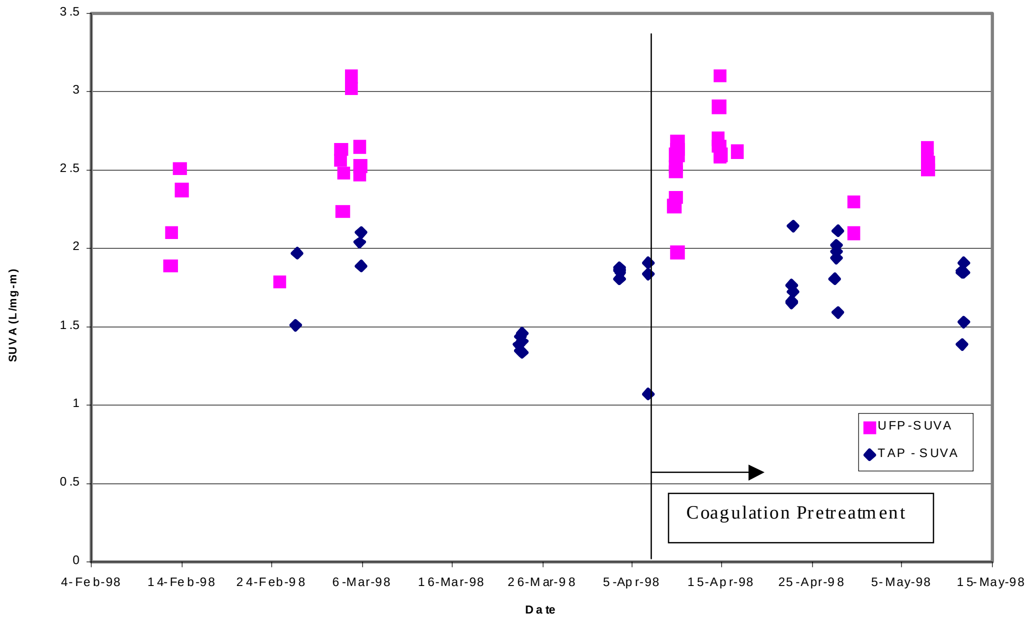

**Figure 5-13**
**Measured SUVA Values with UF Permeate and Tap Water**

## RO Pilot Results

### *Short Term Element Assays*

#### Overall RO Operational Results

The following tables and figures will depict the range of concentrations of contaminants measured before and after RO filtration. Figures 5-14 to 5-29 depict the combined measurements and rejections of all seven membrane elements with varying recoveries. The measured contaminants included nitrate, atrazine, TOC, alkalinity, total hardness, and calcium hardness. UV analysis was conducted at 254 nm as an indication of finished water quality and for calculating SUVA values. Table 5-2 summarizes the range of feed and permeate concentrations throughout testing. Table 5-3 represents the range of recoveries for the compounds measured.

## Measurements

### Temperature

The temperature of the RO permeate varied through the experiments (Figure 5-29), but did not affect DS removal. The ROSA and Winflows RO computer models also suggest that temperature differences from to 0°C to 25°C (32°F to 73°F) may not affect or even decrease DS removal. Using the computer models to predict nitrate removal, nitrate concentrations in the permeate varied by 0.1 mg/L $NO_3$ as N over the range of temperatures. While contaminant removal may not have been affected, the flux/recovery varied. For the purposes of comparison, the temperature correction formula (Equation 3.11) was used to adjust flux measurements to 25°C (73°F).

### Dissolved Solids—Hardness, Alkalinity, and TOC

As a secondary goal, the removal of DS (hardness, alkalinity, TOC) was measured for the RO/NF membranes. RO was effective in removing total dissolved solids from concentrations of 110 mg/L (tap) to 260 mg/L (UF) to levels less than 20 mg/L (Figure 5-28).

As pressure was increased on an individual element, DS concentrations (hardness, alkalinity, and TOC) tended to increase in the RO permeate. However, as the effective recovery increased, the combined results from the seven test elements showed decreases in DS removal (Figures 5-19 and 5-24). Hardness and TOC removals remained above 90% for recoveries from 10% to 50% regardless of membrane material and pressure. Alkalinity rejection was slightly lower than hardness rejection (Figure 5-25). At higher recoveries (44%-54%), the near-NF membrane (TFC-S) exhibited lower overall removal efficiency of hardness, alkalinity and TOC.

Results from previous experiments and numerical models are consistent with these trends for the evaluated constituents. Using a single stage system without retentate recycle, the increase in recovery over the span of RO elements was directly proportional to flux. Higher rejections with most RO elements require a more selective barrier that creates more resistance to solute passage. Membranes with higher flux capacity (e.g., TFC-S) were less selective and allowed passage of more dissolved solids than those with lower flux capacity (e.g., CPA2) for pressures between 100-200 psi (689-1379 kPa).

### UV Absorbance

For all RO permeate samples, UV absorbance at 254 nm-wavelength remained below 0.013 (Figure 5-20). With the high removal of TOC, UV values were expected to be near zero. These UV values were used to calculate SUVA values for the RO permeate

(Figure 5-21). The SUVA measurement evaluates the ratio of UV absorbance to TOC present in the product water.

The RO permeate SUVA values were within the range of 0 - 11 L/mg-m. For most of the elements tested, SUVA values were below 4 L/mg-m. The ESPA membrane exhibited SUVA values below 2 l/mg-m. Because both TOC and UV absorbents were reduced to such low levels in the RO permeate (Figures 5-18 and 5-20), any variations in either measurement could have caused the scatter in the SUVA measurements. Blending the RO water (UV<0.013, TOC<0.930 mg/L) with tap water (UV<0.051, TOC< 7-8 mg/L) would have very little effect in decreasing SUVA measurements.

As suggested by a strict interpretation of the M/DBP rule, enhanced coagulation may still be required prior to RO/tap water blending to reduce SUVA values below 2 L/mg-m. However, this seems unlikely with lower TOC concentrations (<2 mg/L) in the RO permeate.

### Nitrate

While hardness, alkalinity, and TOC removal were consistently high, nitrate rejection varied over the range of experimental recoveries and pressures (Figures 5-15 through 5-17). The cellulose acetate membrane showed poor removals (60% - 90%) of nitrate at low recovery (<15%). TFC-Polyamide membranes exhibited nitrate rejections above 90% for recoveries from 15% to 30%. Beyond 30% recovery, nitrate rejections decreased more rapidly. While there were occasional variances in the nitrate data, the suggested minimum nitrate rejections are indicated by lines drawn into Figures 5-16 and 5-17.

**Table 5-2**
**Summary of Measured Concentrations**

| Membrane | Sample | Nitrate (mg $NO_3$-N/L) | Atrazine (ppb) | TOC (mg/L as C) | Alkalinity (mg/L as $CaCO_3$) | Hardness (mg/L as $CaCO_3$) | Calcium (mg/L as $CaCO_3$) |
|---|---|---|---|---|---|---|---|
| TFC - HR | UFP | 5.2-7.5 | 0.03-0.23 | 1.968-2.225 | 202-220 | 283-300 | 170-176 |
| | TAP | 6.1-7.7 | 0.03-0.17 | 1.629-1.648 | 50-68 | 148-164 | 54-66 |
| | ROP | 0.005-0.4 | 0.004-0.06 | 0.092-0.361 | 2-8 | 2-6 | 0-4 |
| TFC - ULP | UFP | 6.4-7.3 | 0.01-0.19 | 1.732-1.873 | 246-262 | 346-372 | 184-220 |
| | TAP | 6.8-7.4 | 0.09-0.42 | 1.044-1.081 | 46-50 | 154-188 | 62-64 |
| | ROP | 0.55-1.7 | 0.001-0.07 | 0.034-0.930 | 4-8 | 6-12 | – |
| TFC - S | UFP | 6.4-7.4 | 0.12-0.16 | 2.290-2.533 | 186-194 | 268-282 | 154-166 |
| | TAP | 6.4-7.4 | 0.02-0.10 | 1.99-3.560 | 62-80 | 134-160 | 90-98 |
| | ROP | 1.2-3.0 | 0.002-0.05 | 0.149-0.316 | 6-20 | 10-34 | 4-18 |
| CPA2 | UFP | 5.2-6.6 | 0.54-0.10 | 1.288-2.297 | 172-288 | 268-480 | 184-248 |
| | TAP | 6.8-7.4 | 0.15-0.16 | 1.792 | 46-56 | 154-188 | 64 |
| | ROP | 0.15-0.9 | 0.03-0.11 | 0.061-0.378 | 3-4 | 1-6 | – |
| ESPA | UFP | 4.8-7.4 | 0.07-0.51 | 3.150-6.250 | 144-216 | 202-318 | 124-170 |
| | TAP | 6.0-7.1 | 0.10-2.03 | 3.549-3.904 | 84-88 | 156-178 | 128-132 |
| | ROP | 0.4-1.1 | 0.03-0.40 | 0.081-0.561 | 3-8 | 4-7 | -– |
| HR-CA | UFP | 6.9-7.5 | 0.05-0.26 | 1.983-2.479 | 222-232 | 330-358 | 183-192 |
| | TAP | 5.4-7.9 | 0.04-0.16 | 1.651-1.827 | 50-58 | 148-150 | 54-60 |
| | ROP | 1.2-2.9 | 0.055-0.16 | 0.080-0.744 | 4-6 | 6-20 | – |
| HR-PA | UFP | 7.0-9.0 | 5.02-6.62 | 2.351-3.841 | 116-136 | 198-234 | 116-132 |
| | TAP | 6.0-7.4 | 3.81-4.41 | 2.492-2.582 | 55-60 | 130-136 | 100-104 |
| | ROP | 0.2-1.1 | 0.040-0.42 | 0.144-0.258 | 3-4 | 4-9 | – |

**Table 5-3**
**Summary of RO Rejections**

| Membrane | Feed Water | Nitrate | Atrazine | TOC | Alkalinity | Hardness |
|---|---|---|---|---|---|---|
| TFC - HR | UFP | 94.7% - 98.6% | 63% - 76% | 91% - 96% | 98% - 99% | 98% - 99% |
| | TAP | 97.4% - 99.9% | 42% - 95% | 91% - 93% | 87% - 94% | 96% - 99% |
| | Recovery | 22% - 33% | | | | |
| TFC - ULP | UFP | 70.2% - 89.0% | 43% - 94% | 47% - 96% | 97% - 98% | 97% - 98% |
| | TAP | 89.7% - 92.4 | 30% - 94% | 83% - 97% | 91% - 92% | 100% |
| | Recovery | 37% - 52% | | | | |
| TFC - S | UFP | 55.9% - 70.0% | 73% - 97% | 87% - 96% | 89% - 94% | 88% - 94% |
| | TAP | 63.5% - 78.6% | 64% - 76% | 89% - 92% | 84% - 91% | 89% - 96% |
| | Recovery | 49% - 62% | | | | |
| CPA2 | UFP | 85.3% - 98.1% | 33% - 94% | 57% - 93% | 98% | 99% |
| | TAP | 91.3% - 97.8% | 63% - 65% | 90% - 97% | 92% - 94% | 99% |
| | Recovery | 12% - 36% | | | | |
| ESPA | UFP | 77.1% - 87.5% | 67% - 95% | 95% - 99% | 96% - 97% | 96% - 97% |
| | TAP | 82.0% - 94.4% | 58% - 95% | 85% - 91% | 91% - 95% | 97%-98% |
| | Recovery | 42% - 53% | | | | |
| HR-CA | UFP | 58.0% - 70.7% | 42% - 94% | 56% - 82% | 97% | 98% - 96% |
| | TAP | 63.0% - 83.1% | 13% - 59% | 83% - 96% | 92% - 96% | 95% - 06% |
| | Recovery | 8% - 11% | | | | |
| HR-PA | UFP | 86.4% - 93.7% | 92% - 98% | 90% - 97% | 97% | 97% - 99% |
| | TAP | 91.8% - 99.1% | 99% | 85% - 94% | 93% - 94% | 94% - 96% |
| | ROP | 15% - 28% | | | | |

Using the minimum nitrate rejections at each pressure, nitrate rejections appear to have been affected by the type of feed water used with the RO experiments. The UF permeate water had higher hardness and alkalinity levels (Figure 5-22) than the tap water (Figure 5-23). Higher ionic strength (IS) levels in the UF permeate (IS ≈ 0.006) compared to the SWTP effluent (IS ≈ 0.002) were a result of softening in the SWTP product water. The higher IS may indicate the ability for more nitrate to pass through the membrane via charge coupling.

While the charge coupling effect is not completely understood for multi-electrolyte solutions, single or bi-electrolyte solution experiments and theory suggest that ions must be paired (overall charge neutrality) before diffusing through the membrane (Mariñas, 1997). Ions such as calcium could theoretically pair with two nitrate ions and cross the membrane. With increased concentrations of calcium and magnesium, the charge coupling mechanism may have improved nitrate transport as $Ca(NO_3)_2$ versus $HNO_3$ and $NaNO_3$ across the membrane.

As a result, tap water tests showed slightly improved nitrate rejections (1% - 3%) with increasing recovery (Figures 5-16 and 5-17). This difference will be discussed more under Element Design Parameters and Summary.

### Atrazine

Atrazine concentrations in the raw water remained below 0.5 ppb from February to early May (Figure 5-27). The immunoassay test for atrazine was only reliable for concentrations between 0.5 and 5.0 ppb Atrazine. Below 0.5 ppb, however, the atrazine test still indicated removal using RO. Qualitatively, all RO permeate samples (Figure 5-26) were below the RO influent levels. Experiments with the HR-PA (Figure 5-27), showed excellent removals of atrazine (92% - 98%).

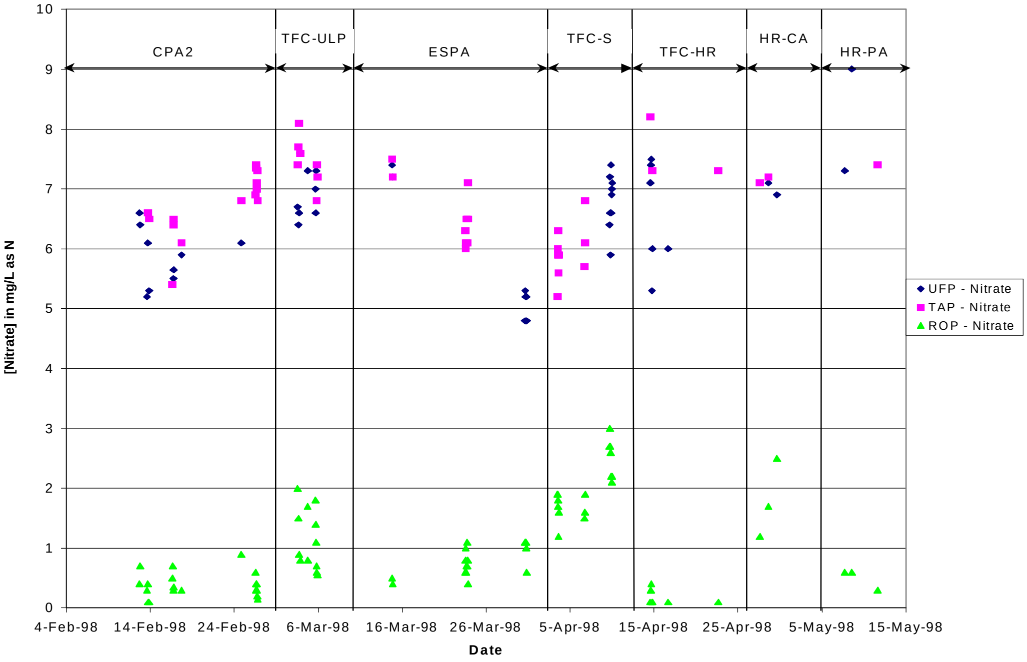

**Figure 5-14**
**Measured Nitrate Concentrations**

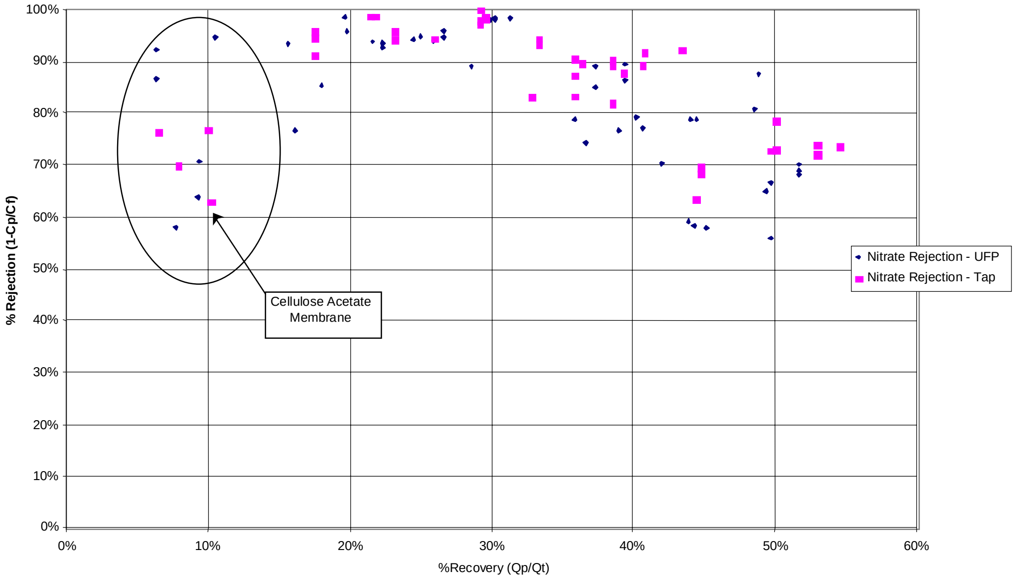

**Figure 5-15**
**Measured Nitrate Rejections with Feed Water Recovery**

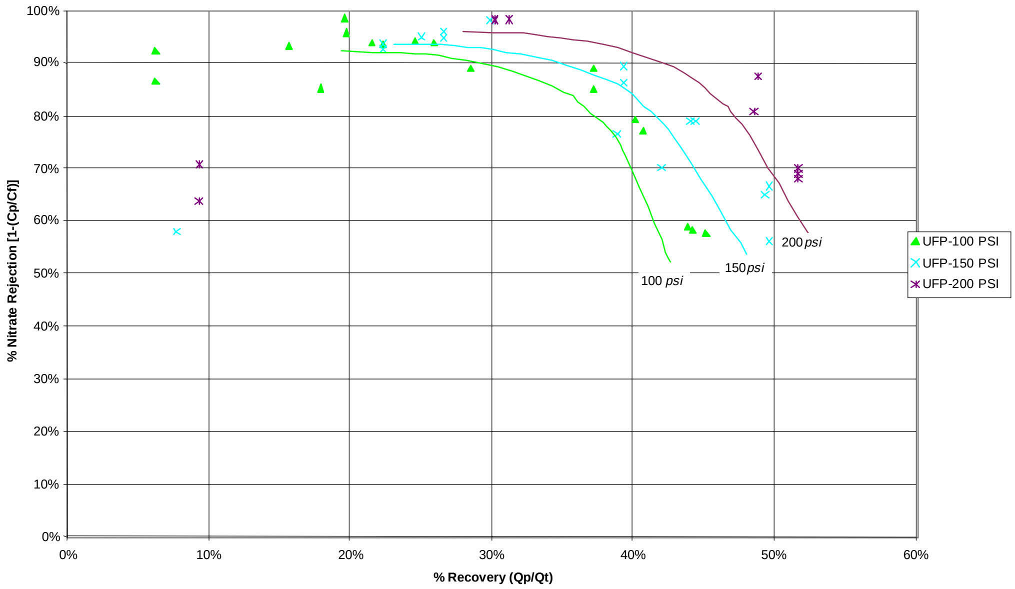

**Figure 5-16**
**Nitrate Rejection from UF Permeate over 100-200 psi (All RO Elements)**

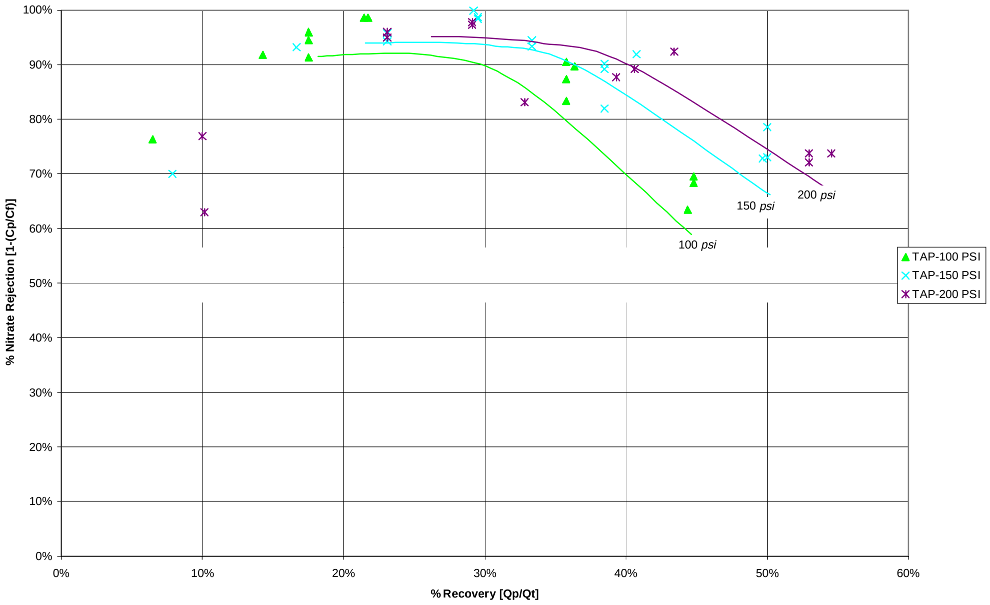

**Figure 5-17**
**Nitrate Rejection from Tap Water over 100-200 psi (All RO Elements)**

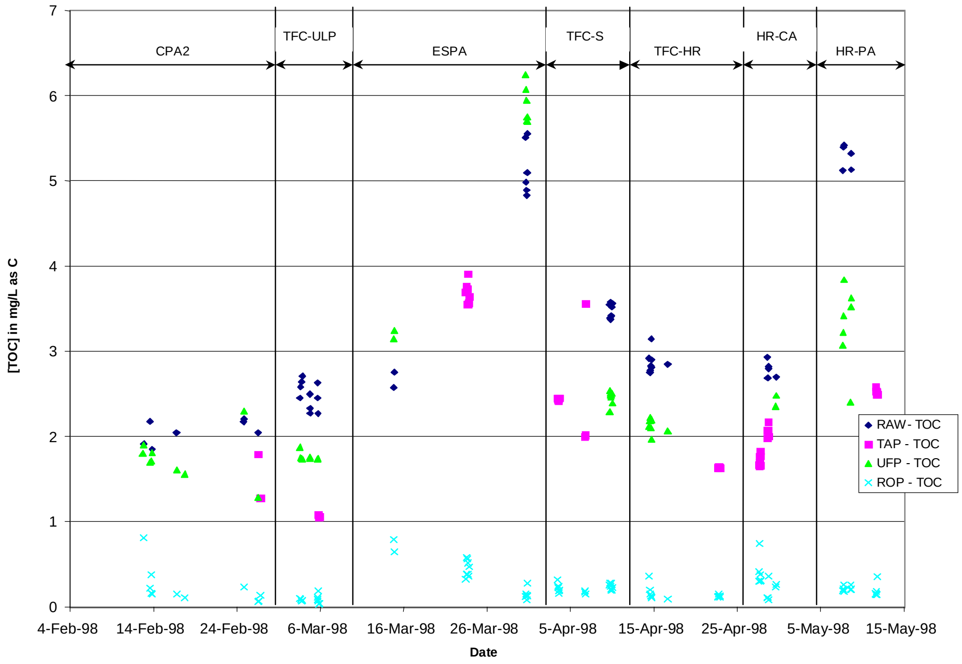

**Figure 5-18**
**Measured TOC Concentrations**

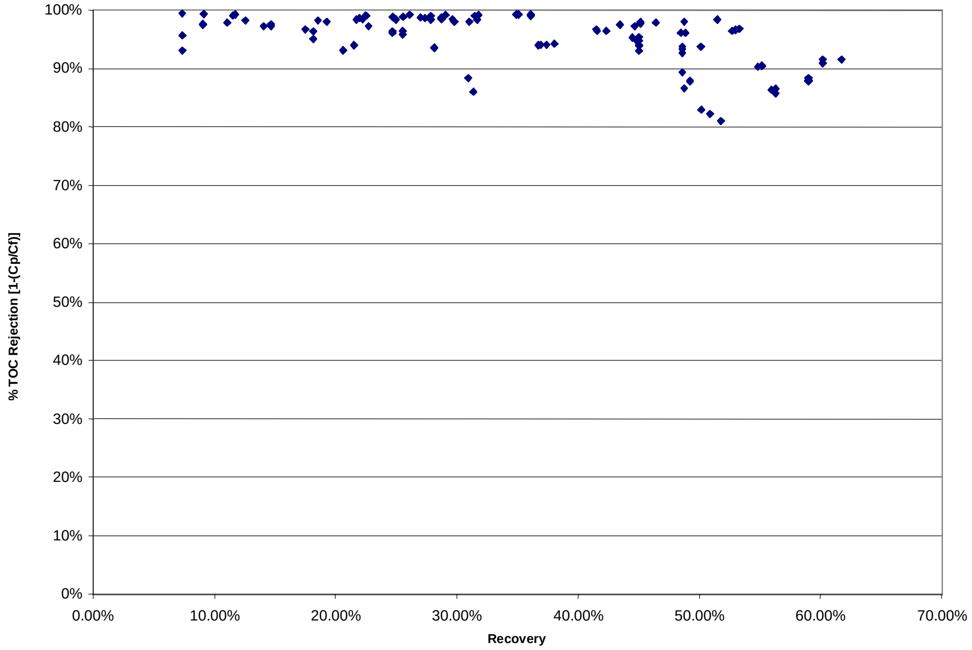

**Figure 5-19**
**Measured TOC Rejections with Feed Water Recovery (All RO Elements)**

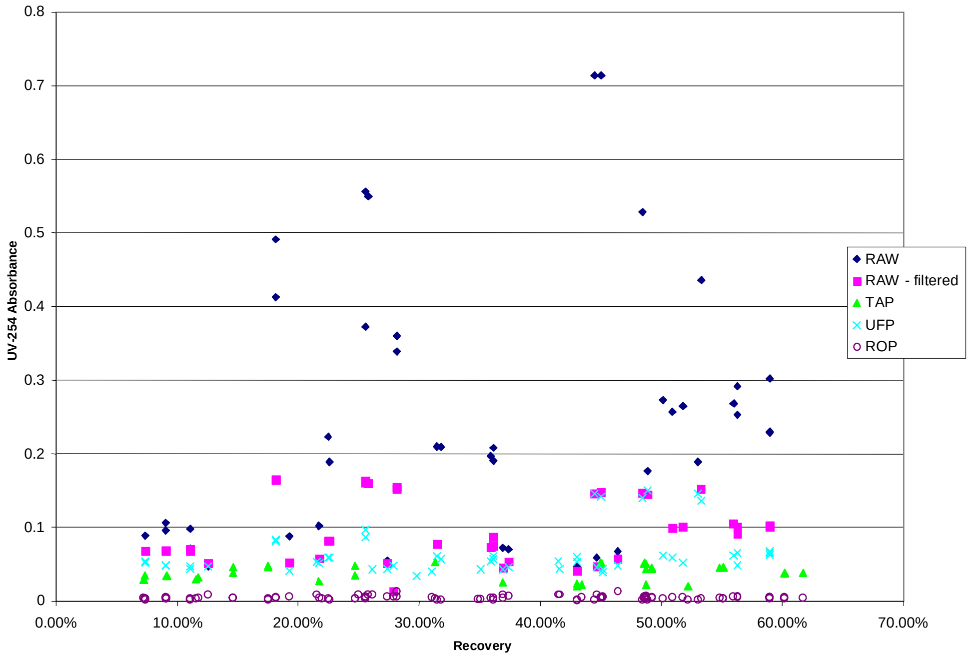

**Figure 5-20**
**Measured UV Absorbance (All RO Elements)**

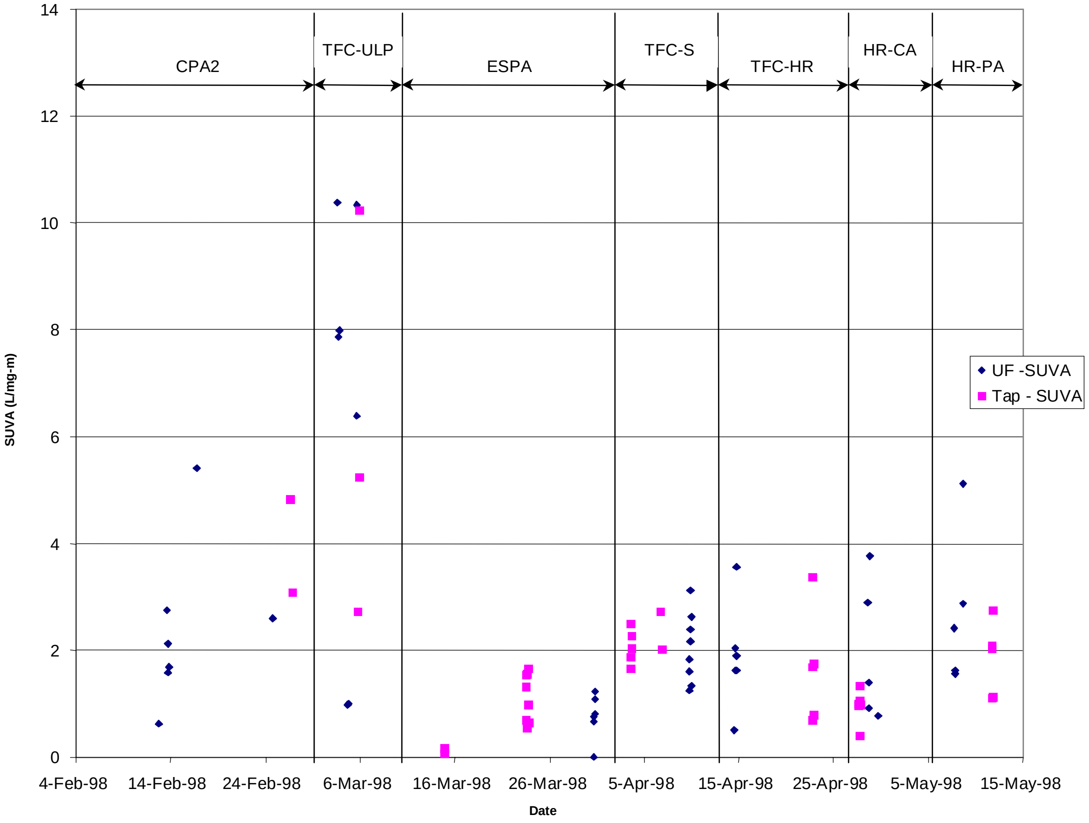

**Figure 5-21**
**RO SUVA Values**

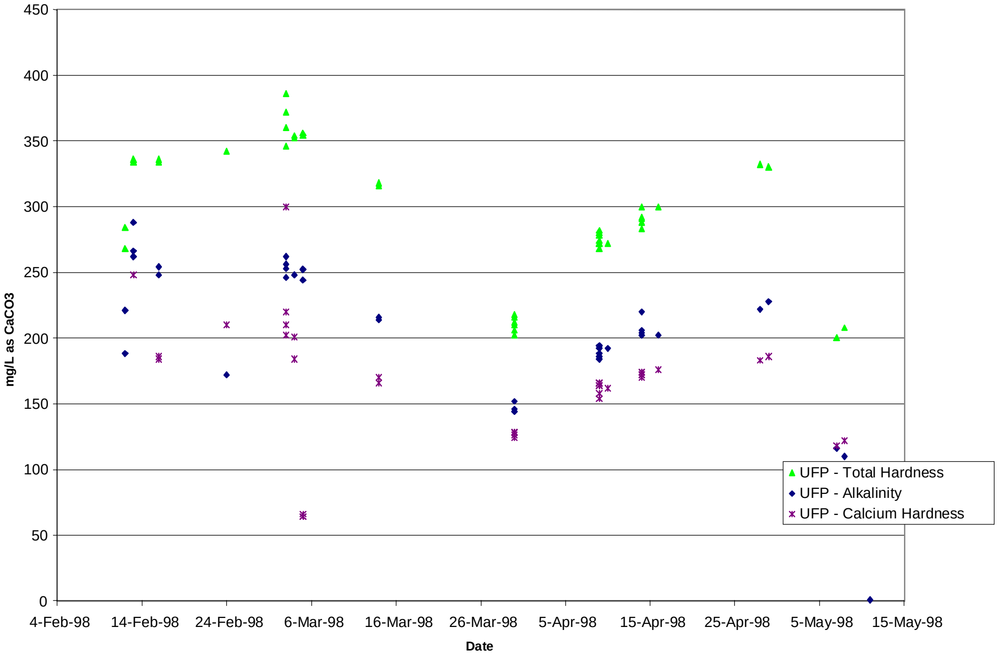

**Figure 5-22**
**Measured UF Permeate Water Alkalinity, Total Hardness, and Calcium Hardness**

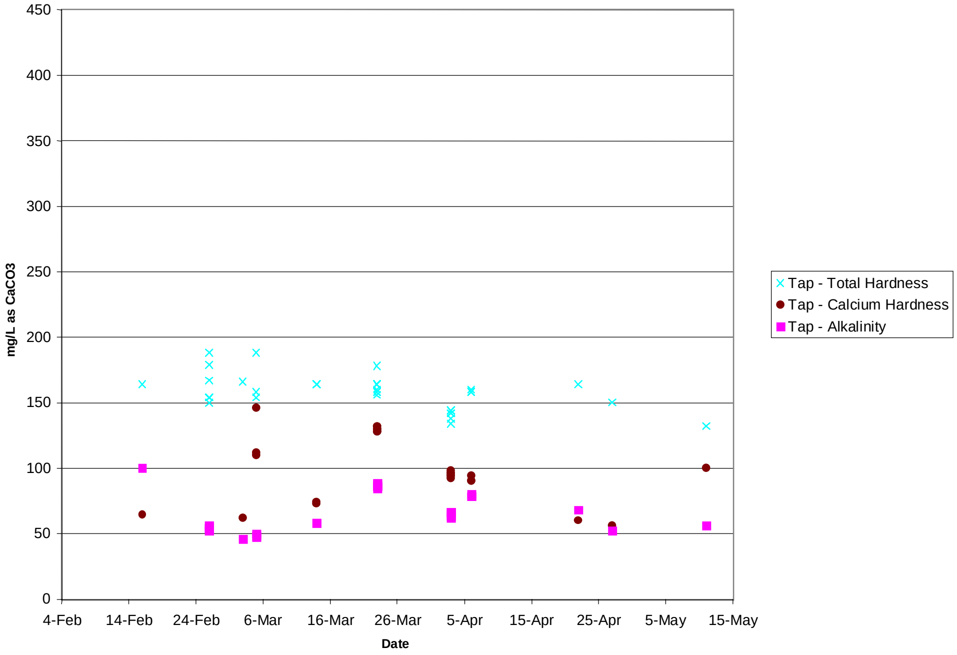

**Figure 5-23**
**Measured Tap Water Alkalinity, Total Hardness, and Calcium Hardness**

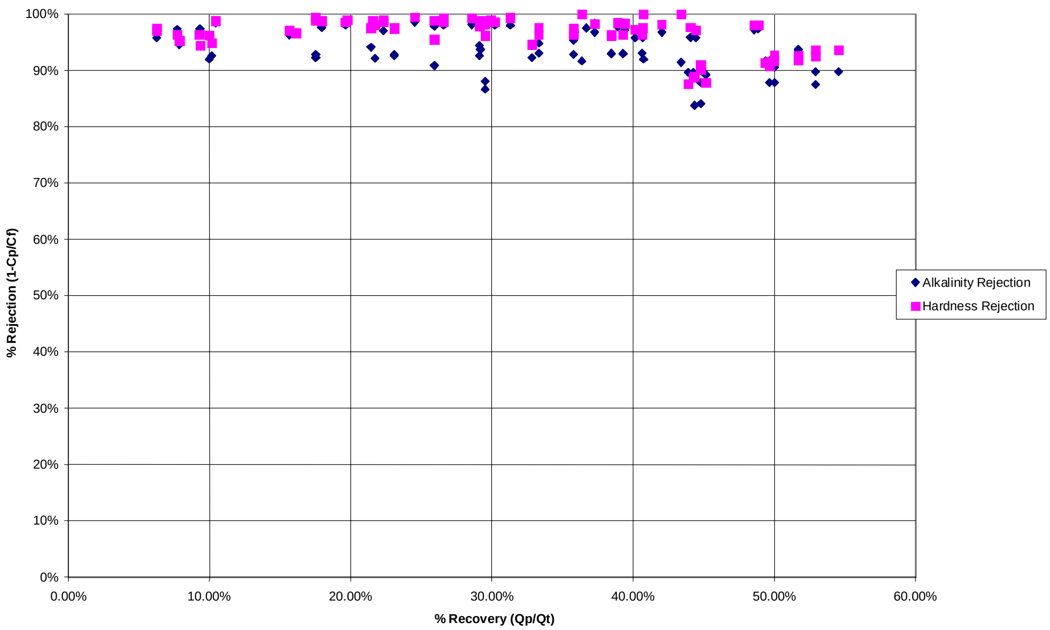

**Figure 5-24**
**Alkalinity and Hardness Rejection with Feed Water Recovery (All RO Elements)**

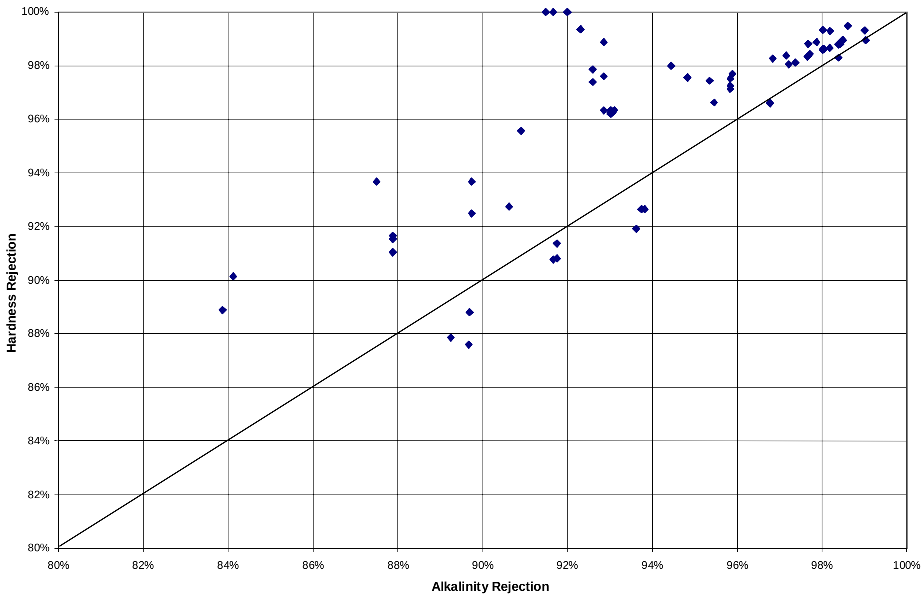

**Figure 5-25**
**Alkalinity Rejection Versus Hardness Rejection (All RO Elements)**

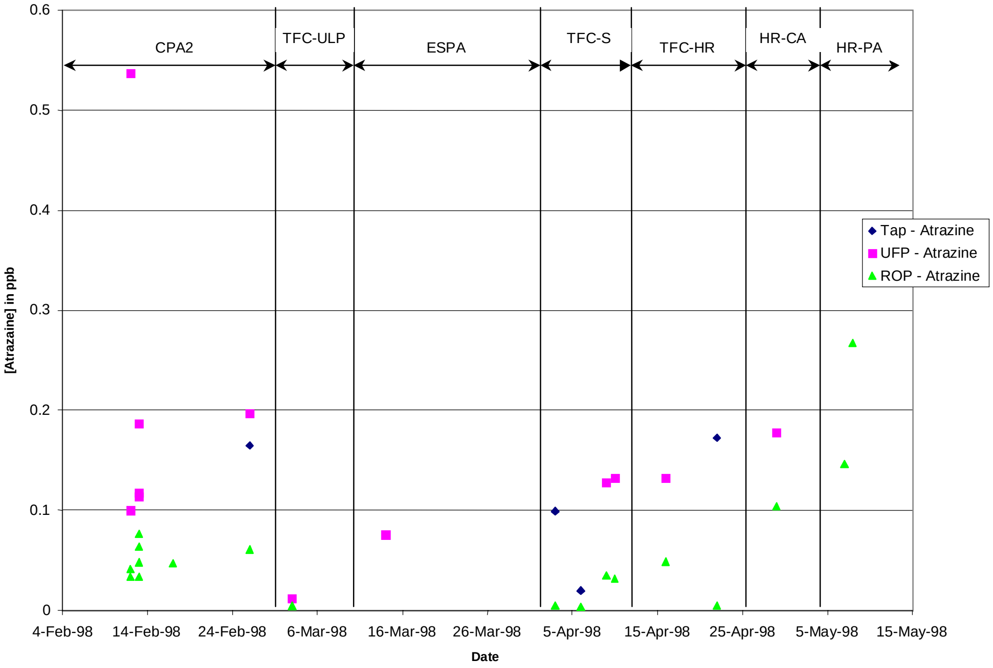

**Figure 5-26**
**Measured Atrazine Concentrations in Permeate**

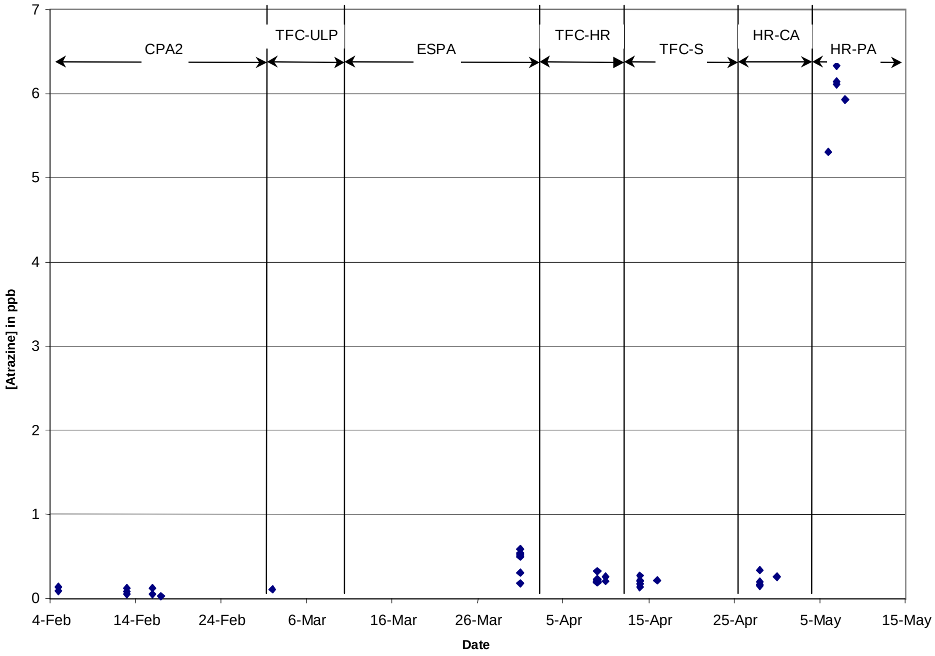

**Figure 5-27**
**Measured Raw Water Atrazine Concentrations**

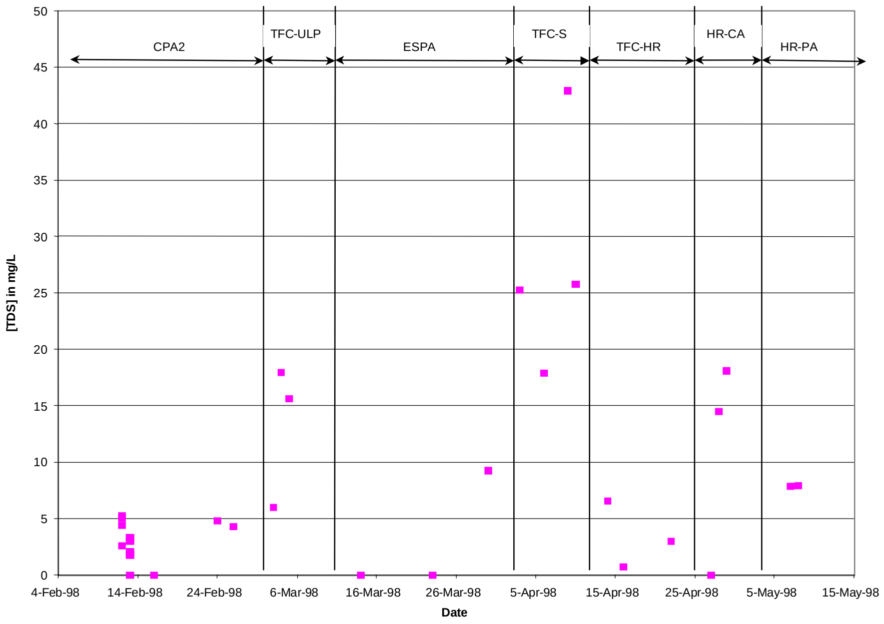

**Figure 5-28**
**Estimated TDS Concentrations in RO Permeate Based on Conductivity Measurements**

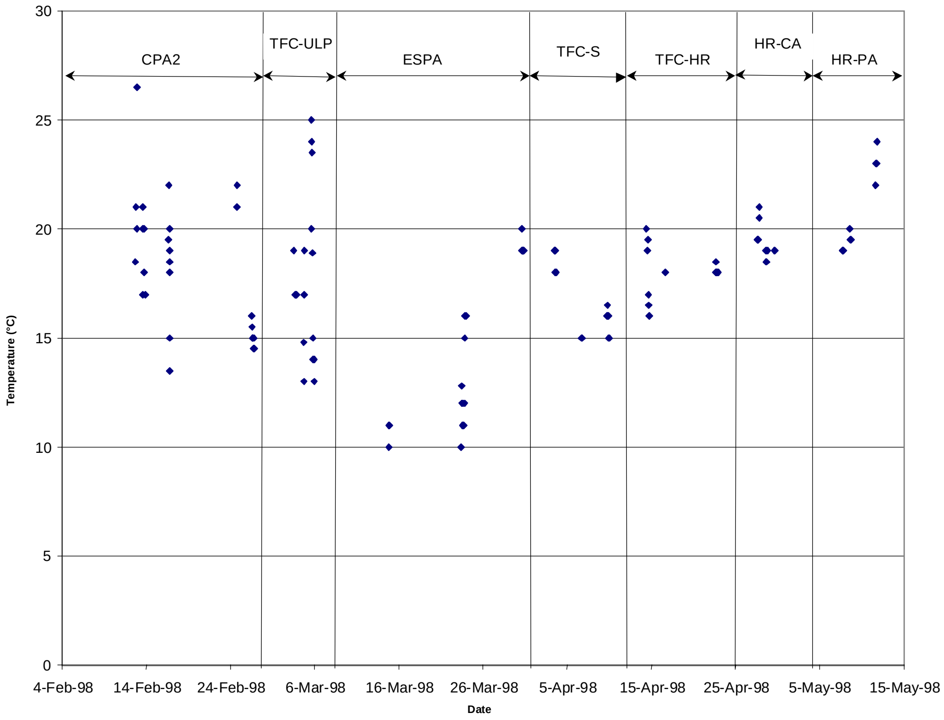

**Figure 5-29**
**Measured RO Permeate Temperature**

## Element Design Parameters and Summary

In order to design full scale RO treatment plants, design parameters for flux versus pressure and contaminant removal are important. From Equation 3.9, the water permeation coefficient, *A*, can be estimated by the linear slope between data points for a particular element when plotting the applied transmembrane pressure ($P_{tm}$ - $\sigma\Delta\pi$) against the measured flux ($J_v$). When comparing different elements, larger permeation coefficients indicate a membrane with higher specific flux.

Using Equation 3.10, the solute permeation coefficient, *B*, can be determined by the linear slope between data points for the same element by plotting the inverse element flux ($J_v^{-1}$) against the percent solute remaining in the permeate $(C_p/(C_f\text{-}C_p))$. For a given membrane, the solute permeation coefficient of different contaminants can be individually estimated. When evaluating several contaminants and elements, the larger "*B*" value indicates that the contaminant is more likely to permeate through the membrane. For this project, only the nitrate "*B*" values will be estimated.

Using the *y*-intercept from the plot of the inverse flux ($J_v^{-1}$) against the percent solute remaining in the permeate ($C_p/(C_f\text{-}C_p)$), the reflection coefficient ($\sigma$) can be determined. This parameter is used to estimate the ratio of diffusive to convective permeation within an element. The reflection coefficient was also only estimated for nitrate.

### Summary of Design Parameter Estimation

For each membrane element, parameters were estimated by graphing the appropriate data using Microsoft Excel. The data points were then fitted with a linear trendline to estimate the overall slope for each plot. Overall, differences in the design parameters can indicate important information about each element. Table 5-4 summarizes the estimated *A*, *B*, and $\sigma$ parameters.

In Table 5-4, the $R^2$ values for each trendline estimate overall error between the sample variance (*SSE*) and fitted line variance (*SST*) (Equation 5.4). $R^2$ values equal to one indicate a perfect fit between the data and the governing model.

(5.4)

$$R^2 = 1 - \frac{SSE}{SST}$$

where

$$SSE = \sum\left(Y_i - \hat{Y}_i\right)^2$$

and

$$SST = \left(\sum Y_i^2\right) - \frac{\left(\sum Y_i\right)^2}{n}$$

**Table 5-4**
**Summary of Design Parameters A, B, and σ**

| Membrane | Feed Water | A<br>$(R^2_A)$ | B<br>$(R^2_B)$ | σ |
|---|---|---|---|---|
| TFC-HR | UFP | 0.000015<br>(0.9950) | -0.000011<br>(0.0277) | – |
| | Tap | 0.000014<br>(0.9898) | 0.000010<br>(0.0594) | – |
| TFC-ULP | UFP | 0.000020<br>(0.9934) | 0.000338<br>(0.7620) | 0.925 (*) |
| | Tap | 0.000017<br>(0.9918) | 0.000186<br>(0.9635) | 0.9253 |
| TFC-S | UFP | 0.000027<br>(0.9911) | 0.001802<br>(0.5923) | 0.7243 |
| | Tap | 0.000027<br>(0.9874) | 0.000989<br>(0.5708) | 0.7493 |
| CPA2 | UFP | 0.000011<br>(0.7546) | 0.000070<br>(0.4811) | 0.983 (*) |
| | Tap | 0.000012<br>(0.9825) | 0.000060<br>(0.4435) | 0.9831 |
| ESPA | UFP | 0.000020<br>(0.9922) | 0.000498<br>(0.5037) | 0.8352 |
| | Tap | 0.000020<br>(0.9476) | 0.000149<br>(0.0439) | 0.8740 |
| HR-CA | UFP | 0.000005<br>(0.9938) | 0.000234<br>(0.8352) | 0.8810 |
| | Tap | 0.000005<br>(0.9975) | 0.000200<br>(0.8340) | 0.9798 |
| HR-PA | UFP | 0.000016<br>(0.9795) | 0.000107<br>(0.4541) | 0.9292 |
| | Tap | 0.000015<br>(0.9573) | 0.000056<br>(0.4268) | 0.9639 |

(*) - parameter estimated using intercept from tap water estimation

When comparing a set of RO elements, the water permeation coefficient, $A$, represents the ratio of applied pressure to flux for a given RO element. Higher $A$ values indicate that flux can be increased with smaller changes in pressure.

For each element, the $A$ parameter for both UF permeate and tap waters was very similar (±0.000001). The similar $A$ coefficients indicate that water permeation through the membrane was not different for the UF permeate and tap water. This is to be expected because the intrinsic membrane properties were virtually unchanged during each test. The high $R^2$ values (>0.94, for most tests) confirm that the data were linear and conformed well to Equation 3.9.

With higher solute permeation coefficients, *B*, RO elements will exhibit increased rejections with higher flux. For comparing the removal of solutes for a set of RO elements, *B* values closer to zero are indicative of membranes that exhibit constant solute passage regardless of flux. Membrane elements with a lower solute permeation coefficients may be preferred over other elements in situations requiring consistent permeate water quality regardless of pressure or flux.

Due to the variance in the nitrate rejections, the *B* parameter was less predictable. Differences in feed water quality over time affected the RO element rejection rate and nitrate measurements. This was expected because of the possibility for nitrate ions to couple in any combination with the different feed water constituents. However, all tests indicate that nitrate rejection is increased when using tap water in comparison with the UF permeate.

With the TFC-HR, a negative *B* value was estimated because the nitrate concentrations were greater with increasing flux. It should be noted that the estimated *B* value was the lowest measured throughout all experiments. During the long term testing with the TFC-HR element, further measurements at 200 psi (1400 kPa) showed that lower nitrate concentrations were possible (0.0 - 0.1 mg/L $NO_3$ as N). This would indicate a water permeation coefficient value nearer to zero. Inaccuracies in the *B* parameter estimation would also prevent the prediction of the reflection coefficient parameter ($\sigma$).

The reflection coefficient parameter estimates the ratio of diffusion through the membrane compared to convective transport for a single solute. Higher $\sigma$ values (>0.5) indicate that diffusion is the prevalent form of solute transport through the membrane. As $\sigma$ increases, the membrane also exhibits higher selectivity that results in higher solute rejections. The higher selectivity of a particular element could be more effective in removing a contaminant (i.e., higher nitrate rejections).

While the TFC-HR $\sigma$ parameter estimation was not possible, the remainder of elements showed somewhat mixed results. Those elements that showed large variances and negative values for the *y*-intercept value had unacceptable reflection coefficient values. To resolve this problem, the reflection coefficient was then set to match the tap water $\sigma$ coefficient. Because the intrinsic element properties are not necessarily changed using different feed waters, the reflection coefficient is believed to be constant. As shown by the ESPA, HR-PA, and TFC-S elements, there were only slight differences (<0.04) in the $\sigma$ value between the tap and UFP experiments.

### Summary of Element Performances

As part of the short-term element testing, the individual elements were run at pressures ranging from 100-200 psi (700-1400 kPa) without retentate recycle. This resulted in low recoveries (<50%) compared to an overall system that would recover more than 70% of the feed water per pass. At these lower recoveries, the short assays provided the information necessary to estimate rejection efficiencies for each RO element. Figures 5-31 through 5-44 illustrate the performance of each membrane.

**TFC-HR**

The TFC-HR membrane provided the best nitrate removal (Figures 5-30 and 5-43). Rejections of nitrate ranged from 94.7% - 99.9% with feed water recovery from 22% to 33%. The $A$ parameter was slightly (-0.0000015) below the $A$ parameter average (0.000016) for all elements. The $B$ parameter was estimated to be near zero. While $\sigma$ could not be determined, it would be expected to close to unity (1) because of the extremely high nitrate and overall salt rejections. Compared to the other elements, these parameter estimates indicate that the TFC-HR membrane produces water with consistently low nitrate concentrations with no advantages in energy savings compared to other membrane.

**TFC-ULP**

With the ability to operate at low pressures and high recoveries, the TFC-ULP membrane (Figures 5-32 and 5-33) offered energy savings with lower RO permeate quality (based on Table 5-2—Hardness and ROP TOC concentrations). Nitrate rejections ranged from 70.2% to 92.4% with recoveries from 37% to 52%. The $A$ parameter was slightly greater (+0.000025) than the average $A$ value. The $B$ values for tap and UF permeate feed waters were 0.000186 and 0.000338 respectively. The tap water solute permeation coefficient was much lower (-0.000150) than both the $B$ value average (0.000336) and the UF permeate $B$ value of 0.000338 (close to the $B$ average). The reflection coefficient, $\sigma$, was higher (+0.0292) than the average $\sigma$ value of 0.8961. Based on parameter estimates for this element, nitrate rejections would be dependent upon the feed water quality. The high reflection coefficient also indicates that permeate nitrate concentrations would be low. The higher than average water permeation coefficient may indicate a possible cost advantage using this membrane over other elements through this range of pressures (100-200 psi (689-1379 kPa)) if slightly lower permeate quality is acceptable.

**TFC-S**

The TFC-S softening element (Figures 5-34 and 5-35) represented a near-NF membrane (based on reduced rejections of calcium and magnesium). The TFC-S membrane showed the highest recoveries (49%-62%) with the lowest nitrate rejections (55.9%-78.6%). The $A$ parameter (0.00027) was the largest measured for all of the membranes. With tap water ($B$ = 0.000989) and UF permeate ($B$ = 0.001802) the estimated $B$ parameters for this membrane were the highest. The $\sigma$ value was the lowest of all values. Based on the difference between tap and UF permeate solute permeation coefficients, these extreme parameter estimates indicate that nitrate rejections would depend on the feed water quality. The reflection coefficient (0.7493) suggests that this membrane would produce permeate with the lowest water quality of all seven membranes. This result is comparable to the measured nitrate rejections (55.9% - 78.6%). However, this membrane would have the greatest cost advantage over other elements because of its ability to produce more permeate with small increases in pressure.

### CPA2

The CPA2 element (Figures 5-36 and 5-37) showed fairly high nitrate rejections in the range of 85.3% - 98.1% with feed water recovery from 12%-36%. The *A* parameter was below (-0.000004), the *A* parameter average. The *B* parameter was also lower than the *B* parameter average (0.000336) for all elements. The membrane exhibited the highest $\sigma$ value (0.983) of the remaining six elements. The CPA2 parameter estimates suggest that this membrane's permeate would have consistently low nitrate concentrations regardless of feed water quality with higher energy costs compared to other membranes for similar increases in productivity (flux).

### ESPA

Tests with the ESPA element (Figures 5-38 and 5-39) were similar to those of the TFC-ULP membrane. Nitrate rejections ranged from 77.1% to 94.4% over a range of recoveries from 42% to 53%. The water permeation coefficient, *A*, was above average (+0.000004) for this element. The *B* values for both the tap (0.000149) and UF permeate (0.000498) were below and above the average (0.000336) respectively. The reflection coefficient, $\sigma$, (0.8546) was below the average (0.8961). Using both the tap and UF permeate water sources, the difference in the *B* parameter values shows that nitrate rejections are dependent on feed water quality. Moderate removals of nitrate (>77.1%) are possible with this membrane as suggested by the average reflection coefficient. The ESPA membrane may have a cost advantage over other elements because of its above-average ability to produce more permeate with small increases in pressure.

### HR-CA

The HR-CA membrane (Figures 5-40 and 5-41) was the only cellulose acetate membrane used during the short term tests. While nitrate rejections ranged from 58.0% to 83.1%, experimental recoveries were the lowest (8% - 11%) with this membrane. The *A* parameter (0.000005) was the smallest measured for all membranes. The *B* parameter (0.000217) was slightly lower than the overall *B* value average (0.000336). The reflection coefficient showed the largest variation (0.0988) of all other elements. While HR-CA nitrate rejections were lower than other RO elements, the solute rejections do not appear to be dependent on feed water quality. Due to its high variation, the estimated reflection coefficient could not be used by itself to determine if the HR-CA could remove nitrates. The small water permeation coefficient indicates this membrane would be at a disadvantage for energy costs because it would require larger pressure increases to match increases in permeate flow.

## HR-PA

Similar to the CPA2 element, the HR-PA element (Figures 5-42 and 5-43) was able to reject 86.4% to 99.1% of nitrate feed concentrations. The *A* parameter for this membrane represented the average overall value (0.000016). With the tap water *B* value (0.000056) lower than the UF permeate *B* value (0.000107), there is a possibility that feed water quality will affect the HR-PA element's ability to remove nitrate. The reflection coefficient was above average (+0.0331). The HR-PA parameter estimates suggest that this membrane's permeate would have consistently low nitrate concentrations regardless of feed water. Compared to the other six elements, no advantage or disadvantage exists for energy costs savings using this membrane (average *A* value).

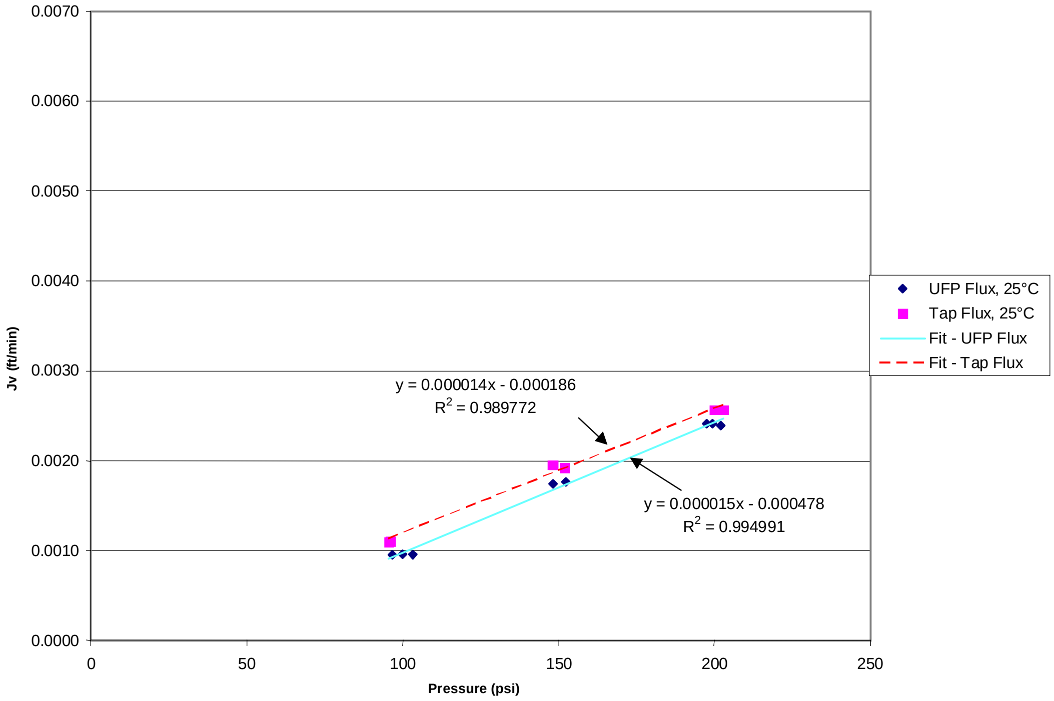

**Figure 5-30**
**TFC-HR Element—"A" Parameter Estimation**

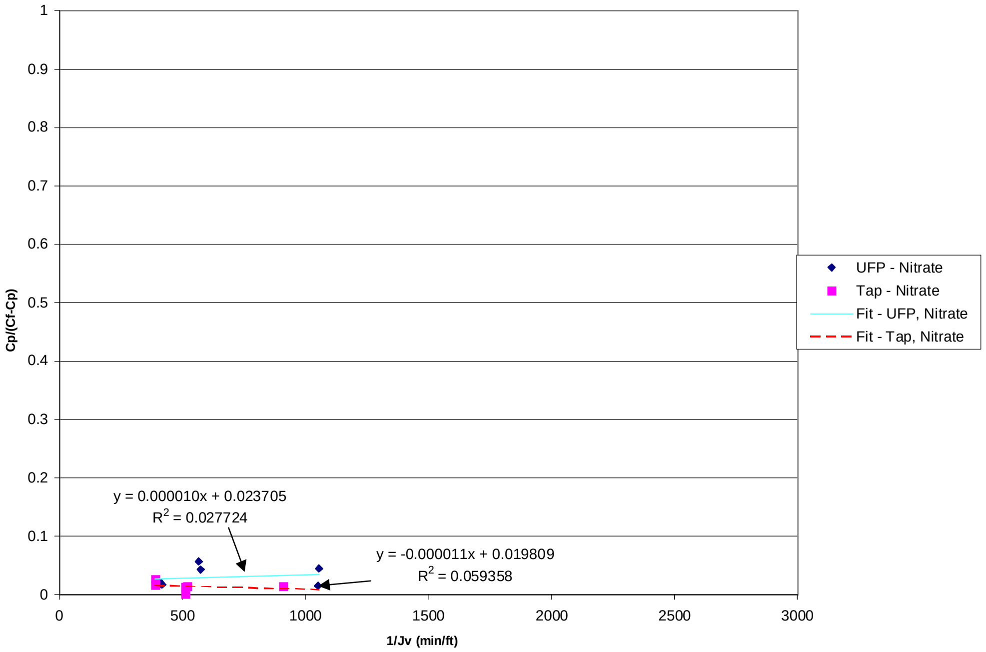

**Figure 5-31**
**TFC-HR Element—"B" Parameter Estimation**

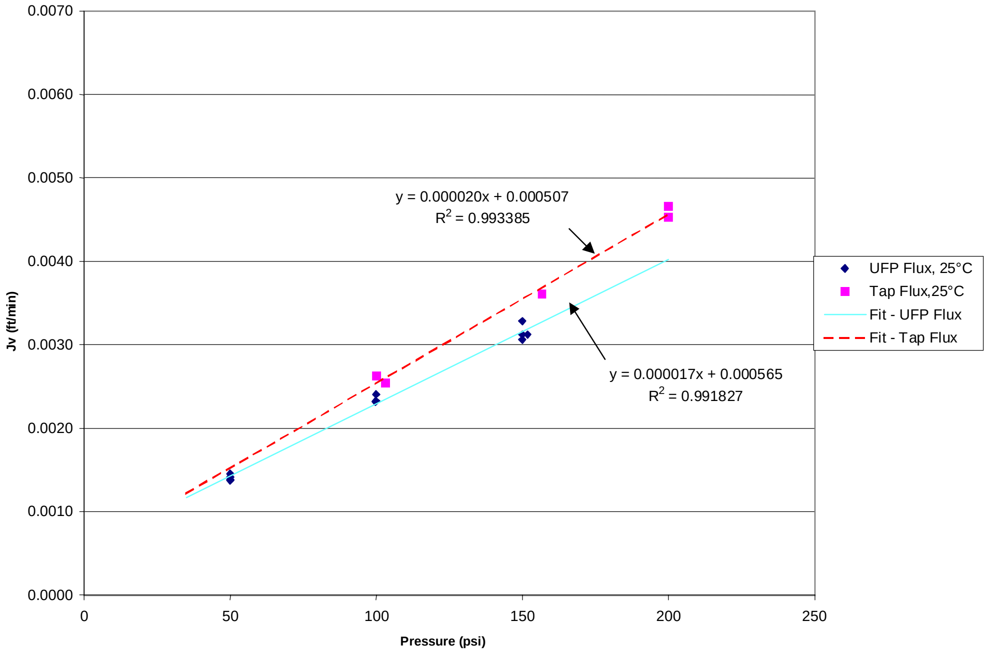

**Figure 5-32**
**TFC-ULP Element—"A" Parameter Estimation**

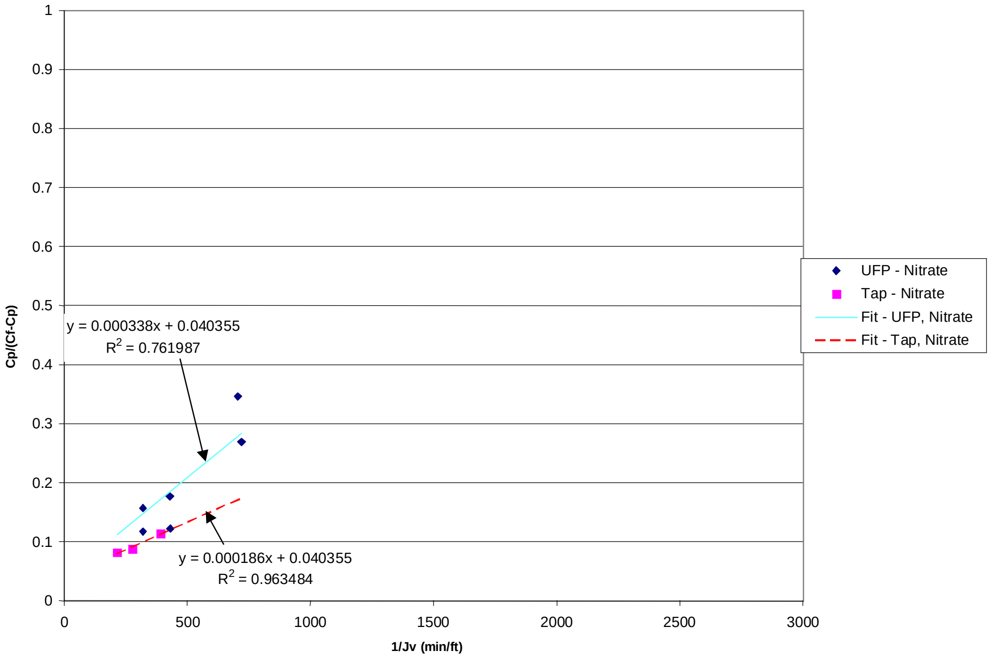

**Figure 5-33**
**TFC-ULP Element—"B" Parameter Estimation**

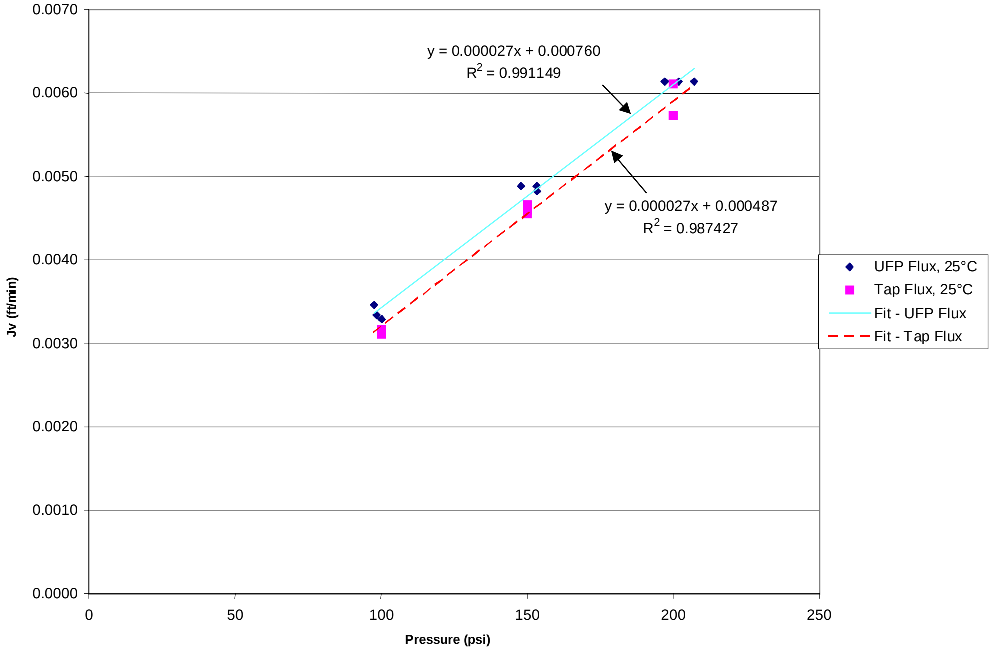

**Figure 5-34**
**TFC-S Element—"A" Parameter Estimation**

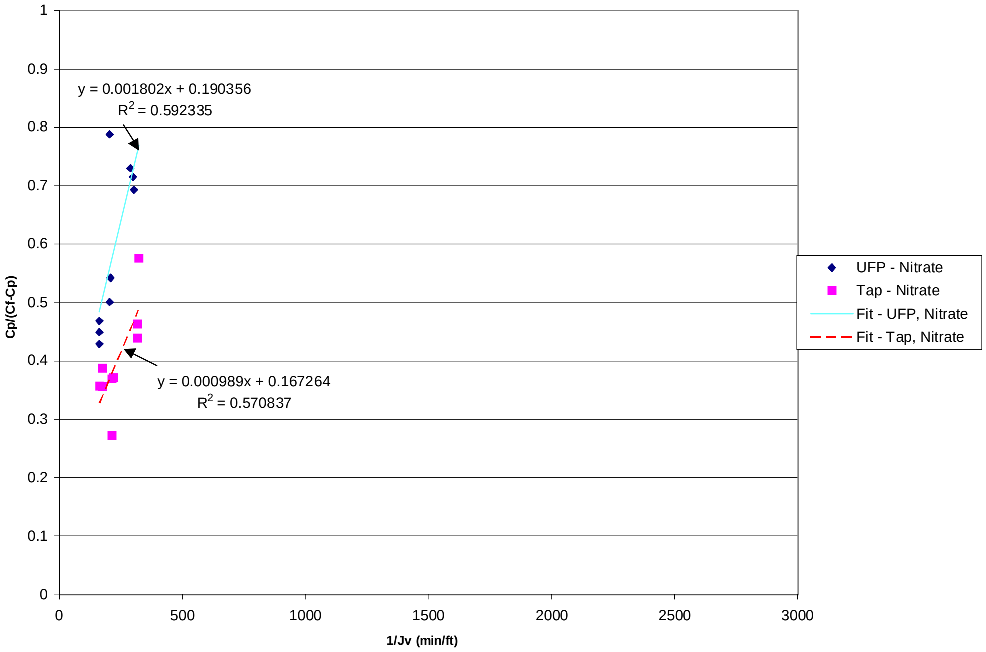

**Figure 5-35**
**TFC-S Element—"B" Parameter Estimation**

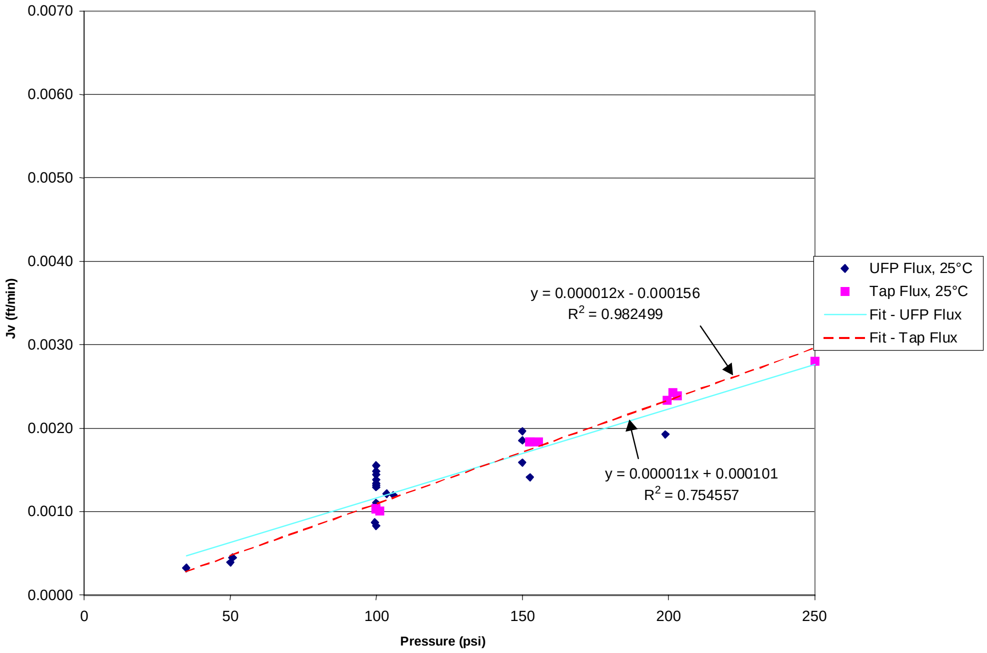

**Figure 5-36**
**CPA2 Element—"A" Parameter Estimation**

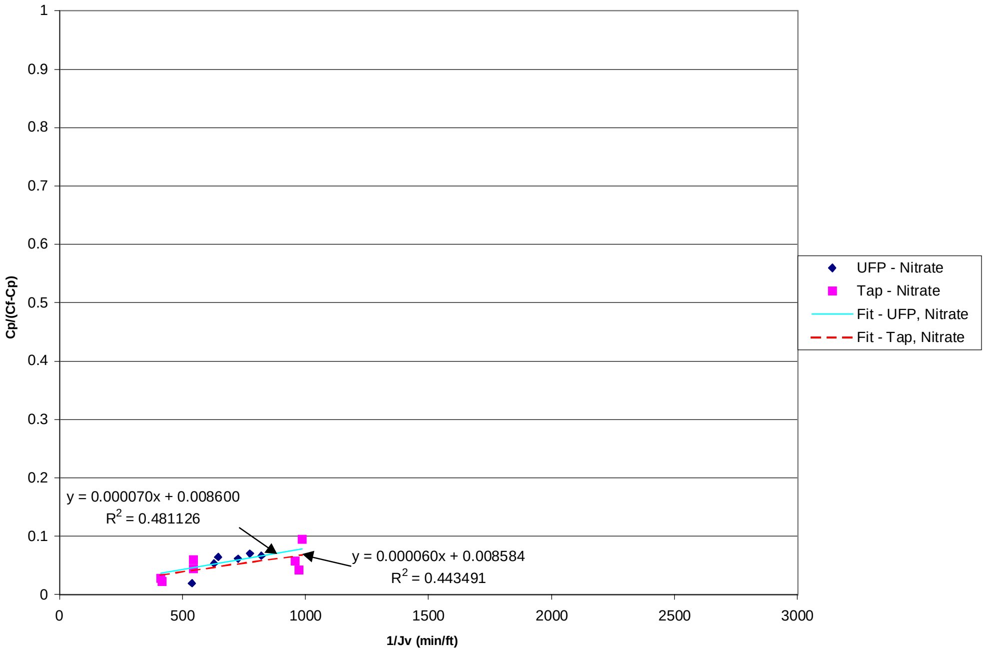

**Figure 5-37**
**CPA2 Element—“B” Parameter Estimation**

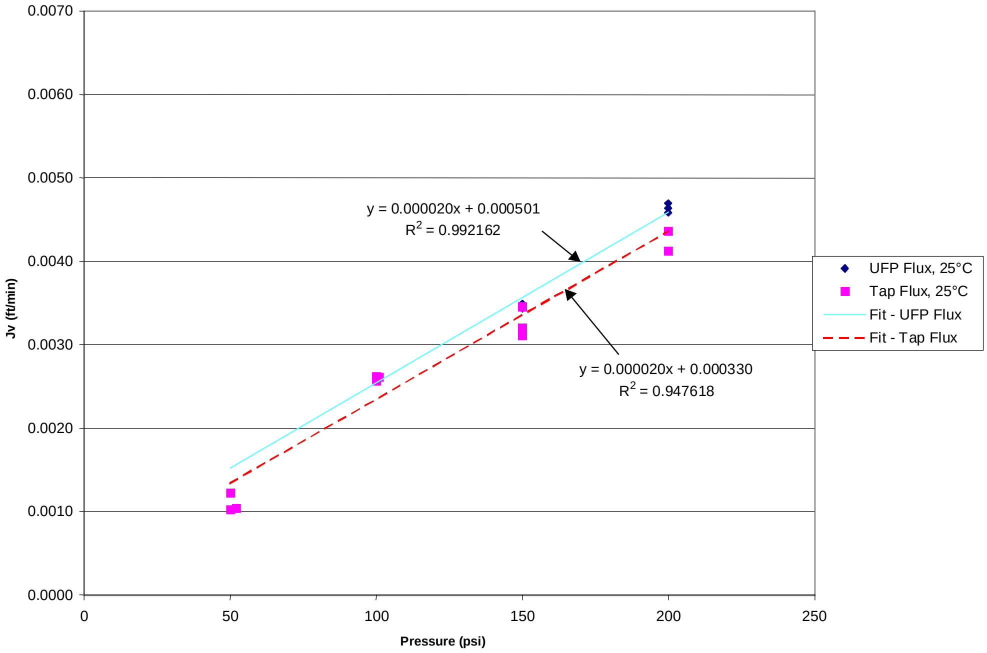

**Figure 5-38**
**ESPA Element—“A” Parameter Estimation**

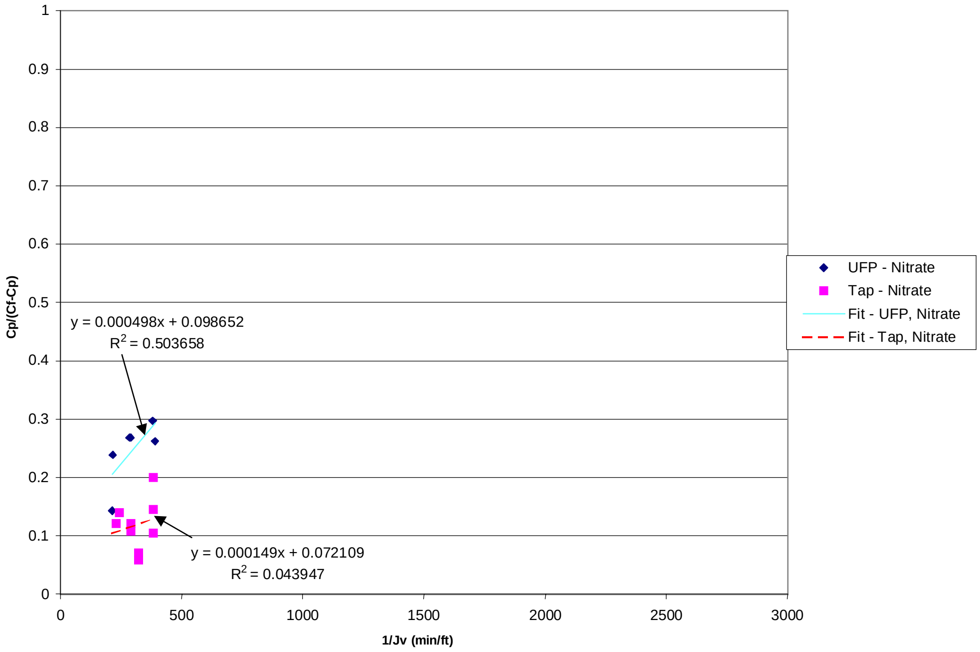

**Figure 5-39**
**ESPA Element—"B" Parameter Estimation**

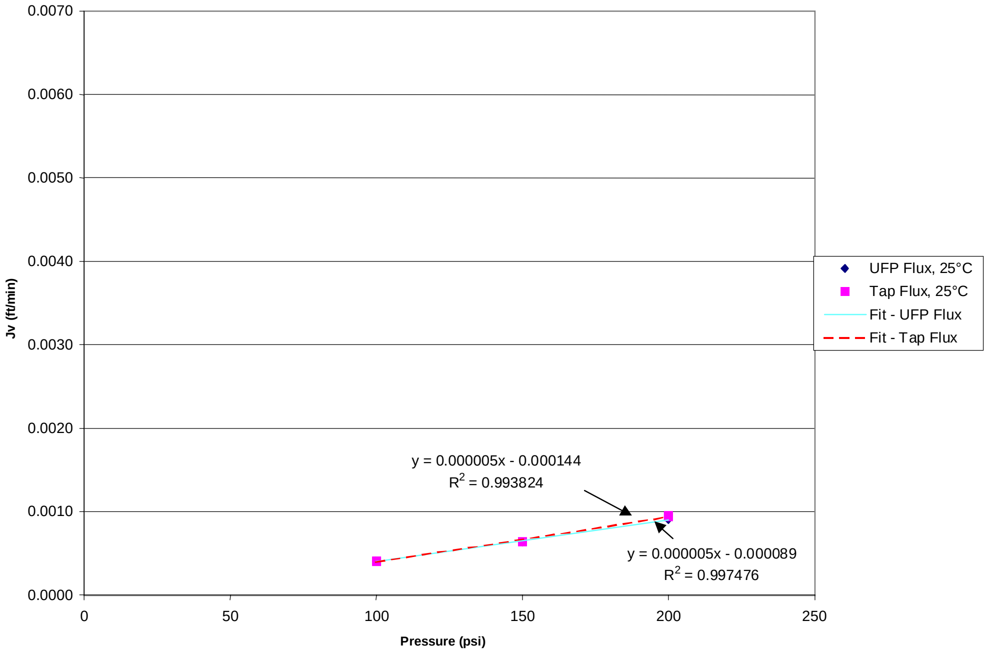

**Figure 5-40**
**HR-CA Element—"A" Parameter Estimation**

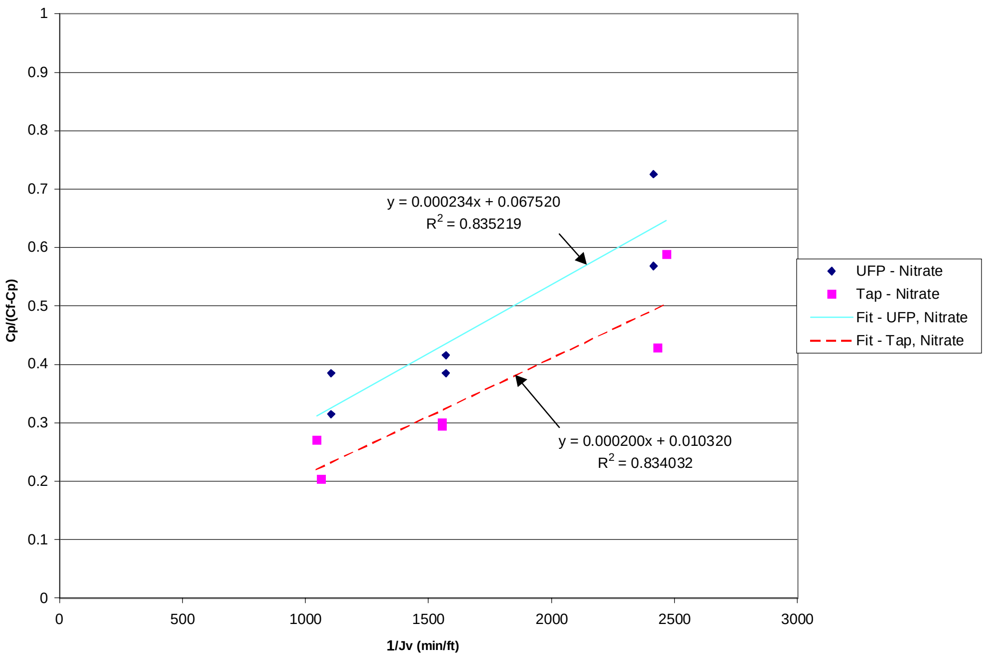

**Figure 5-41**
**HR-CA Element—"B" Parameter Estimation**

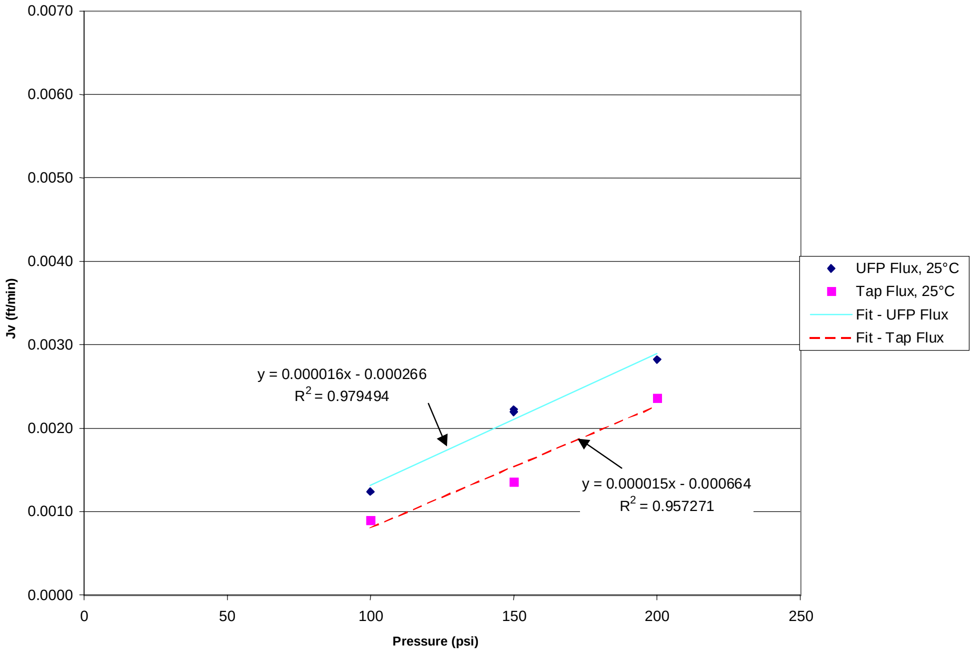

**Figure 5-42**
**HR-PA Element—"A" Parameter Estimation**

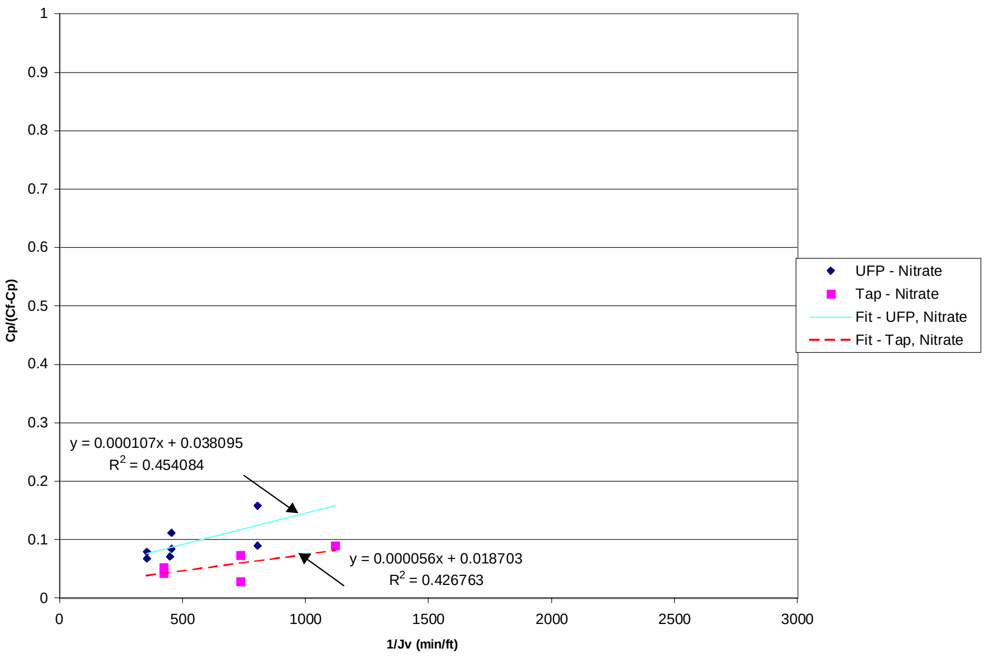

**Figure 5-43**
**HR-PA Element—"B" Parameter Estimation**

### *Long-Term Element Testing*

To evaluate the RO membranes during longer term operation, two elements (TFC-HR and CPA2) were selected to run for one month. These two elements were selected because of their ability to remove more than 90% nitrate at low pressures (200 psi (1400 kPa)). The RO pilot was setup as a single stage/element to filter tap water at recoveries from 38% to 50%. Tap water was filtered during both tests because of the slightly improved nitrate rejections in comparison with tests filtering UF permeate. Table 5-5 summarizes the results of the long-term test.

**Table 5-5**
**Summary of Long-Term Tests**

| Membrane | Pressure | Recovery (25°C) | RO Permeate [Nitrate] as N (% Nitrate Rejection) | RO Permeate [Atrazine] | RO Permeate [TOC] |
|---|---|---|---|---|---|
| TFC- HR | 201.5±3.4 psi | 40%-45% | 0.0-0.6 mg/L as N (97.5±1.9%) | <0.1 ppb | 0.01-0.40 |
| CPA2 | 199±2.8 psi | 32%-38% | 0.2-0.7 mg/L as N (89.8±2.7%) | <0.1 ppb | 0.02-0.38 |

## TFC-HR

The first experiment with the TFC-HR was started in late May and was completed in late June. During this period, concentrations of atrazine were at higher levels (3.7 ppb) than previous short term tests. Nitrate levels remained stable between 6.1 to 8.4 mg/L $NO_3$ as N.

With the presence of polyphosphates (SHMP) in the SWTP tap water, antiscalant pretreatment was not required during the long-term tests. However, the free chlorine in the SWTP tap water required dechlorination pretreatment with sodium metabisulfite. The sodium metabisulfite was added until the 23rd day of the test when the bisulfite feed tank ran out.

Throughout the test, pressure was held at 201.5 ± 3.4 psi (1389.3 ± 23.4 kPa). The ambient water temperature ranged from 24°C to 28°C (75°F to 82°F). The measured element feed water recovery ranged from 33.5% to 45% (Figure 5-44). When corrected to 25°C (77°F), the recovery ranged from 40% to 45% (Figure 5-45). Recycling of the RO retentate to increase system recovery was not performed.

During the test run, the feed water recovery at 25°C (77°F) appeared to increase (Figure 5-45). This may be in part to possible membrane surface deterioration over the test run. After the dechlorination pretreatment was stopped, a decrease in nitrate rejection was measured. The free chlorine had possibly attacked the membrane and caused it to deteriorate. However, nitrate concentrations did not follow the same increasing trend in element recovery prior to the loss of dechlorination pretreatment. Permeate nitrate concentrations would have gradually increased over the length of the experiment if membrane deterioration occurred. If membrane fouling had occurred, it was not noticeable during the test.

As can be seen in Figures 5-46 through 5-48, the removal of the primary contaminants ($NO_3$ and Atrazine) and secondary contaminant (TOC) were high (>90%). Nitrate concentrations in the feed water (Figure 5-46) ranged from 6.1 to 8.4 mg/L $NO_3$ as N. Nitrate levels were reduced below 0.4 mg/L $NO_3$ as N with nitrate rejection averaging 97.5 ± 1.9%. Considering that the TFC-HR element had been rated at 99.4% overall salt rejection by its manufacturers, this value appears to be within reason. The overall salt rejection indicates the removal of total dissolved solids from the water. While nitrate rejections were below the 99.4% salt rejection value, average rejections of hardness were higher (>99.5%) and could be used to show that this element did reject overall solids above 99.4%.

Removal of atrazine was also consistently high. Atrazine concentration in the RO permeate was reduced below the immunoassay detection limit of 0.1 ppb. The TFC-HR element can effectively remove all atrazine (<0.1 ppb) in feed levels below 3.7 ppb. Based on the TFC-HR (Figure 5-47) and HR-PA element (Figures 5-26 and 5-27) atrazine removals, higher concentrations of atrazine (5-7 ppb) would not increase the permeate concentration above 0.3 ppb of atrazine.

During the tests, the removal of TOC was monitored. The TFC-HR element was able to reduce TOC concentrations of 1.4 - 2.4 mg/L below 0.22 mg/L. Measured rejections of TOC ranged from 88% to 97%. The high TOC removal can be also attributed to the low MWCO of the TFC-HR element.

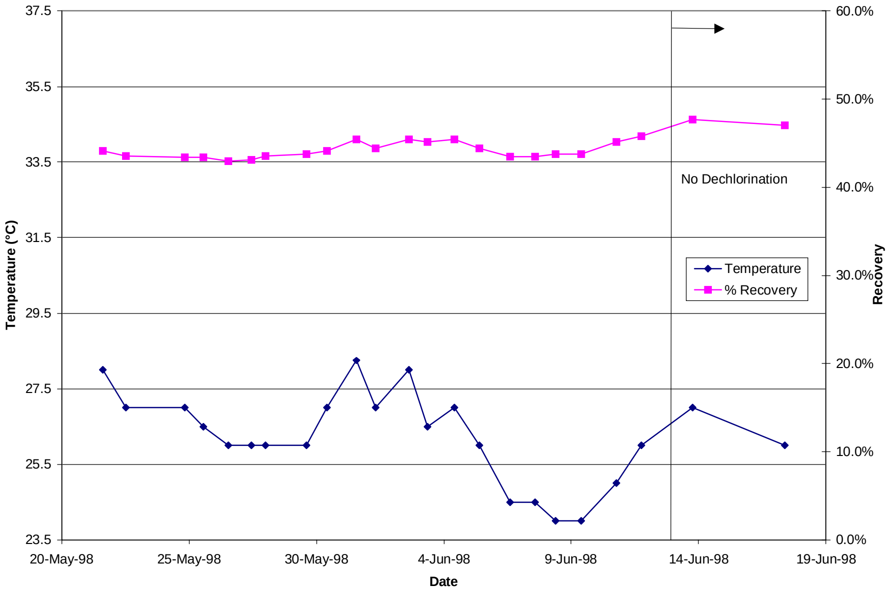

**Figure 5-44**
**Measured Recovery and Ambient Water Temperature for TFC-HR Long-Term Test**

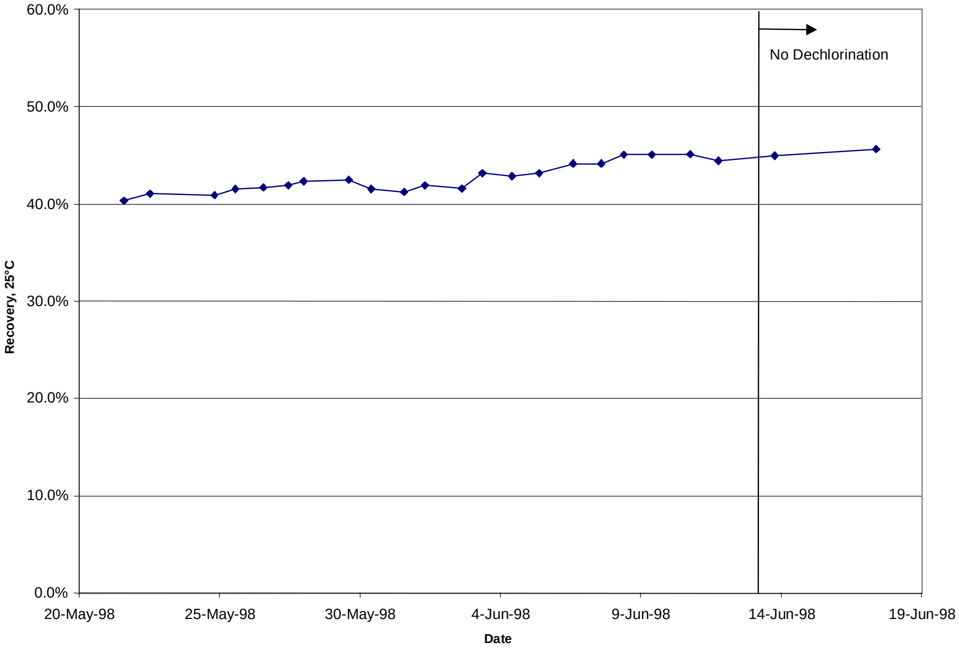

**Figure 5-45**
**Temperature Corrected Recovery at 25°C**

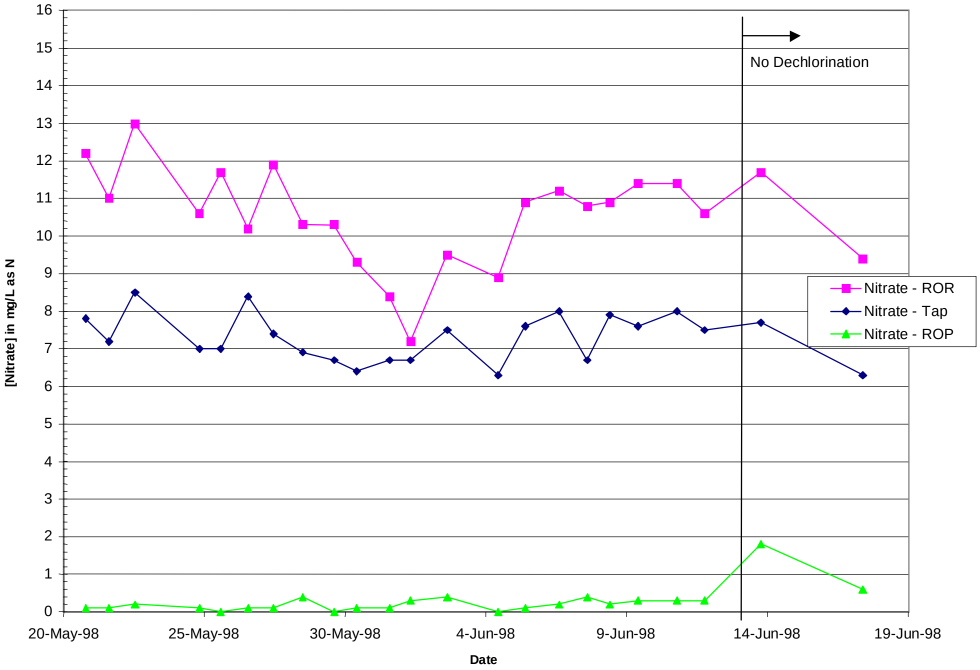

**Figure 5-46**
**Measured Nitrate Concentrations from TFC-HR Long-Term Test**

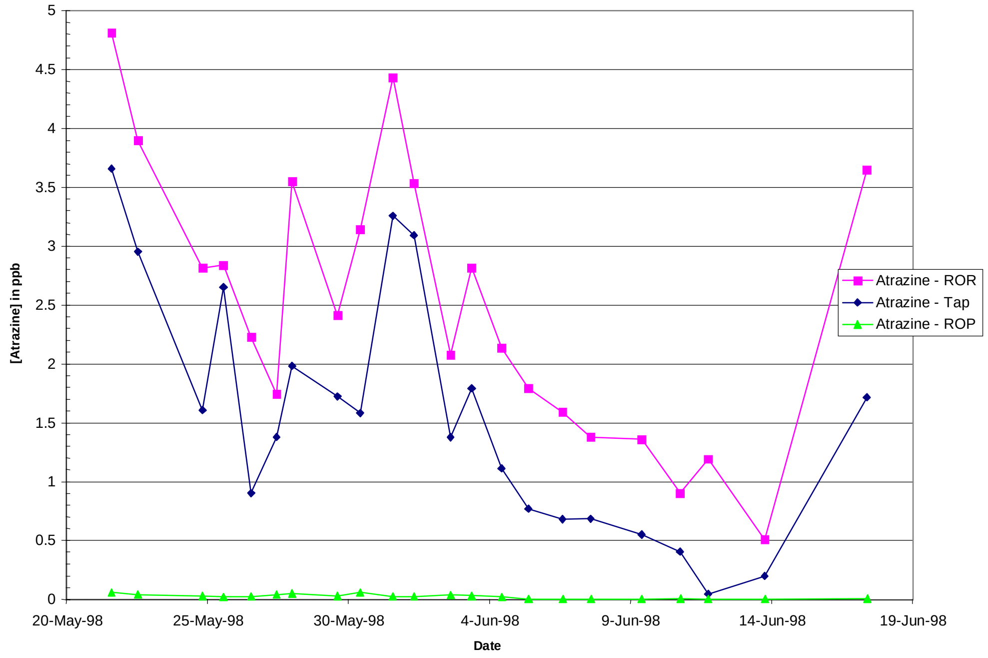

**Figure 5-47**
**Measured Atrazine Concentrations from TFC-HR Long-Term Test**

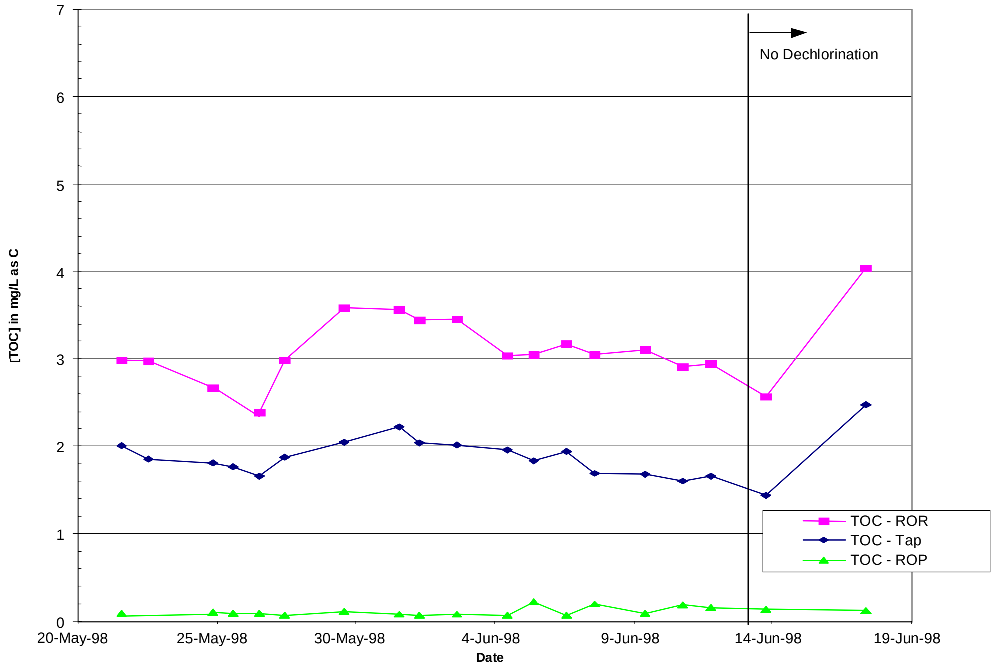

**Figure 5-48**
**Measured TOC Concentrations from TFC-HR Long-Term Test**

## CPA2

The second long term experiment used the CPA2 element because of its ability to remove nitrates. During the test, pretreatment with sulfuric acid was added to drop the pH near 7.0. The lower pH was used to prevent scaling on the membrane with higher recoveries (>50%). With an increase in the element feed concentration, retentate recycling was hoped to show the elements capacity to remove higher nitrate levels from the feed water with an increased fouling potential. The increased fouling was expected to be similar to that experienced by a second or third stage element in a multi-stage array system. Dechlorination was continued through the entire test to prevent membrane degradation.

During this experiment, the pressure was held at 199.0 ± 2.8 psi (1327 ± 19.3 kPa) with temperatures ranging from 29°C to 35°C (84°F to 91°F). Measured element feed water recovery (Figure 5-51) ranged from 41% to 45. Temperature corrected element feed water recovery at 25°C (77°F) (Figure 5-52) was lower at 32% to 38%.

By recycling the RO retentate to the feed tank (Figure 5-49), increased solid concentrations were achieved. Higher solids concentrations (800%-900%) in the RO retentate are normally experienced with multi-stage systems with an overall system recovery of 80% to 90%. When using a single element to simulate multi-stage performance, the membrane element can be exposed to higher DS concentrations by recycling the RO retentate. While this system design would not be used for full scale operation, the increase in solids concentration was hoped to increase fouling potential on the membrane.

The percent solids concentration increase could be calculated using Equation 5.3. This allowed for the comparison of tap water solids concentration, $Q_f$ and $C_f$, with that of the RO feed water into the element, $Q_{RO}$ and $C_{RO}$. Because hardness rejections were above 99% ($X_H = 0.99$), the hardness concentrations in the retentate and permeate flow were used in a mass balance to estimate the solids concentration, $C_{RO}$ (5.1).

$$C_{RO} = \frac{Q_r \cdot C_r + Q_p \cdot C_p}{Q_{RO}} \tag{5.2}$$

$$\%\, Solids Concentration\ Increase = \frac{C_{RO} - C_{feed}}{C_{feed}} \tag{5.3}$$

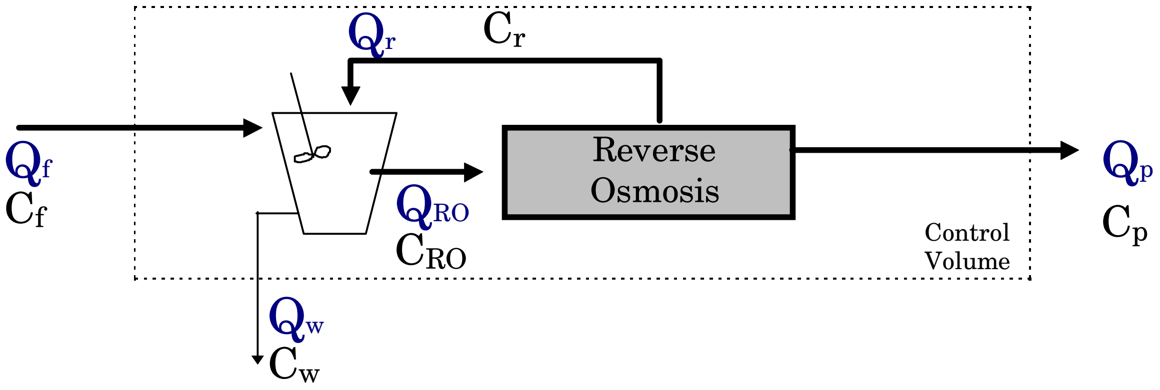

**Figure 5-49**
**RO Control Volume Approach for System Mass Balance**

Overall, the element RO feed water recovery at 25°C (77°F) showed a slight decrease of about 2% over the 28 day test (Figure 5-51) . This result was expected because of the higher solids levels present in the feed water. The higher solids concentrations are often associated with increased fouling potential as mentioned in Chapter 3. Multi-stage systems used to increase feed water recoveries would be exposed to even higher DS concentrations (<900% solids concentration).

Nitrate removal (Figure 5-52) did not appear to be affected by changes in increased solids concentration. While the RO retentate nitrate levels ranged from 10.0 to 15.6 mg/L $NO_3$ as N, tap water nitrate concentrations varied between 4.2 and 7.0 mg/L $NO_3$ as N (Figure 5-52). Permeate concentrations of nitrate remained below 0.8 mg/L $NO_3$ as *N*. Nitrate rejection for the CPA2 element (98.0% overall salt rejection) averaged 89.8 ± 2.7%. When compared with the TFC-HR element, the CPA2 element showed lower nitrate removal at higher nitrate concentrations (higher recovery) in the RO influent.

The removal of atrazine (Figure 5-53) and TOC (Figure 5-54) were similar to that of the TFC-HR membrane element. Due to the high rejection characteristics (low MWCO) of the CPA2 element, atrazine was removed below detection limits for the immunoassay test (<0.1 ppb). Feed water concentrations of atrazine were lower during this test (0.4 to 0.8 ppb). Even though the TFC-HR membrane was exposed to higher atrazine concentrations, the CPA2 element performance at higher atrazine concentrations could be similar (<0.3 ppb). TOC concentrations were reduced from 1.8 mg/L to levels below 0.18 mg/L. During this test, there were no significant deviations of the TOC and atrazine concentrations in the RO permeate.

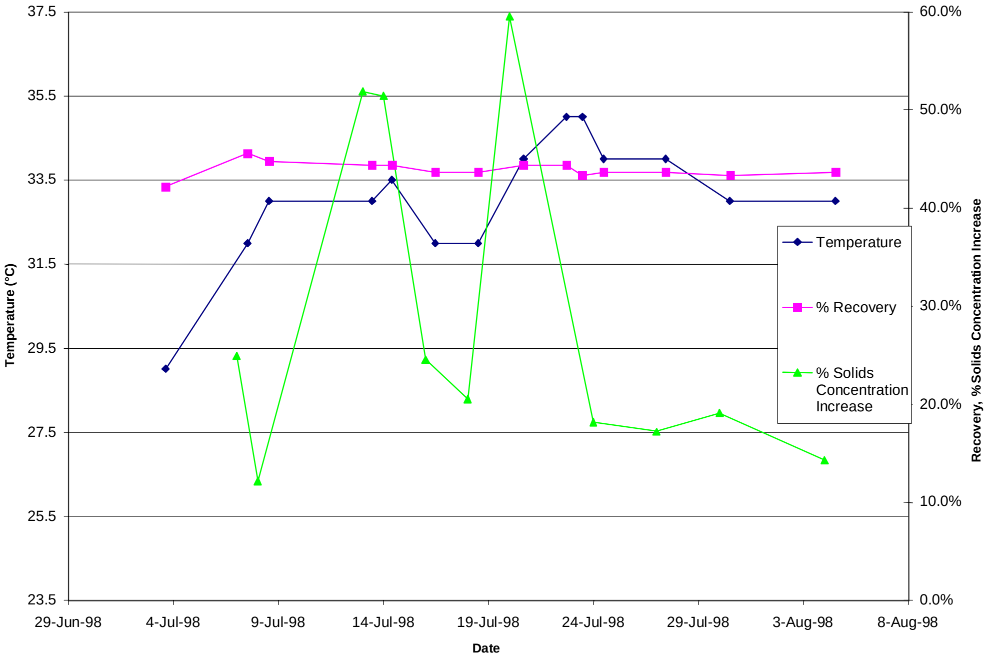

**Figure 5-50**
**Measured Recovery, % Solids Increase, and Temperature for CPA2 Long-Term Test**

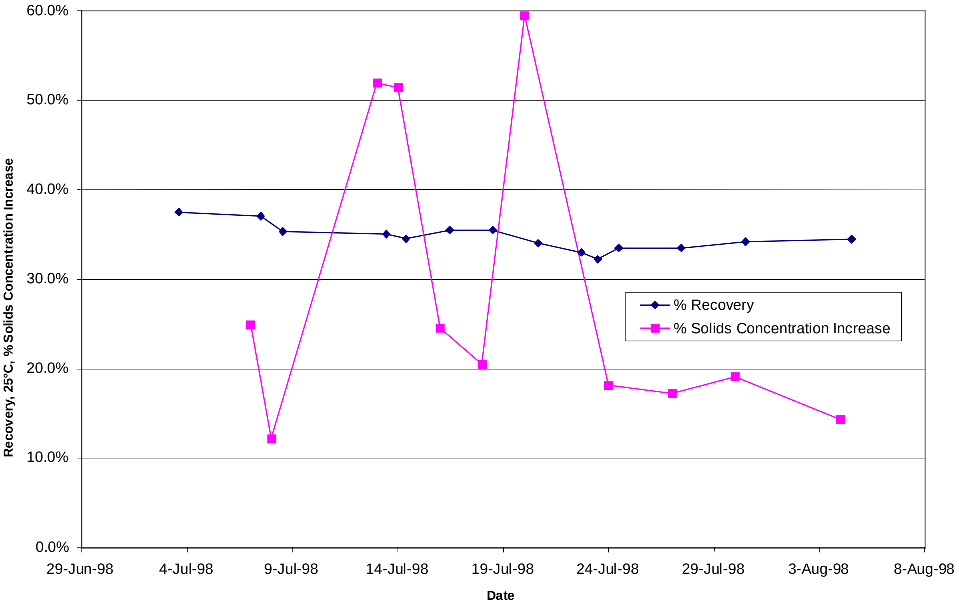

**Figure 5-51**
**Temperature-Corrected Recovery at 25°C**

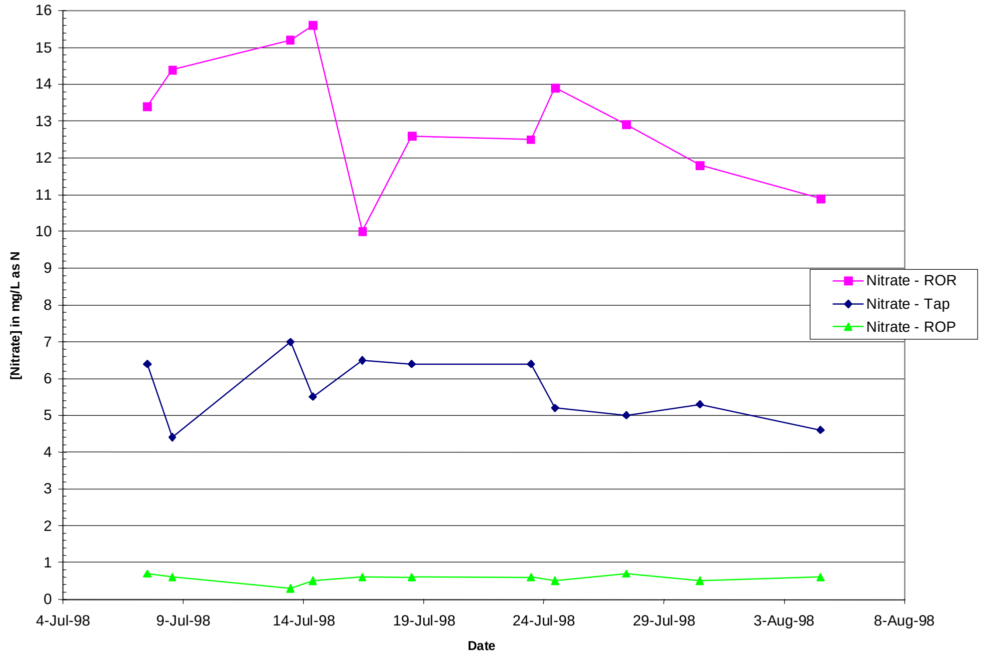

**Figure 5-52**
**Measured Nitrate Concentrations from CPA2 Long-Term Test**

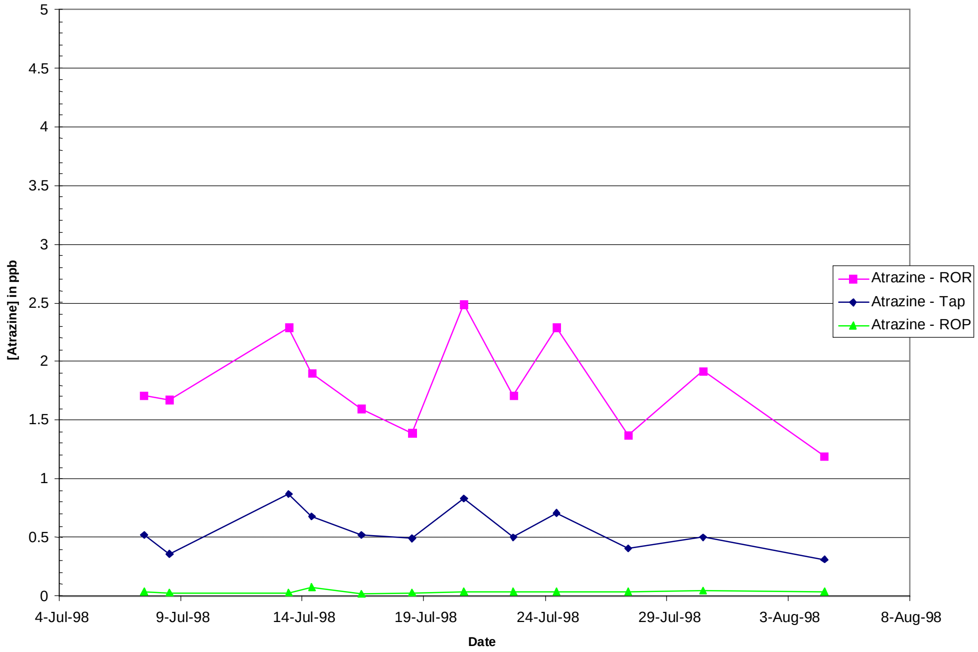

**Figure 5-53**
**Measured Atrazine Concentrations from CPA2 Long-Term Test**

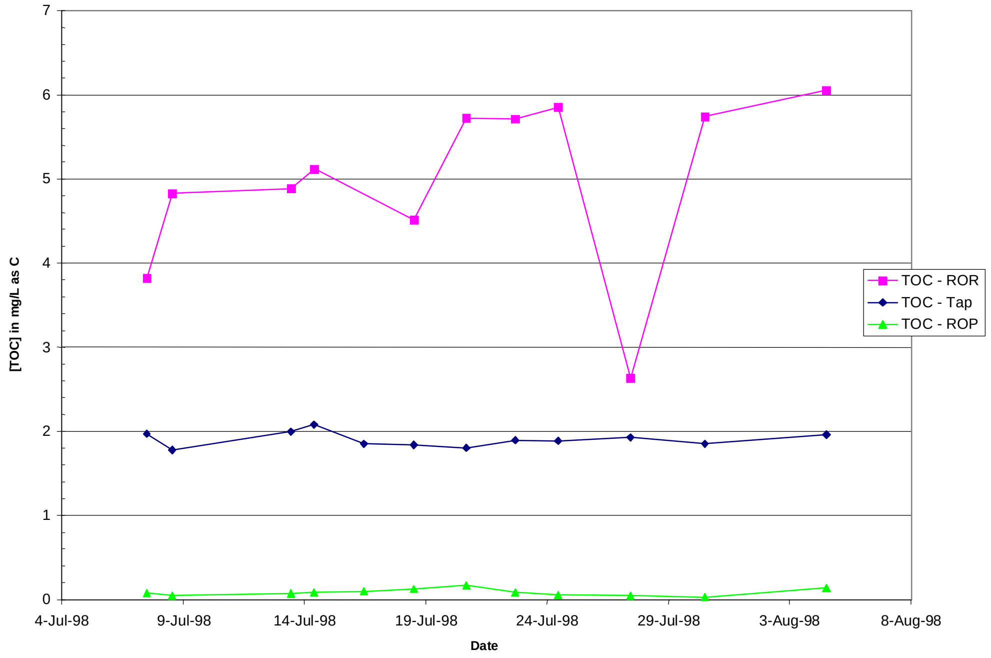

**Figure 5-54**
**Measured TOC Concentrations from CPA2 Long-Term Test**

## Remote Operation

The first year and a half of this project was spent designing, building, and automating the COMFOWARD system for remote operation. These capabilities allowed for complete monitoring and operation of the system between Decatur and Urbana, Illinois. Without these capabilities, testing would have been extremely difficult. The ability to exchange the megabytes of information directly between computers reduced data errors and allowed for regular backups of the pilot data. Operation of the UF pilot plant via the remote connection was conducted during storm events when increased numbers of backwashes were required during turbidity spikes (100-150 NTU). Monitoring of the RO pilot was conducted through a QuickCam Digital Camera which allowed digital images of the flow, pressure, and temperature gauges to be taken (Figure 5-55).

**Figure 5-55**
**Simulated QuickCam View of RO Pilot**

Feed tanks with PAC, sulfuric acid, and sodium metabisulfite were not fully automated and required refilling by a plant operator. Grab samples were taken manually at various intervals for each test. Improvements towards automated sampling or continuous nitrate measurements would have made sampling easier and more consistent. However, nitrate probes do require more frequent calibration (days to 1 week) and replacement (3-6 months based on use).

While not all components of the RO system could be controlled or monitored, the computer programming allowed for instant notification when system failures occurred during the long term test. This proved to be very beneficial and reduced down time during tests from a couple of days to a few hours. Testing was virtually continuous through the long-term tests.

Even with such a small scale plant, the COMFOWARD system shows that it is possible to semi-automate the UF and RO membrane process. With more probes and sensors, further programming could be developed that would allow a “smart” system to react to changes in water quality (ie. turbidity, hardness, temperature). In combination with VFD pumps, the system operation could be optimized according to electrical costs (based on peak vs. non-peak rates), system demand, and membrane performance. The ability for data collection by regulatory agencies could help to enforce legislation or gain knowledge from a range of systems in operation. Communications between various agencies, water treatment plants, and utilities could be reduced via a common GIS platform that provides current and accurate information. While more industrial strength controls and equipment would be required, this project shows that off-the-shelf technologies can be used towards mimicking similar operations at both pilot and full-scale.

## Cost

Current models to estimate costs for a full scale plant require design capacities, energy, and good knowledge of costs associated with membrane processes. For cost estimating, only the full-scale RO system will be considered. The current SWTP effluent is of high quality and already incorporates the necessary pretreatment (polyphosphates, turbidity removal) to implement RO.

In order to determine a design flow to meet regulatory standards, a suggested maximum nitrate level of 8.5 mg/L $NO_3$ as N could be maintained. This level represents the IEPA trigger for public notification of the distribution of nitrate free bottled drinking water. Nitrate levels less than 10 mg/L $NO_3$ as N would be in compliance with current regulations.

Over the past 72 years, the maximum recorded nitrate level in the Decatur drinking water has been 16 mg/L $NO_3$ as N (Illinois State Water Survey, 1996). While this value has only been measured twice, a design maximum of 15 mg/L $NO_3$ as N (94% of the maximum) was chosen.

It is also important to note that nitrate levels have surpassed 10 mg/L $NO_3$ as N at water temperatures near 8°C to 10°C (46°F to 50°F). Temperature will influence the amount of flux produced. Lower temperatures for a full-scale RO system would require higher pressures to meet a constant design flow throughout the year.

An RO treatment plant using the TFC-HR elements can then be used to remove nitrates at an estimated value of 97.5 ± 1.9% (minimum suggested rejection = 95.6%). The maximum recovery rate for a single TFC-HR membrane element at 25°C (77°F) and 200 psi (1379 kPa) is 45% (30% at 10°C (50°F) and 200 psi (1379 kPa)). Without recycle, a three-stage array of TFC-HR elements at 30% to 45% recovery would allow for 65% to 83% overall recovery (Figure 5-56). These minimum and maximum conditions can then be used to determine the maximum amount of water that must be treated with RO prior to the distribution system (Figure 5-57).

In order to meet the 10 mg/L $NO_3$ as N standard, maximum recovery with minimum amounts of RO feed water will aid in reducing costs. For the purpose of summarizing possible costs, the design flow of 18 MGD (68,000 $m^3$/d) will be used with element nitrate rejections at 95.67%. This feed flow will allow for concentrations to remain below the 10 mg/L $NO_3$ as N for a three-stage array with element recoveries ranging between 30% (9.40 mg/L $NO_3$ as N) to 45% (8.65 mg/L $NO_3$ as N).

The computer models (ROSA, Winflows, and Hydranautics) helped to determine the number of elements required to produce a given permeate flow. For a two-stage system, over 300 pressure vessels or 2100 elements would be required to produce 10 MGD (38,000 $m^3$/d) at pressures from 150 to 200 psi (1034-1379 kPa). The ROSA program was limited to a 10 MGD flow rate. The Winflows program suggested similar values.

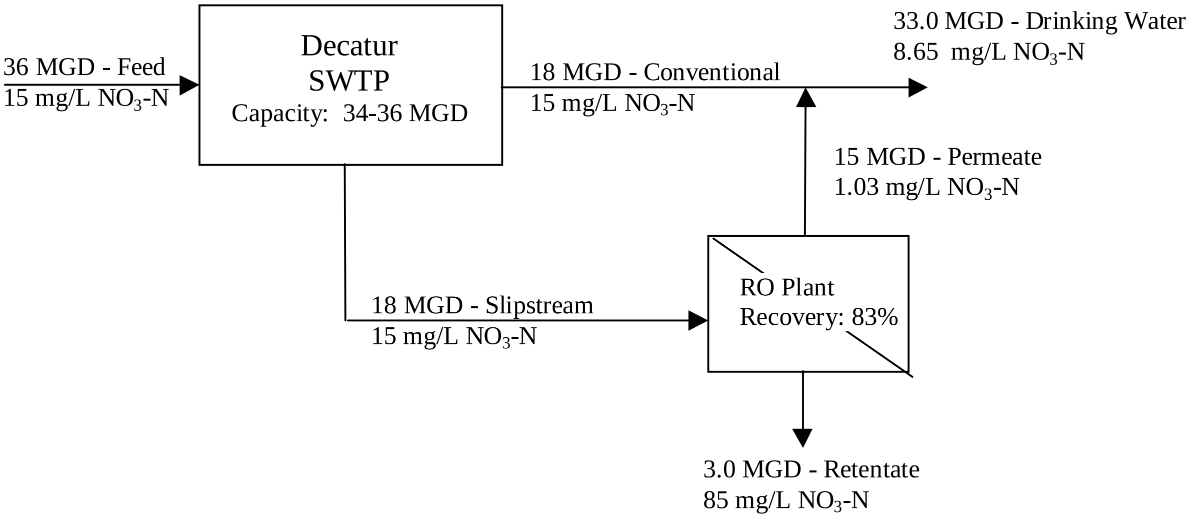

**Figure 5-56**
**Slipstream Model to Determine Design Flow**

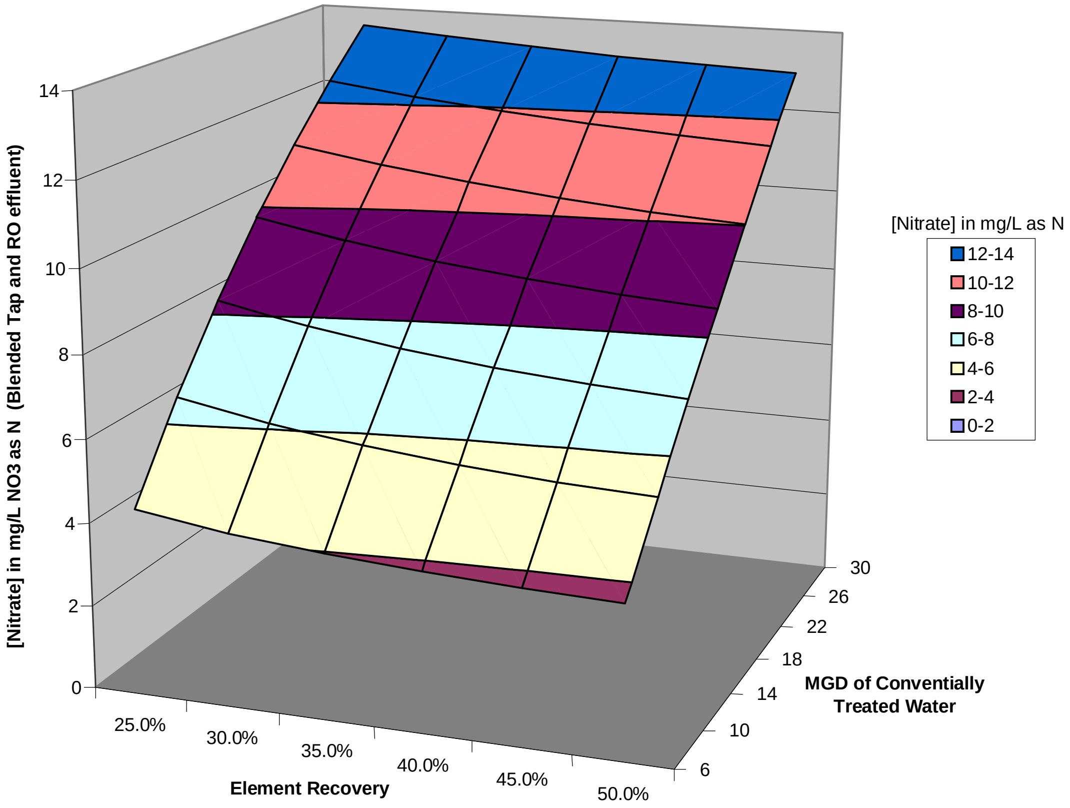

**Figure 5-57**
**Nitrate Concentrations in Blended Distribution Water for 3-Stage Array with Varying Element Feed Water Recovery (Nitrate Rejection = 95.67%)**

The Hydranautics RO Design program allowed the higher permeate flow rates to be calculated (18 MGD (68,000 $m^3$/d)). Approximately 2000 high rejection (99.5% overall salt rejection) CPA2-8540 elements and 300 pressure vessels would provide the required amount of permeate to reduce nitrate. Given the costs of the elements and pressure vessels and a more optimized solution, the membrane capital costs could be calculated.

The non-membrane capital cost (Equation 3.35) showed that the costs for construction, equipment, and instrumentation would be near $1,866,000 or $0.01/1000 gal ($0.003/$m^3$). The capital cost estimate assumes a 7.0% nominal interest rate over a design life of 20 years.

Energy costs for the 18 MGD (68,000 $m^3$/d) plant can be estimated using electric use rates for summer ($0.07619) and winter ($0.04619) seasons. At 200 psi (1400 kPa), the cost of energy would be $0.25/1000 gallons ($0.066/$m^3$) during the summer months and $0.17/1000 gallons ($0.045/$m^3$) during the winter months. That represents an average rate of $0.21/1000 gallons ($0.055/$m^3$) or a yearly energy cost of $1,379,700.

Operation and maintenance costs for the 12 MGD (45,000 $m^3$/d) plants (presented in Table 3-4) could be used to estimate overall costs. While these costs may be lower due to the economies of scale, a rate of $0.36/1000 gallons ($0.095/$m^3$) for a 12 MGD (45,000 $m^3$/d) plant could be used as a maximum estimate of O&M costs for an 18 MGD (68,000 $m^3$/d) RO plant. At the rate of $0.36/1000 gallons ($0.095/$m^3$), $2,365,200 per year would be spent to maintain the RO system and operations.

Total costs before membrane capital costs are estimated to be near $0.58/1000 gallons ($0.15/$m^3$) produced ($3,810,600 per year at 18 MGD (68,000 $m^3$/d) production). Additional costs in the form of additional chemical pretreatment, retentate disposal, and plant design/startup may also affect this total cost value. While more detailed cost estimates would be developed for an actual design, these costs represent an overall estimate for a base comparison of other nitrate removal technologies.

## Preliminary Design and Related Costs

Since completion of the thesis entitled *Membrane Technologies for the Removal of Nitrate and Atrazine from a Surface Water Source*, a more detailed preliminary plant design has been prepared for the City of Decatur as promised through the project. Original cost estimates presented in the MS thesis were based on existing literature describing RO systems in operation. Most systems considered were entirely new treatment plants. Due to the design of the South Water Treatment Plant, modifications can be made to the plant that would allow for full use of an RO system and the existing plant. The proposed transfer of the North WTP to ADM will make it possible to use a smaller RO treatment facility compared to published projections. Reverse osmosis could also be used to solve other potential problems in the future as well.

### *Preliminary Objectives and Previous Work*

In order for the City of Decatur to implement RO treatment, several design constraints were taken into account. These were:

1. Maintain existing water treatment capacity and quality.
2. Use a high rejection type RO membrane for the best nitrate removal as proven with previous pilot work.
3. Provide a flexible system according to demand and contamination requirements.
4. Provide similar construction to existing facilities.
5. Reduce both capital and operating costs to be competitive with other treatment and management issues.

6. Minimize the amount of waste produced.

7. Provide a low maintenance and operator friendly system.

8. Provide a long-term solution to Decatur's water quality issues.

Original estimates presented in previous work (Hackman, 1998) showed that an 18 MGD (68,130 $m^3$/d) RO system would be required. All cost estimates for that system were based on literature reviews of existing treatment plants in operation. The literature cost estimates were based on RO treatment plants built "as new" to replace existing facilities. Additional costs for land acquisition, site preparation, new construction, and additional pretreatment may have increased the cost estimates compared to this proposed RO system with existing pretreatment (Decatur's South WTP). Based on information provided by the Water Production Staff, water demand for the City of Decatur is estimated to decrease from 34 MGD to 24-25 MGD (128,690 to 90,840-94,625 $m^3$/d) at the South WTP as the sale of the North WTP to ADM, Inc is completed. This reduction in water production is significant enough to affect the amount and cost of RO treatment required to remove nitrates.

### *Preliminary Design Work*

With the objectives in place, preliminary design work continued. Drawings provided by the City of Decatur Water Production Staff aided in making preliminary design layouts. Additional water quality data was used with computer models and the help of membrane component manufacturers to predict pretreatment requirements, reverse osmosis treatment capacity, and final water quality. Equipment manufacturers were able to provide equipment dimensions to be included in a floor layout design. The scope of the preliminary design included:

1. Estimation of the reverse osmosis system size, cost, and operation.

2. Estimation for the necessary pumps, chemical feed equipment, and instrumentation.

3. Estimation for a new electrical substation, motor controls, and SCADA operation.

4. Modification to the existing operations building to extend treatment.

5. Development of a floor layout showing the new system in place.

6. Construction timeline of new facilities to incorporate a working RO system.

7. An estimation of operation costs associated with the new facilities.

To allow for future modifications, several design options were considered. The designs included options for different membrane element and component manufacturers, pumps, chemical manufacturers, and electrical controls. The preliminary work presented here will focus on a base design option that should be optimized for construction time, product support and maintenance, and cost prior to final design.

## Sizing the Reverse Osmosis System

With the change in water demand at the South WTP, a smaller RO system could be designed for reducing nitrates below the MCL level of 10 mg/L NO3- as N. The maximum observed contaminant level of 15 mg/L NO3- as N was used for design purposes. At this contaminant level, up to 10-11 MGD (38,000-42,000 $m^3$/d) of RO permeate would be required to reduce levels at or below 9.0 mg/L NO3- as N at a total plant flow rate of 24 MGD (90,000 $m^3$/d). The SCADA and valve systems should allow for automatic adjustment of the RO system to meet the 9.0 mg/L NO3- as N contaminant level or better.

Given the existing water quality data, computer design software could be used to determine the number of membrane elements, pressure vessels, and/or skids required. The previous 10 MGD (38,000 $m^3$/d) ROSA model estimates were consistent with this new RO design, 300 pressure vessels with 2100 elements. Osmonics currently produces a 1 MGD (3,800 $m^3$/d) RO membrane skid design which could also be used. Ten to eleven RO skids would be required in this case for 10-11 MGD (38,000-42,000 $m^3$/d) production. These membranes would be operated in the range of 225 to 250 psi (1600-1700 kPa) for an optimal flux of 15 gfd as per manufacturer's recommendation. With the permeate side of the membranes connected to the 7.0 mg reservoir, a 20 psi (138 kPa) backpressure on the permeate side of the membrane was included to assure that flow to the reservoir would be possible. Recovery rates between 75% to 85% could be achieved.

With this design in place, budgetary cost estimates from the various membrane manufacturers were obtained. The following table is a summary of those membrane capital costs.

**Table 5-6**
**Summary of Membrane Capital Costs**

| Membrane System Component | Description | Qty | Cost |
|---|---|---|---|
| **Skid Mounted Pre-fab System** | | | |
| 1 MGD skid (Osmonics, Inc.) | Includes HR-PA membrane elements, stainless steel pressure vessels, intervessel piping, connection piping, 6 in. SS316 Feed, 6 in. PVC Permeate, 3 in. SS316 Retentate, preassembled, on-site support | 11 | $3,637,000 ($3,627,000 + $10,000 shipping charges) |
| **Component System** | | | |
| Pressure vessels | 8 in. diameter Code Line Systems FRP vessels or Osmonics Stainless Steel | 300 | $850,500 max. |
| Membrane Elements | 99.4% overall salt rejection, polyamide, 8 in. x 40 in. membrane elements | 2100 | $1,984,500 max. |
| Vessel Supports, Membrane Installation, Shipping, and Piping | Stainless Steel and PVC piping connections, steel rack for pressure vessel mounting, installation | | $1,012,000 |
| Probable Total Component System Cost | | | $3,847,500 |

The component system allows for more flexibility for system design and construction. However, membranes would have to be installed by the contractor and their installation would not be guaranteed by the membrane manufacturer. This could present an unacceptable risk unless a qualified contractor is on-site.

The best option from a cost and quality standpoint would be a prefabricated skid unit system. Osmonics, Inc., for example, would be responsible for construction and membrane installation for each skid. Once on site, the skids could then be connected to pipe manifolds with the necessary valving and be ready for operation.

## Pretreatment Chemicals

Providing the quality and quantity of water necessary to run the RO is another consideration. While the previous report states that no additional pretreatment was required other than sodium metabisulfite, standby pretreatment processes should be available for long-term operation to maximize membrane life. Previous long term tests using the TFC-HR and CPA2 membranes with sodium metabisulfite pretreatment showed no or minimal flux decline over the month long tests using the Decatur South WTP effluent. Full scale systems often incorporate other pretreatment chemicals with raw or untreated waters or those waters with the potential to scale (positive Langlier Saturation Index).

The pretreatment chemicals suggested for this preliminary design include sodium bisulfite for dechlorination, sulfuric acid addition for pH adjustment, antiscalants for lower fouling potential, and an NSF-approved disinfection agent to prevent inadvertent biofouling of the membranes. The additional feed lines would be available for operation similar to the ammonia and chlorine lines used to supply raw water to the COMFOWARD pilot system.

The chemical additions would occur within the pipe from the existing clearwell to the newer clearwell extension. Additions within the connecting pipe will help to promote chemical mixing prior to the RO system and prevent membrane damage. This will also be discussed in further detail in the section titled Chemical Feed System and Instrumentation.

Sodium bisulfite addition is important to dechlorinate the water. Previous work present in the thesis explains how sodium bisulfite is used to remove free chlorine ($OCl^-$ or $Cl_2(g)$). Free chlorine in the feed water would cause damage to the membranes that is irreversible and could reduce nitrate rejection rates. Dosage rates were determined from the AWWA standard B601 for taste and odor control using sodium metabisulfite. Quantities of dry chemical would be mixed with treatment water at the plant. Food-grade-quality chemical sodium metabisulfite suppliers and distributors are located in the central Illinois area (Ashland Chemical).

Sulfuric acid addition is used to adjust the pH of the water to prevent scaling on the membranes. Scaling reduces the amount of water produced through a membrane over time. The pH adjustment of the water will also precipitate aluminum from the source water. Any aluminum that passes through the multimedia filters will be removed prior to the RO system via the clearwell through additional settling. At this time, sludge production is anticipated to be minimal, if any, and could be removed from the clearwell during washout.

Antiscalants may be added to prevent calcium precipitates from scaling the membranes. Higher recovery rates can be achieved with the use of antiscalants to minimize waste formation. The antiscalants are selected in combination with the membranes used and chemical manufacturers provide software to optimize the antiscalant dosage rates. Some antiscalants use dispersants to help prevent cake formation from suspended solids. This may help reduce particulate fouling on the membranes when multimedia filter effluent turbidities are between 0.1 and 0.2 NTU. The product presented in this opinion of probable cost is supplied by Professional Water Services, Inc. Their cost opinion was used in this report because it was the highest priced of the antiscalants available to determine the maximum operation cost associated with this preliminary design. Calgon product number EL-5600 should be able to provide the same antiscalant protection at a lower operational cost. Calgon EL-5600 is approved by the EPA for potable water production using reverse osmosis membranes.

Antiscalant use can be minimized further with the presence of SHMP (sodium hexametaphosphate) in the treatment system. SHMP is currently used to prevent calcium scaling on the multimedia filters. During pilot tests, flux decline was not detectable during the long-term tests (1 month in length with two membranes). The antiscalant cost estimate was developed assuming no SHMP addition and using one of the more expensive antiscalant chemicals available for this opinion of cost.

Disinfectants can be used to prevent bacterial growth. With the amount of DOC, sulfates, nitrates, and polyphosphates in the water, bacterial growth is possible. Other nitrate removal technologies could experience similar bacterial contamination problems due to the fact that biological nutrients (carbon, phosphorous, and nitrogen as nitrate, and sulfur) are present in the water, and bacteria could grow without the presence of chlorine disinfectants that are removed prior to treatment. Protection from biofouling of the membranes can be done with non-chlorine disinfectants during operation that do not affect the final permeate water quality or membrane performance. These disinfectants are NSF-approved for potable water use and can be provided by Professional Water Services. The product proposed is Biogard ACS (Professional Water Services, Inc.), an anti-microbial growth compound.

Of the pretreatment chemicals, sodium metabisulfite and disinfectant chemicals should be the only pretreatment required for normal operation. Antiscalants and sulfuric acid addition may be required on an occasional basis to prevent or reverse scaling as water quality changes throughout the year (i.e., higher hardness and turbidity levels)

Dosage rates and costs per day for feeding 14 MGD (53,000 $m^3$/d) to this preliminary RO system are as follows in Table 5-7.

**Table 5-7**
**Dosage Rates and Costs Per Day for Feeding 14 MGD (53,000m$^3$/d) to this Preliminary RO System**

| Chemical | Maximum Feed Rate | Unit Cost | Maximum Day Cost |
|---|---|---|---|
| Sodium Bisulfite | 1:1.4 $Cl_2$ to $NaS_2O_5$ = 24.1 gal/day (91.2 l/d) | $0.37 per pound dry weight ($0.81/kg) | $61 per day |
| Sulfuric Acid, 25% | pH from 9.7 to 8.5 = 6 gal/day (22.7 l/d) | $2.10 per gallon ($0.55/l) | $13 |
| Antiscalant estimate 1 (optional dosage rate provided by Professional Water Services, Inc.) | 2.8 ppm @ 25°C- 3.5 gal/day max. (13.2 l/d)<br>10 ppm @ 5°C - 13.75 gal /day max.(52.0 l/d) | $150 per gallon ($39.63/l) | $525 per day at 25°C<br>$2,100 per day at 5°C |
| Disinfectant | 3.5 ppm = 3.6 gal/day (13.6 l/d) | $180 per gallon ($47.56/l) | $648 per day |
| | | Maximum Total Daily Chemical Cost | $1,250 per day to $2,850 per day |

## High Pressure Pump System

In order to supply the pressures and flow required for treatment, pumps are required to operate the system. Using a Fairbanks Morse vertical turbine design, pressures from 240 to 280 psi (1655 to 1931 kPa) can be achieved for a flow of 4 MGD (15,000 m$^3$/d) per pump (6 or 7 stages). A four-pump design would allow for a total of 16 MGD (61,000 m$^3$/d) of flow. A firm capacity of 12 MGD (45,000 m$^3$/d) could be maintained should one pump go off-line for any reason. These pumps would pump from the clearwell extension into a pump header for distribution to the membrane skids. Based on the amount of nitrate present and the WTP production rate, the amount of permeate production can be controlled by changing the number of pumps and membrane skids in operation. Booster pump control and backpressure sustaining valves would be incorporated to control system pressures and prevent system surges. Each pump would require a 600 HP motor to be driven by 4160 VAC 3 phase 60 Hz electricity.

The clean-in-place process can be incorporated with the proposed equipment. The pump system with chemical injectors should allow for periodic cleaning of the membranes. Chemical pumps will feed from a CIP chemical feed tank (5,000 - 10,000 gallons (19,000-38,000 m$^3$)) and be connected to an injector tap at the pump discharge to provide CIP capabilities to each 1MGD skid. Waste would be discharged to the existing sewer and not be allowed to contaminate plant potable water.

The proposed piping system would consist of epoxy-coated ductile iron on the interior and exterior surfaces to prevent corrosion during pH adjustment and normal operation. Stainless steel piping would be more expensive, but better suited for use with corrosive feed waters is antiscalants and SHMP are not used. If antiscalants and SHMP are used on a regular basis, the feed and retentate waters would be non-corrosive. Due to the nature of the feed water and retentate, the only corrosive water present within the system should be the clean permeate.

All permeate piping should be CPVC schedule 80 to prevent any corrosive permeate from damaging the piping. The permeate side of the membrane should see pressures in the range of 20 to 25 psi (140 to 170 kPa). Schedule 80 CPVC is rated for pressures to 250 psi (1700 kPa). The permeate main will be piped back to the 7.0 mg (26,500 $m^3$) reservoir through a connection to the 54-inch (137 cm) reservoir feed main. The 54-inch (137 cm) reservoir feed main will be used to mix the permeate and non-membrane treated plant effluents to dilute the nitrate concentrations.

During system operation, the pipe trench will include all the necessary piping below grade for a clean appearance and to allow waste draining during system startup and operation. Retentate waste would be piped directly to the sewer connection. A 24-inch (61 cm) sewer would be connected to the existing 48-inch (122 cm) sewer main with a manholed invert.

The opinion of probable cost associated with the pumps and piping are presented in Table 5-8

**Table 5-8**
**Probable Costs Associated with the Pumps and Piping**

| Piping Component | Quantity/Description | Total Cost |
|---|---|---|
| **Piping** | | |
| 6 in. DI Pipe | 150 feet (installed) | $7,500 |
| 12 in. DI Pipe | 150 feet (installed) | $15,000 |
| 24 in. DI Pipe | 100 feet (installed) | $15,000 |
| 36 in. DI Pipe | 50 feet (installed) | $12,500 |
| 24 in. PVC Pipe | 200 feet (installed) | $40,000 |
| 24 in. Sewer Extension | 40 feet and Saddle Invert (installed) | $2,000 |
| **Pumps** | | |
| 2800 GPM at 270 psi | 4 Pumps with Motors | $317,500 |
| **Valves** | | |
| Valve System | Includes major valve assemblies for each pump and RO skid | $162,000 |
| **Total Pump System** | | **$571,500** |

Motor and SCADA controls to the high pressure pumps will be discussed in the section titled Electrical Substation, Motor Controls, and SCADA Programming.

### Chemical Feed System and Instrumentation

The chemical feed system provides the necessary feedback and control to maintain the RO feed water quality. Instrumentation through various probes and flow meters can be used to control the chemical feed pumps. The instrumentation package for this system includes the necessary flow, water quality, and pressure gauges for full monitoring of the system (Table 5-9).

**Table 5-9**
**Water Quality Instrumentation**

| Water Quality Instrumentation | Description/Quantity | Cost |
|---|---|---|
| Pressure | 27 Pressure gauges and transducers, pressure switches | $16,200 |
| Temperature | 3 Probes w/Transducers | $1,200 |
| pH | 3 Probes with Meters and digital output | $3,000 |
| Conductivity | 3 probes with Meters and digital output | $3,000 |
| Flow | 2 turbine meters, flow switches | $7,600 |
| Nitrate | 3 probes with meters and digital output | $4,500 |
| Nitrate Probes | 12 replaceable probe tips per year | $480 per year |
| Turbidity | 2 Turbidometers with flow through cells | $2,000 |
| Shipping | | |
| Total Instrumentation Package | | $37,980 |

The chemical feed system, based on the chemical feed rates, are as follows in Table 5-10. Tanks for the CIP Process are included. Chemical feeds are in duplicate for maintenance purposes.

**Table 5-10**
**Chemical Feed System Based on the Chemical Rates**

| Chemical Addition | Quantity | Cost |
|---|---|---|
| **Chemical Feed Pumps** | | |
| Sodium metabisulfite | 2 | $20,300 |
| Sulfuric Acid | 2 | $20,300 |
| Antiscalant | 2 | $20,300 |
| Disinfection | 2 | $20,300 |
| Clean-In-Place Injector Pump | 2 | $27,000 |
| **Tanks** | | |
| 10,000 Gallon FRP | 5 | $40,500 |
| 200 Gallon FRP | 8 | $5,400 |
| **Total** | | **$154,100** |

### Electrical Substation, Motor Controls, and SCADA Programming

Because of the high horsepower requirement, medium voltage was selected for the preliminary design. Existing facilities use 4160 voltage to power large pumps within the treatment plant. In order to provide the new facilities with power, additional transformers would be required. This would include underground transmission lines (345 kV) to a transformer, construction of a transformer pad, connections to a 4160 V secondary and motor control system for 1750 kW maximum power demand, and connections for a 75 kVA or larger secondary transformer for 480, 240, and 120 VAC. Additional lighting and support equipment would be pulled from the secondary transformer.

The motor controls would allow for soft or reduced voltage start of each pump motor brought on-line. The slow VFD or reduced voltage start in combination with slow opening/slow closing valves will reduce the risk of inadvertent membrane damage from pressure shocking. A variable frequency drive would be used for system startup and for trimming the system flow rate to match demand. The switch gear design would incorporate the ability to ramp through each pump as water demand changes occurred. More pumps would be brought on-line to increase permeate production as needed.

SCADA control would be used to collect data and control system pumps, valves, and feed equipment. To assure operation safety, manual controls would be in place should automatic operation fail. The SCADA system would be similar to controls in place utilizing the Interloution FIX 32 software/hardware interface. All controls and transducers should allow for 4-20 mA or equivalent signals to the computer system. Electrical components could be similar to previous equipment (i.e., manufacturer) for maintenance purposes. The following table summarizes the electrical and SCADA control costs.

**Table 5-11**
**Electrical and SCADA Control Costs**

| Electrical Component | Quantity/Description | Total Cost |
|---|---|---|
| **Substation** | | |
| Power Line Transmission | Underground Installation of two 345 kV power lines (Illinois Power) | $50,000 |
| Transformers | 345 kV primary to 4160 V secondary, 3 phase 60 Hz AC, installation, 75 kVA or larger tertiary transformer (480 VAC and below) | $200,000 |
| Switch Gear and Appurtenances | Includes necessary switches, conduit, transformer concrete foundation, connections to 4160 VAC motor controls and panels | $100,000 |
| **Motor Controls** | | |
| Soft Voltage Start System | 3 Soft Start Systems for 500 HP motor size, 1 VFD with softstart capability for 600 HP motor, Motor Control Panel (similar components to plant) | $304,000 |
| Panel Control Center | Panel control for valves, instruments, chemical feed pumps, accessories | $81,000 |
| **SCADA System** | | |
| Computer and Software | Pentium 333 Mhz Computer with CDW Drive for data storage, Interloution Fix 32 software package and hardware components | $25,000 |
| SCADA Programming, Setup, Start-up | 300 hours per project engineer, two engineers assigned to project ($30,000), installation by electrician ($30,000) | $60,000 |
| **Total** | | **$820,000** |

**Table 5-12**
**Electricity Costs for Different Pump Configurations**

| Electricity Cost | Pump Configuration | Cost |
|---|---|---|
| Maximum Day Usage | 1 Pump at full operation, 3 MGD (11,000 $m^3$/d) permeate production | $250 per day (winter)<br>$433 per day (summer) |
| | 3 Pumps at full operation, 10-11 MGD (38,000-42,000 $m^3$/d) permeate production | $750 per day (winter)<br>$1,300 per day (summer) |
| Total | | $250 per day minimum<br>$1,300 per day maximum |

### Treatment and Building Modifications

When originally built, the existing treatment system at the South WTP was designed for additional capacity. Extra appurtenances and land were originally appropriated for future expansion of the treatment facilities as per design drawings. The future system included two 36-inch (91 cm) diameter pipe connections at the east end of the filter clearwells for additional filters. Currently, the pipe connections are plugged.

To incorporate the RO facilities with the existing building, the east side of the operations building was chosen for this preliminary design for several reasons. These include:

1. Ease of connecting to the treatment facility.
2. Road access for construction and long-term chemical supply/maintenance access.
3. Nearby utilities (sewer, electrical) to avoid long trenching and complicated design considerations.
4. Promote system operation and maintenance by keeping the system as part of the operations building compared to an isolated building elsewhere on-site.
5. Building exterior and construction could be matched to past construction.

The modifications to the site would include:

1. A 25-30 foot (7.6-9.1 m) wide by 80-foot long (24.4 m) wetwell constructed in a similar fashion (floor elevation of 611 feet) to the existing reservoir pump wetwell.

The two 36-inch clearwell pipe connections would be extended into the wetwell and valved so that each clearwell can be operated independently.

2. Above the wetwell, a pump area and chemical storage/clean-in-place area would be below grade. This area would allow for maintenance of the pumps and operation of the clean-in-place system. A manway access would allow for pump maintenance within the wetwell.

3. At grade, the new building would be near to the level of the operations building. The main level would include all the chemical storage and feed equipment, electrical controls, and the RO system. The chemical storeroom for the sulfuric acid would be isolated from outside exposure both from a spill and fume standpoint with ventilation and spill containment/neutralization equipment. Electrical controls would be mounted close to both pumps and electrical supply but away from areas where inundation could occur. Pipe trenches would supply the necessary connections to the RO skids and draining for the system. Open air access to the lower pump room level through hatchways would allow for a crane lift for maintenance and repairs.

4. Construction with prefabricated concrete slabs, steel beams, and masonry would allow for fast construction and be able to match the operation building exterior. A slab for the additional substation would be poured just northeast of the clearwell.

5. An access road would be included to the north of the new construction. Should trenching be completed for the installation of the electrical services, the trench would be brought back to grade with materials suitable for an access road (asphalt or concrete).

6. Connection of a 24-inch sewer main to the existing 48-inch main in the vicinity for retentate disposal.

Treatment modifications would include:

1. The connection of a 24-inch diameter pipe to the reservoir feed line.

2. No drastic changes in treatment are planned. Minor changes to the amount of lime softening or SHMP addition would require further refinement with operation.

3. Reservoir feed pump operation would be decreased as necessary to maintain clearwell levels.

### Floor Layout of New Facilities

To develop a preliminary opinion of probable costs, floor layout drawings were prepared based on a skid design. The floor layout design is functional for both operation and maintenance. Figure 5-58 shows the layout configuration and a view of the facilities after construction. This layout can be modified prior to final design. In particular, it may be desirable to move the new chemical storage facilities to the north side of the addition.

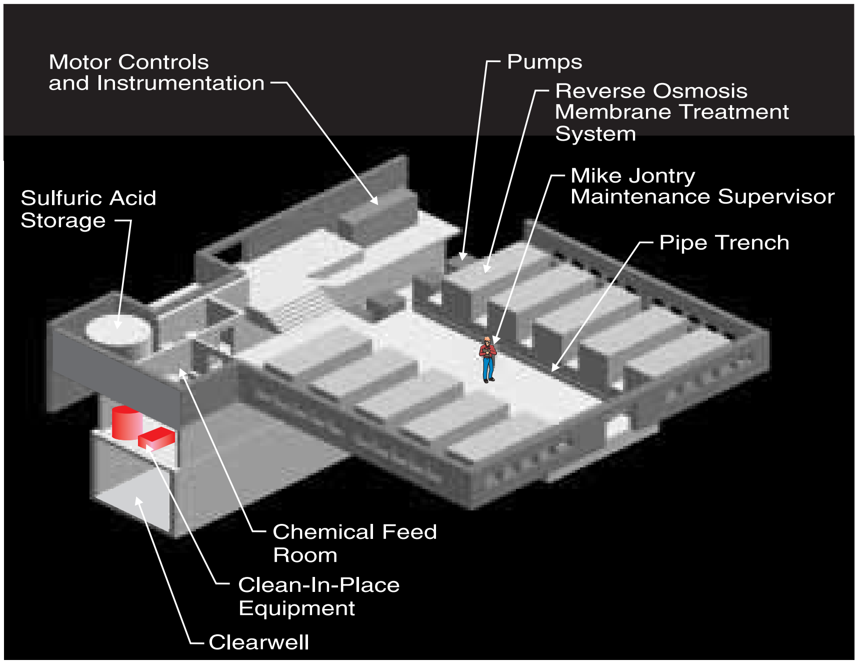

**Figure 5-58**
**City of Decatur 3D Rendering of Reverse Osmosis Treatment Facility**

### Construction Timeline and Costs

The City of Decatur is required to meet an EPA compliance order by the year 2001 with a plan or water treatment process set in place to reduce nitrate levels below the MCL of 10 mg/L NO3-as N throughout the year. Therefore, time is a consideration in meeting the EPA's deadline. Figure 5-59 presents a probable timeline for completion of such an RO treatment facility before January 1, 2001.

Depending on the actual project start date, this probable timeline should be modified to meet any time constraints. Installation times will vary between design options, weather, and other contingencies. Scheduling with equipment manufacturers' can be completed in more detail when the authorization for work has been given. Should work on this project be delayed, the schedule should be pushed back accordingly. However, consideration for seasonal-type work (i.e., construction) may be required.

Based on the preliminary design, an opinion of probable for construction costs was developed for the building. These costs include the work to design, prepare, and install all systems in place and in working order. Pictures of the operations building interior and exterior aided in the cost opinion. Existing drawings helped to determine the necessary design elements for this report.

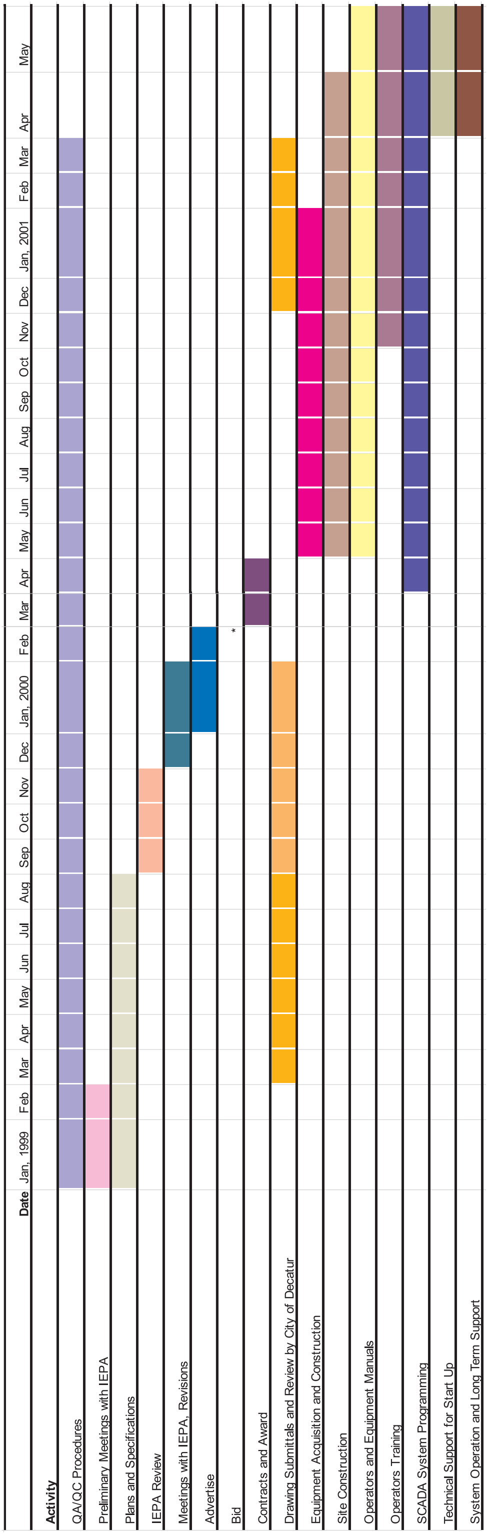

**Figure 5-59**
**Probable Construction Timeline**

The estimated costs for enclosing the RO treatment system and wetwell construction are detailed in Table 5-13.

**Table 5-13**
**Estimated Costs for Enclosing the RO Treatment System and Wetwell Construction**

| Construction Component | Description | Costs |
|---|---|---|
| Concrete | All concrete slabs, walls, and foundations to support equipment | $760,000 |
| Masonry | Exterior and interior masonry walls | $129,000 |
| Roof | Precast and membrane roof for enclosure | $206,000 |
| Site Excavation | Possible open cut, backfill, and dewatering operations | $163,000 |
| Trench Covers | Protection of pipe trench for safety purposes | $27,000 |
| Doors | Standard doors for room and exterior exits | $15,000 |
| Hoist | For pump and equipment maintenance on I-beam rail system | $10,000 |
| Paints and Coatings | Paint finishes for interior surfaces | $47,000 |
| Piping and Miscellaneous Finishes | Epoxy type coatings or equal for wear resistance and aesthetics | $25,000 |
| Stairs | OSHA approved stairs for access to different floor elevations | $4,500 |
| Construction Sub-Total | | $1,386,500 |
| 20 % Contingency | For items not accounted for and demo purposes | $278,000 |
| Construction Total | | $1,664,500 |

## Cost Summary

With the preliminary design, an estimated budgetary cost for a full scale system can be used to determine the feasibility of this treatment option to other alternatives for nitrate removal. One disadvantage of this treatment option is the high cost associated with energy and chemical use for just nitrate removal. Some attention is also required to maintain the membrane system with cleaning and maintenance.

However, the one major advantage is that the RO treatment can be used to meet current and future regulations regardless of feed water quality from Lake Decatur. This is especially important considering that RO removes total organic carbon and disinfection byproduct precursors which could be more closely controlled in the future. Watershed management or ion exchange options for nitrate abatement fail to provide long term solutions as regulations change. Unlike ion exchange and chemical disinfection, RO is an invasive process that does not introduce a byproduct into the water. The RO would also serve as a second barrier to pathological contamination. This preliminary design and cost estimate would help to solve a number of anticipated problems. Future public demands for improved water quality could be met with RO treatment.

Table 5-14 summarizes this preliminary design cost estimate. An additional 35 percent contingency and professional services fee is used to cover additional costs related to the project not included in this report. The contingency is based on previous experience with similar projects.

**Table 5-14**
**Preliminary Design Cost Estimate**

| Cost Component | Budgetary Amount |
|---|---|
| Membrane Component System | $3,637,000 |
| Valves and Piping | $571,500 |
| Water Quality Instrumentation | $37,980 |
| Chemical Addition Equipment | $154,100 |
| Electrical Components | $820,000 |
| Construction Costs | $1,664,500 |
| **Subtotal** | **$6,885,080** |
| 35% Contingency and Professional Service Fees | $2,410,000 |
| **Total** | **$9,295,080** |

An optional budgetary opinion of $5 to $7 million for an RO system without civil or site work was produced by Water Services, Inc., located in Milwaukee, WI. Their budgetary opinion is included as an appendix to this report. All probable costs have been determined using list prices and mark-up as provided by either manufacturer or distributor to anticipate installation and setup charges to be incurred. Actual design and construction costs may be less than this opinion of probable cost.

### Possible Cost Recovery

Reverse osmosis membrane elements are a replaceable wear item within the RO treatment system. Over the span of four to five years, the membrane elements may lose their capacity to produce clean permeate. The reduced capacity is a result of membrane fouling and wear with chemical cleaning. After the time span, new membranes would be installed and the plant would continue production.

Membrane elements may lower in cost and may have some resale value to companies like ADM, Caterpillar, and other manufacturers in the area that use RO technologies for production purposes. Membrane replacement would currently cost about $1.05 million dollars. The cost of membrane elements could decrease with increased use of the process and development in the future. RO prices have significantly reduced in cost over the past ten years. This trend may continue. Also, any revenue generated from the sale of spent RO elements would decrease this cost. The spent RO elements can be used in processes for metal plating, degreasing and paint preparation solvent recycling, and boiler water softening. Other applications such as heat exchangers or turbine generators could be used to recover energy costs.

### Membrane Storage

Membrane storage with a solution of glycerol, water, and sodium metabisulfite (as mentioned in the report) can be used to store the membranes for long-term storage periods Storage could last from September to February when nitrate concentrations are well below the MCL level of 10 mg/L $NO_3^-$ as N. The 8.5 mg/L trigger level could be used as a set point to begin preparing the RO system for operation and storage.

# 6

# CONCLUSIONS

One of the project objectives was to show the effectiveness of membrane technologies in removing nitrate and atrazine from a natural surface water source. Regardless of feed water quality variations, nitrate and atrazine were found to be significantly reduced (97% reduction of nitrate, and atrazine reduced below detection limits) using reverse osmosis membrane elements. Nitrate rejection rates greater than 90% can be achieved with polyamide elements specifically designed for overall salt rejections of 98% (CPA2, HR-PA) or better (TFC-HR). Energy saving (TFC-ULP, ESPA) and softening membranes (TFC-S), while effective in removing hardness and alkalinity, exhibited lower nitrate rejections of more than 80%. With lower nitrate rejections, higher recoveries were achieved. The cellulose acetate membrane (HR-CA) used in this study showed lower nitrate rejections (<71%) with low recovery (<11%) and was not considered feasible for this particular application.

Even though all elements were not exposed to the same levels of atrazine, the high rejection membranes reduced atrazine levels below detection limits. In particular, the TFC-HR, CPA2, and HR-PA membranes reduced atrazine below 0.1 ppb regardless of feed water concentrations. Future testing with the softening and ultra low pressure membranes should be conducted to see if atrazine removal is reduced.

While the UF pilot possibly improved RO pretreatment for longer term usage by reducing NOM/TOC concentrations, nitrate removals by the RO elements were slightly decreased or not improved with prior ultrafiltration. The UF pilot operation improved throughout the project. Further experience was gained in using ultrafiltration with raw lake water and remote operation and control. The automation software allowed for continuous monitoring and operation that was beneficial to data collection. The UF pilot also proved again that pretreatment with coagulation aids was necessary to prevent rapid build-up of organic and cake fouling on the membrane. However, the required chemical cleanings (every 3-5 days of operation) would increase operational costs. When compared to the UF permeate, the Decatur SWTP product water showed similar removals of TOC and turbidity with lower total hardness and alkalinity.

With such small differences in finished permeate water quality when using conventional tap water and UF permeate, UF does not appear to be an applicable pretreatment for RO to remove nitrates in Decatur, Illinois. Removal of TOC and atrazine may have been enhanced if a PAC/Coagulation/UF system was implemented prior to the RO plant. Other full scale plants could use the PAC/UF pretreatment option before RO to reduce organic fouling with high TOC waters and increase the operational life of the RO elements.

In order to implement RO membrane filtration at Decatur's SWTP, the following items could be required for a full scale RO plant:

- Dechlorination of the Tap Feed Water prior to RO (if free chlorine added prior to filtration)
- SCADA system to monitor and control system parameters
- Permits for construction and operation of the RO plant
- Permits for retentate disposal (as necessary)

While long term cleaning frequencies could not be determined during this project, chemical cleaning of the RO elements is usually conducted every six months according to the manufacturer's specifications and cleaning regime. Increased fouling of the membrane during specific time periods (i.e., storm events and increased feed turbidity) may require more frequent cleaning.

Overall costs for implementing an RO system were estimated using a design feed flow of 18 MGD (68,400 $m^3$/d). At 18 MGD, the design maximum nitrate concentration of 15 mg/l $NO_3$ as N could be removed to meet the MCL for nitrate with rejections greater than 95.67% (TFC-HR element). Overall costs, without the membrane elements included, could be expected to be \$0.56/1000 gallons (\$0.15/$m^3$). While the cost of RO may be high, the secondary benefits from softening (>98% hardness removed) and TOC removal (>95% TOC removed) could decrease chemical costs associated with lime softening and enhanced coagulation.

Overall, this project proved to be worthwhile in demonstrating the potential use of UF and RO to remove compounds like nitrates from a surface water. This project served more as a feasibility study, and there is still the possibility for future work. Longer testing of both the UF and RO elements would allow for better conclusions to be made on cleaning frequencies and contaminant removals. With the use of the larger 8.5 in. elements, full-scale testing (1 MGD) (3,785 $m^3$/d) may be appropriate to familiarize management and operators with full-scale membrane filtration while readjustments to the system design can be made if flux and rejection results differ from this pilot study. Future studies on particle fouling and surface fouling characteristics will hopefully offer more opportunities for high resolution surface analysis and improved options for increasing the membrane life. RO offers a cost-effective solution that has the potential to improve the lives of people drinking water with high levels of contaminants.

# 7

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