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Precipitative Softening and Ultrafiltration Treatment of Beverage Water

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Precipitative Softening and Ultrafiltration Treatment of Beverage Water

by

Jorge T. Aguinaldo

A thesis submitted in partial fulfillment
of the requirements for the degree of
Master of Science in Environmental Engineering
Department of Civil and Environmental Engineering
College of Engineering
University of South Florida

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Precipitative Softening and Ultrafiltration Treatment of Beverage Water

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ABSTRACT

Lime softening, chlorination, clarification and filtration have been long recognized treatment processes for beverage water specifically the carbonated soft drink (CSD) because it provides consistent water quality required for bottling plants, however these processes are becoming uneconomical and causes more problems than the benefits they offer. These processes require very large foot print, occupy large plant volume, and generate large volume of sludge which causes disposal problems. Chlorination produces trihalomethanes (THMs) and other by-products which are detrimental to health and imparts tastes to the final products. Using the newly developed submerged spiral wound ultrafiltration membranes in conjunction with lime softening may replace the conventional lime softening, clarification and filtration processes.

This research was conducted to demonstrate the feasibility of integrating immersed ultrafiltration (UF) membrane with lime softening. The objectives of this research was to achieve the water quality required by the CSD bottlers; determine the relationships of operating parameters such as pH and membrane flux with trans-membrane pressure (TMP), and membrane permeability; determine the optimum dosage of lime; evaluate the operating parameters as basis for the design and construction of the full scale plant; and predict the membrane cleaning intervals.

A pilot unit consisting of lime reactor and UF system was designed and built for this research. The pilot unit was operated at various pH ranging from 7.3 to 11.2 and at membrane flux rates of 15, 30 and 45 gfd. The pilot unit was also operated at the CSD bottler's operating conditions which is pH 9.8 at flux of 30 gfd. The pilot unit operated for a total of 1800 hours. The raw water source was from city water supply.

The filtrate from the pilot unit achieved alkalinity reduction to 20 to 30 mg/L preferred by CSD bottlers, with lime dosage close to the calculated value. The filtrate turbidity during the test was consistently within 0.4 to 0.5 NTU. The TMP values obtained during the test ranges from 0.1 to 2.5 psi, while the permeability values ranges from 18.19 to 29.6 gfd/psi. The increase in flux results to corresponding increase in TMP, and increase in operating pH, increases the rate of TMP. Permeability decreases with increasing operating pH. The TOC reduction ranges from 2.6 % to 15.8% with increasing operating pH. No scaling of the UF membranes was observed during the test. Thirty days UF membrane cleaning interval was predicted. The results from this research can use as the basis of designing and operating a full scale Lime Softening UF Treatment Plant.

Chapter One

Introduction

The ingredients used in carbonated soft drinks (CSDs) including water are approved and closely regulated by the US Food and Drug Administration (FDA), but there are no defined water quality standards as long as it meets the federal and local drinking quality standards. The source water for soft drink manufacture is typically the municipal water supply, and at minimum it should comply with the primary and secondary National Drinking Water Standards. The municipal water supply however vary from one area to another and may not be able to provide consistent quality required for soft drink manufacture, therefore additional treatment is necessary. Most of the impurities that concerns the carbonated soft drink bottlers are those that affect the appearance and flavor of the product. The important ingredients of CSDs, aside from water are sugar, flavors and carbon dioxide. Carbon dioxide is the essential characterizing ingredient in all soft drinks, the “tingly fizz” which gives a refreshing taste. When CO₂ is dissolved in water, it imparts a unique taste. Natural carbonated or effervescent mineral water was popular because the minerals dissolved in water were believed to have beneficial medical properties. By 1800, artificial effervescent mineral water were introduced in Europe and North America. Then the innovative step of adding flavors to these popular “soda water” gave birth to the soft drink beverage we enjoy today.

Originally, carbon dioxide was made from sodium salts and the carbonated beverage became known as “soda water” (American Beverage Association, 2005).

Lime softening is the most common water treatment process in CSD bottling plants. The typical water treatment process includes pre-chlorination, lime softening with ferric salt dosage, media filtration or manganese greensand filtration. The addition of coagulants, such as ferric salts in lime softening process promotes better sludge settling and also can reduce organic matter in the raw water. The unit processes above when accompanied by super chlorination followed by activated carbon filter and polishing filter comprise the conventional system for CSD product water (Morelli 1994).

Lime softening has been the choice of bottlers because it provides consistent water quality suitable for bottling operations, regardless of the raw water quality. Recently, many bottling plants are replacing the lime-soda softening with other processes such as reverse osmosis, microfiltration and/or ultrafiltration. These processes, in most cases provide treated water that meets the quality requirements of the bottling. However, there are cases that lime softening can not just be replaced by reverse osmosis, especially when the high concentration of hardness in the raw water limits the recovery in the RO system. RO is excellent in reducing total dissolved solids, hardness and alkalinity in raw water, but it requires pre-treatment such as media filter or membrane microfiltration or ultrafiltration. The major CSD bottlers require the raw water feed to the RO system to be chlorinated to prevent biological fouling of the RO membranes. The drawback of chlorination of RO feed water is the breakdown of organic matter into smaller molecules forming trihalomethanes (THMs), which are not rejected by the RO membranes.

The activated carbon, as part of the process removes residual chlorine and most of the organic matter that may impart off-taste and odor in the final product.

Table 1
Selected Contaminants Limits in the National Primary and Secondary
Drinking Water Standards
(EPA, 2003)

Primary Drinking Water Standards	
Turbidity:	< 1 NTU or < 0.3 NTU in 95% of daily sampling in a month
Chlorine (as Cl ₂):	MRDL = 4.0 mg/L
Cu:	1.0 mg/L
Pb:	0.015 mg/L
As:	0.010 mg/L
F:	4.0 mg/L
Hg:	0.002 mg/L
NO ₂ (as N):	10 mg/L
NO ₃ (as N):	1.0 mg/L
TTHMs:	0.08 mg/L
Secondary Drinking Water Standards	
Aluminum:	0.05 to 0.2 mg/L
Chloride:	250 mg/L
Sulfate:	250 mg/L
Color:	15 (CU)
Foaming Agents:	2.0 mg/L
Iron:	0.3 mg/L
Manganese:	0.05 mg/L
Odor:	3 threshold number
pH:	6.5 - 8.5
TDS:	500 mg/L

In US there is no standard water quality specifically for CSD, because every bottler has its own quality control requirements. At minimum, the water supply to CSD bottling plant should meet the National Primary and Secondary Drinking Water Standards as shown in Table 1.

In 1958, the Society of Soft Drink Technologists carried out survey among bottlers on the quality of water they require for their plant (Morelli 1994). The result of the survey is shown in Table 2. In Canada, the Agriculture and Agri-Food Canada issued a water quality guideline (Agriculture and Agri-Food Canada 2000) for food and beverage industry which specifically includes carbonated beverages as shown in Table 3. In the CDS beverage industry variations in taste could be caused by the variations in the alkalinity of the product water. Lime softening seems to be the only treatment process that can provide consistent quality of treated water. Lime softening primarily will reduce and/or maintain alkalinity in the treated water to less than 50 mg/L as CaCO₃ and a pH range of 8 to 9, however most CSD bottler operators are aiming for 20 to 30 mg/L alkalinity because it provides better yield (or less rejection of final product due to off taste). Another criterion is the hydroxide concentration which should be between 2 to 7 mg/L as CaCO₃ based on calculation using Phenolphthalein and Methyl Orange Alkalinity values.

Table 2
CSD Bottlers Water Quality Survey
(Morelli 1994)

	Max.	Min.	Avg.	Median Avg.
Turbidity, NTU	10	0	2.3	2.0
Color, CU	20	0	4.8	3.5
Organic, Matter, ppm	5	0	0.4	0
Taste & Odor	0	0	0	0
Chlorine, ppm	0.2	0	0.03	0
Alkalinity, ppm CaCO ₃	130	0	70	50
Sulphates, ppm	900	0	240	225
Chlorides, ppm	525	0	210	225
Iron & Manganese, ppm	1.8	0	0.4	0.1
Copper, ppm	0.05	0	--	--
Calcium, ppm	500	25	182	150
Magnesium, ppm	650	0	160	80
Sodium, ppm	900	500	--	--

Table 3
Canadian Water Quality Guidelines for Carbonated Beverage
(AAFC, 2000)

pH	< 6.9
Color	< 10 Hazen Units
Turbidity	1 – 2 NTU
Taste, Odor	N.D.
TDS	< 850 mg/L
Iron	< 0.1 mg/L
Manganese	< 0.1 mg/L
Carbonate	< 5 mg/L
Sulphate	< 200 mg/L
Chloride	< 250 mg/L
Fluoride	0.2 to 1.0 mg/L
Hardness	200 to 250 mg/L
Alkalinity	50 to 128 mg/L

The CSD bottlers apply the multi-barrier concept in treating raw water into product water. Shachman (2004) defines multi barrier system as an orderly series of reliable processes that, in a complementary and incremental manner, completely removes or reduces targeted raw water adverse quality factors to acceptable levels, at lowest practical cost. To apply this concept, many CSD bottlers are incorporating membrane treatment processes, such as ultrafiltration, microfiltration, nanofiltration and reverse osmosis in their existing processes. In many cases, the membrane processes alone cannot provide the required product water quality. It is common to find membrane treatment after lime softeners. Talking to CSD quality personnel and plant operators, the majority expressed desire to simplify the lime softening and membrane processes, possibly to combine both processes. It is common for UF systems in CSD bottling plants to dose coagulant, such as ferric sulfate or ferric chloride. A novel approach is to dose lime to achieve softening. The application of lime for softening is not

the same as dosing ferric salts. By combining both the lime dosing and membrane treatment, it will be possible to reduce the lime dosage and sludge production, and achieve the desired product water quality at reduced cost. The membrane ultrafiltration is a barrier that can physically prevent microorganism from passing through into the treated water. The existing lime softening facilities can be integrated with ultrafiltration. Additional minor modifications will increase the existing plant's capacity.

The purpose of this research is to demonstrate the feasibility of combining lime softening with membrane ultrafiltration to achieve the water quality required in the bottling process with minimum usage of chemicals and eliminating continuous chlorination of the raw water.

Chapter Two

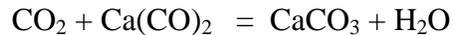
Background

2.1 Lime Softening

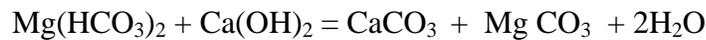
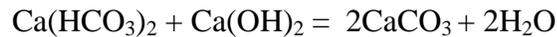
Lime softening has been long recognized as an effective process to reduce calcium and magnesium hardness in water by adding CaO or Ca(OH)_2 (lime) and/or Na_2CO_3 (soda ash) to precipitate calcium as CaCO_3 and magnesium as Mg(OH)_2 . It will also remove CO_2 in the water. In addition to hardness, other impurities such as iron, manganese, fluoride, phosphates, heavy metals, silica, chloride and total dissolved solids in the water are also removed with the addition of lime alone or in combination with other chemicals such as alum, sodium silicate, ferric and ferrous salts, flocculant, etc. The elevated pH required in the process also inactivates many microorganisms. Lime softening has been known to remove natural organic matter (NOM) in water specifically trihalomethane (THM) precursors (Collins, Amy, and King 1985). Lime softening was found to remove significant fraction of fulvic acid extracted from ground water (Liao and Randke 1985), and the NOM removal was achieved by the adsorption onto calcium carbonate and magnesium hydroxide formed in the process. EPA (1999) recommendation to enhanced total organic carbon (TOC) removal using precipitative softening is to provide the conditions that favor the formation of magnesium hydroxide and small calcium carbonate particles. This can be achieved by elevating the pH to 10.8 or higher, delaying carbonate addition and sludge recycling.

The degree of precipitation of calcium, magnesium and other impurities depends on the operating pH. Soda ash is also added to precipitate non-carbonate hardness and to precipitate excess lime. Caustic soda is also added to adjust the operating pH and promote precipitation of calcium and magnesium. This process is often called caustic soda softening. This process is applicable if there is enough calcium in the raw water to complete the softening reactions. The typical reactions in lime, or similar precipitative softening processes are:

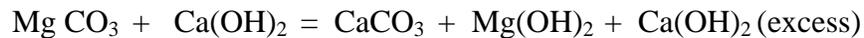
Lime as CaO when water is added becomes Ca(OH)₂



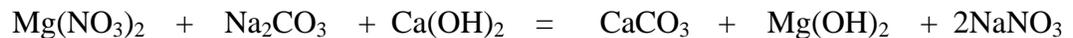
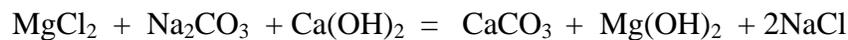
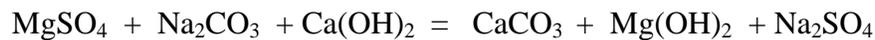
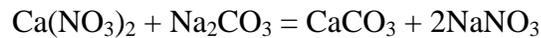
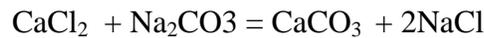
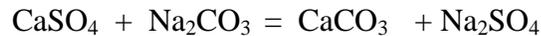
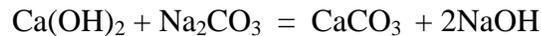
At pH 9.5 or above the following reaction will occur:



At pH 11 or above



Reactions with soda ash



In lime softening, additional and/or excess chemicals are often added to increase the mass of sludge to promote settling.

2.2 Limitations/ Problems Associated with Lime Softening

The lime softening although reliable and being used in the beverage industry for almost a century, has its limitations and problems. Some of the limitations and problems associated with lime softening are:

- Disposal of large amount of sludge generated by the process
- Requires larger foot plant print for the lime reactor, as well as the sludge handling equipment, lime preparation and storage facilities.
- Additional chemicals are required to promote settling of the sludge and solids.
- Requires media filtration after clarification.
- The lime softening plant should be continuously running and requires longer time to stabilize after start-up.
- The lime-soda softening is more expensive compared to other competing processes.
- There are very limited companies now specialized in the manufacture of lime softening systems.

2.3 Precipitative Softening

The USEPA, acting on the 1986 Amendment to the Safe Drinking Water Act (SDWA), set maximum contaminant level goals (MCLGs) for a variety of contaminant that is present in drinking water. The disinfectants and disinfection byproducts (DBPs) are among the list of contaminants for regulated in the Disinfection Byproduct Rule (DBPR). USEPA developed treatment techniques or a maximum contaminant level (MCL) that is as close to the MCLG as is feasible with the use of the best available technology (BAT). As part of the DBPR has USEPA, included a treatment technique requirement to remove natural organic matter (NOM) which serves as the primary precursor for DBP formation. The goal of this pre-treatment technique is to provide additional removal of NOM, measured by total organic carbon (TOC). The USEPA Enhanced Coagulation and Enhanced Precipitative Softening Guidance Manual define enhanced coagulation as a term to represent the process of obtaining improved removal of DBP precursor by conventional treatment whereas enhanced softening refers to the process of obtaining improved removal of DBP precursors by precipitative softening.

In the implementation of the enhanced coagulation and enhanced softening requires process modification in the existing plants and will have some impacts which may be either beneficial or detrimental. USEPA cited some of the impacts as:

- Inorganic constituents levels (manganese, aluminum, chloride and sodium)
- Corrosion control
- Disinfection
- Particle and pathogen removal

- Residuals (handling, treatment, disposal)
- Operation and maintenance
- Recycle streams

In addition to the above, from the beverage plant operator's point of view, the impacts are:

- Maintaining treated water quality suitable for beverage bottling operations that are often to higher quality standard compared to the municipal drinking water quality.
- Operating costs
- Limited plant area to implement process modification
- Additional cost associated with the plant upgrade

Precipitative softening specifically lime process, comes in various forms and variations. Humenick (1977) listed four process types, based on the amount of chemicals added:

- Single-stage lime process is used when the source water has high calcium, low magnesium carbonate hardness (usually less than 40 mg/L as CaCO_3), and no noncarbonate hardness. Single-stage lime softening is not intended for magnesium hardness removal. Lime is added upstream of the reactor in a separate flash mixing chamber or into the reactor-clarifier. The pH of the water leaving flash mixer is about 10.2 to 10.5.

- Excess lime process is used when the source water has high calcium, high magnesium hardness, and no noncarbonate hardness. Excess lime process can be single or in two stages. Excess lime is added to precipitate magnesium carbonate hardness as magnesium hydroxide. The pH of the water after flash mixing will be from 10.2 to 11.2. Above pH 10.2, causticity will be present.
- Single-stage lime-soda process is used when the source water has high calcium, low magnesium hardness, (usually less than 40 mg/L as CaCO_3), and some calcium non-carbonate hardness. This is similar to the single-stage lime process, except that the soda ash is added for the removal of non-carbonate hardness. The soda ash is added in the flash mixer or sequentially after the lime has been added.
- Excess lime-soda process is used when the source water has high calcium, high magnesium carbonate hardness and some non-carbonate hardness. The addition of soda ash in the excess lime process will allow removal of non-carbonate hardness, while removing calcium and magnesium hardness. Excess lime-soda process can be in one or two stages, however two stage process is common practice, because the soda ash added in the second stage will remove the excess lime.

In addition to the above, other variations of lime softening include the following:

- Pellet softening (Van der Veen, C. & Graveland, A., 1988) uses fluidized bed of grains on which crystallization of CaCO_3 takes place.

The softening reaction takes place in the presence of suspended bed of fine sand or crushed CaCO_3 that acts as catalyst. Feed water and chemicals enter tangentially at the bottom of the pellet reactor chamber and mix immediately. The treated water rises through the reactor in swirling motion. The upward velocity is sufficient to keep the sand fluidized. The precipitated hardness particles attaches to the surface of the sand grains and the sand diameter increases. Large grains are continuously removed.

- Ultra high lime softening (Batchelor, B; Lasala, M. McDevitt, M; Peacock, E., 1991) is another variation of lime softening and is used when the source water has high calcium and magnesium hardness, and high silica concentration. Excess lime is added to the reactor to increase the operating pH to above 11. Ultra high lime softening is usually is a two stage process.
- Other modification of lime softening is the addition of caustic soda instead of lime to achieve the reaction pH.

In all the processes above the softening is achieved by precipitation of CaCO_3 and Mg(OH)_2 at elevated pH, where the solubilities of CaCO_3 and Mg(OH)_2 are relatively low. The various process modifications in lime softening also enhances the removal of the precipitate through effective settling or, in case of pellet softening attachment to the fine sand grains

2.4 Ultrafiltration

Ultrafiltration is a pressure driven membrane process, where the source water is passed through a membrane with nominal pore size of 0.01 to 0.1 μm , and suspended solids, colloidal particles, bacteria and other particles are retained. Ultrafiltration also removes high molecular weight organic matter. The typical ultrafiltration membranes have a typical molecular cut-off of 150,000 daltons (1 dalton or Da = 1/12 mass of one atom of Carbon-12), however through the addition of coagulants, it can effectively remove organic matter with molecular weight down to less than 20,000 daltons. The addition of coagulant in the form of ferric salts, poly aluminum chloride or alum is common in ultrafiltration process. The addition of lime in the feed of ultrafiltration membrane was never been reported in the literature, but there were published reports integrating pellet softening with UF membrane treatment (Li, Jian, and Liao, 2004). In most membrane processes especially in reverse osmosis and nanofiltration, CaCO_3 scaling is a common problem. In treating hard water using ultrafiltration, the precipitation of CaCO_3 can be a problem, especially occurring in capillary or small diameter tubular UF membranes. The development of the Spirasep UF membrane, which air-scoured immersed membranes in spiral configuration developed by Trisep, will minimized the build up of scale in the UF membrane surface. Compared to RO or NF, there is no change in salt concentration in the membrane surface, therefore formation of scale will be minimized. The Spirasep membrane is similar in appearance to 8" diameter x 40" length RO membrane, made of polyethersulfone, and with effective membrane area of 178 ft^2 . The operating pH is from 4 to 11 on continuous basis and pH of 2 to 12 for cleaning. The Spirasep membrane has chlorine tolerance of 2,000 mg/L.

In this research the manufacturer's operating guidelines were strictly followed because the UF unit is a working commercial unit with single UF element and to limit the variables. Among the operating conditions maintained were the following:

- Continuous aeration at the recommended aeration rate of 0.02 to 0.05 scfm per square feet of membrane area. Continuous aeration was recommended for water with high suspended solids concentration.
- Back flushing was set every 15 minutes with 30 seconds duration at the rate of 45 gfd.
- Trans-membrane pressure was defined and measured as per the membrane manufacturer's guidelines as shown in Appendix B.

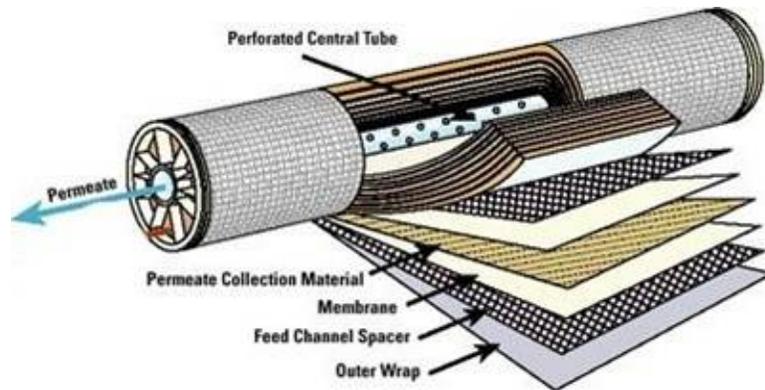


Figure 1 Spiral Wound Membrane

Chapter Three

Materials and Methods

3.1 Experimental Plan

The objective of this experiment is to demonstrate the applicability of combining lime softening with ultrafiltration membrane to produce water that meets the beverage water quality. The specific objectives are:

- Reduce hardness in the feed water and, at the same time, maintaining alkalinity of the treated water to less than 50 mg/L as CaCO₃
- Determine the relationships of operating pH and membrane flux with trans-membrane pressure and membrane permeability.
- Compare the lime dosage in this research with the lime dosage used in conventional lime softening plant, treating similar water source.
- Evaluate the operating parameters important in designing a full scale plant. These includes membrane flux rates, permeability, recovery, backwash intervals, cleaning intervals, and trans membrane pressure.

3.2 Pilot Lime Softening Ultrafiltration Unit

The pilot lime softening ultrafiltration system was designed and built by Doosan Hydro Technology, Tampa, Florida. The details of the plant are described in Appendix A. The pilot unit is a full scale commercial operating plant with one SpiraSep UF membrane immersed in a reactor tank.

3.2.1 Lime Reactor

The lime reactor is a polyethylene cylindrical conical bottom tank, with maximum capacity of 200 gallons, to allow 30 minutes retention at the maximum flow of 5.7 gallons per minute (gpm). The tank was provided with discharges at three different levels for the different flow rates. The elevation of the lime reactor is adjustable, in order to allow gravity flow into the membrane reaction tank. The lime solution or slurry was fed by a BLUE-WHITE Model A-100N Peristaltic Metering Pump, with a maximum capacity of 2.3 gallons per hour (gph). The flow rate of the metering feed pump was controlled by the pH transmitter.

The lime slurry or solution was fed to the incoming raw water into the mixing chamber which directed the flow to the bottom of the lime reactor. The mixing chamber was provided with a mixer (FPI Model PM1/20 PE) driven by a 1/20 horse power (hp) electric motor. The precipitate, or sludge, settles at the bottom of the tank. Sludge was expected to be carried over to the membrane reactor tank. The bottom of the lime reactor was provided with a connection for pumping out the sludge at scheduled interval. The pH sensor was installed at the inlet of the membrane reactor tank.

3.2.2 SpiraSep Ultrafiltration Membrane

The SpiraSep UF membrane manufactured by Trisep Corp. of Goleta, California is an immersed, negative-pressure ultrafiltration process, which will remove suspended solids, turbidity, viruses, bacteria, and some organic compounds. A typical SpiraSep system consists of an array of spiral wound elements submerged inside a process tank. The membrane elements are attached to a manifold assembly, consisting of a central permeate header with an array of membrane permeate ports, which connects to the SpiraSep membrane. A vacuum is generated by the suction of a centrifugal pump, creating the necessary net drive pressure to “pull” water through the SpiraSep membrane. Air is bubbled up through each membrane element via bubble diffusers, creating tremendous shear forces on the membrane surface that remove any suspended solids. A small amount of a coagulant is injected into the process influent. The enhanced coagulation process will help reduce organic fouling and improve TOC and color reduction.

Periodically (on a timed basis), permeate water is reversed through the membrane, or back flushed, to help further remove the accumulated suspended solids. This process also introduces a small amount of disinfectant to help control the microbial activity on the membrane surface. Concentrate is removed from the process tank, and is typically less than 10% of the influent rate. SpiraSep membranes can also be chemically cleaned through one of two processes: a periodic flux enhancement (PFE) or a flux recovery clean (FRC) procedure.

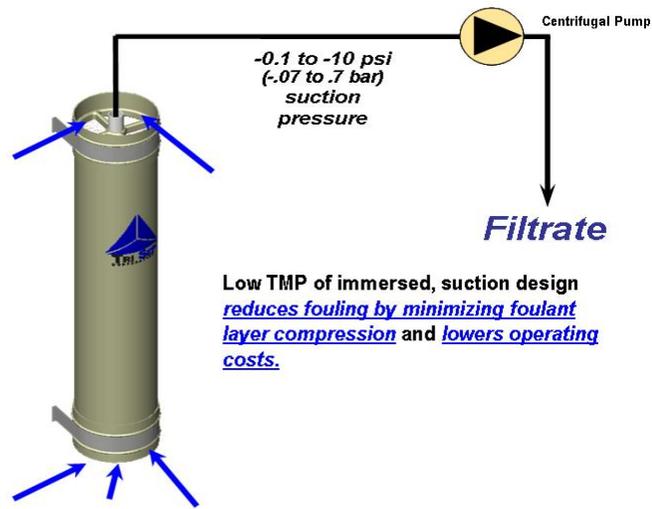


Figure 2 SpiraSep Immersed UF Membrane Configuration

The pilot plant was manually controlled and operated with several automated Features, such as backwashing. Feed from a pressurized source is delivered to the UF system, and is controlled by a feed control valve. A blower is operated continuously to deliver pressurized atmospheric air to the membrane element. Membrane backwashing is controlled by a timer, and is performed on a timed basis. Membrane cleaning is operator initiated.

3.2.3 Pilot Lime Softening Ultrafiltration Process Control Description

The feed water to the pilot unit was delivered to the lime reaction tank and was controlled by a control valve and rotameter. A sample line from the feed was connected to the in-line turbidity analyzer. Lime solution was added to the feed water at the flash mixing chamber. Lime was dosed by a peristaltic chemical dosing pump, drawing lime solution or slurry from a solution tank. The dosing rate of the chemical dosing

pump was controlled by the pre-set operating pH. The pH probe measures the pH of the water in the overflow. From the flash mixing chamber, water flows downward to the conical bottom of the lime reaction tank. A provision for another coagulant dosing was included, in the event that another coagulant will be added in conjunction with or to supplement the lime. The CaCO_3 and other precipitates settled in the conical bottom of the lime reactor tank and softened water overflowed to the UF process or membrane tank. Carryover CaCO_3 and/or precipitate were expected in the overflow.

Feed to the ultrafiltration unit results in two streams: filtrate and concentrate. Feed was introduced to the membrane tank from the overflow in the lime reaction tank. Once feed water was introduced to the membrane tank, the blower was turned on. The air flow was manually adjusted to provide the proper air flow rate to the element. The air flow rate was measured using a flow meter. The concentrate valve was set to obtain the proper concentrate flow rate.

Once the membrane tank was completely filled, the Process Logic Controller (PLC) will start the filtrate pump and open the concentrate valve. The UF filtrate pump provides the necessary net drive pressure to force feed water through the membrane surface. A self-priming centrifugal pump generates a vacuum, typically less than -10 psi, drawing water through the UF membrane surface. Filtrate flow was manually set with a control valve but pump operation is controlled via the PLC.

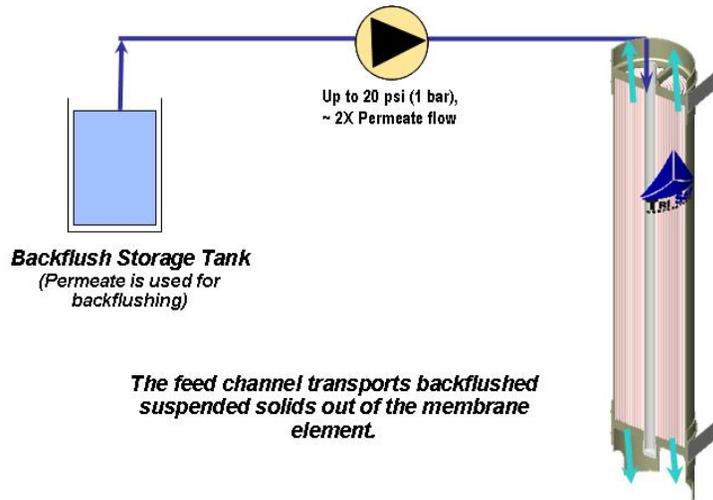


Figure 3 SpiraSep UF Membrane in Backflushing Mode

The filtrate pump flow rate was adjusted manually with the permeate control valve. The UF membrane was back flushed at set interval. The water required for the membrane back flush was taken from the UF filtrate tank and pumped to the membranes using a separate backwash pump. The backwash pump reverses the flow of water through the UF membranes. A membrane back flush was performed every 15 minutes for 30 seconds and is automatically controlled by the PLC.

Once filtrate production started, timers for the back flush frequency and Periodic Flux Enhancement (PFE) are started. The blower remains on running at the manually set value.

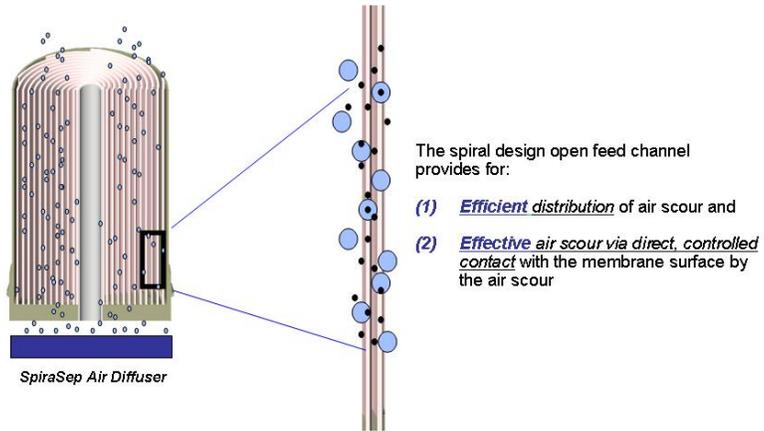


Figure 4 Spirasep UF Membrane Air Scour

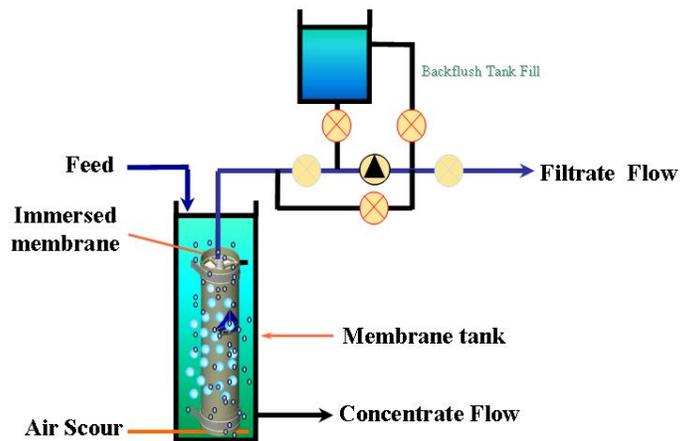


Figure 5 UF System During Filtration

When a back flush sequence is started, the automatic feed valve was closed, and the filtrate pump and blower were automatically turned off (concentrate valve remains

open). UF filtrate water and chlorine were then backflushed through the membrane for a period of about 30 seconds. A Variable Frequency Drive (VFD) adjusts the back flush pump speed, to the manually set value. Output of the metering pump was manually adjusted. Excess water introduced to the tank was removed via a tank overflow and/or concentrate line. Once the back flush sequence was completed, the back flush pump and chlorine metering pump were automatically turned off. The blower was turned on and allowed to operate for 10 – 15 seconds before the filtrate pump was restarted and the feed valve opened to allow normal filtrate production.

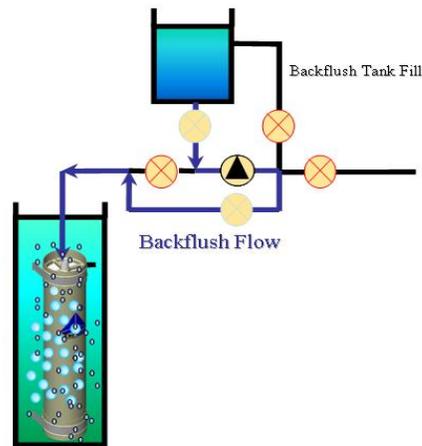


Figure 6 UF System During Backflushing

The UF membrane was continuously aerated to prevent and minimize membrane fouling. A blower takes atmospheric air and bubbles them up through individual membrane module via an aeration disc. The blower was operated using a VFD, and the motor speed is set manually. The operation of the blower was controlled by the PLC.

Air was delivered to the UF membrane through a coarse bubble diffuser. The air diffuser was attached to an aeration pipe. The aeration pipe contains a manual flow control valve and air flow indicator to ensure proper air flow.

Various chemicals were dosed for various system operations. Chlorine was dosed during each back flush, in addition to PFE and Clean-In-Place (CIP) processes. Sodium hydroxide was injected for just PFE and CIP processes. Citric acid was dosed for PFE and CIP processes. The flow rates of the chemical dosing pumps were set manually. Operation of the chemical dosing pumps during backwash, PFE, and CIP was controlled by the PLC.

Operating performance can be optimized through the use of PFE. A chemical solution was backwashed through the membranes in situ to perform a quick chemical treatment. This process was performed while the membrane tank was filled with process water, requiring approximately 20 – 30 minutes. This was done on a daily or every two days. When a PFE process was initiated, the feed valve was closed, and the filtrate pump and blower were turned off. UF filtrate and chemicals were then automatically back flushed through the membranes while they are still immersed in the feed water (i.e. membrane tank is not drained for this process). Excess water introduced to the tank was removed via a tank overflow and/or concentrate line.

During membrane cleaning, a cleaning solution was back flushed through the membranes until the filtrate tank was completely filled. The membrane was statically soaked in the cleaning solution for approximately 4 – 8 hours. A CIP process is typically performed once every 3 months for municipal water treatment. Actual CIP frequency is determined through pilot testing and actual plant operation. CIP is a manual operation. In high suspended solids environment like in lime softening CIP every 2-3 weeks

3.3 Chemicals

The chemicals used in the pilot test are:

- Hydrated Lime, $\text{Ca}(\text{OH})_2$, 93%, CAS 1305 – 78-8, technical grade
- Sodium Hypochlorite, NaOCl , 12% chlorine CAS 7681-52-9
- Sodium Hydroxide, NaOH , 45% CAS 1310-73-2
- Citric Acid Anhydrous 99.5%, $\text{C}_6\text{H}_8\text{O}_2$ CAS 77-92-9

A 3.2 % lime slurry was prepared by adding 32.24 grams of hydrated lime (93% $\text{Ca}(\text{OH})_2$) per liter of water mixed into the slurry tank. The 3.2 % lime slurry has a specific gravity of 1.020 or 2.84 Baume, which will be verified using a Hydrometer (Cole Palmer Cat# C-08287-55, range SG 1.000 to 1.225, Baume 0 to 26 deg).

The sodium hypochlorite (12% chlorine) was dosed at 10 mg/L during back flush and 100 mg/L during Periodic Flux Enhancement (PFE). The sodium hypochlorite solution for both the back flush and the PFE back flush were dosed by metering pumps drawing directly from the sodium hypochlorite container.

The citric acid crystals was dissolved in water at 200 grams/L solution. From this stock solution, the citric acid was dosed directly to the PFE back flush line at rate of 2 l/h. during CIP. The citric acid was dosed to the CIP line at the rate of 20 l/h.

Caustic soda, 45% solution was dosed at 0.1% or 1,000 ppm using chemical feed pump at a rate of 0.63 l/h drawing directly from the caustic soda container.

3.4 Experimental Procedures

The pilot unit was initially operated for one week without any chemical addition to stabilize the flow and calibrate the instruments. After one week the pilot unit was operated for approximately one month with varying dosage of lime to determine the conditions that can provide the desired water quality. The pilot unit was operated for another month at the selected optimum operating conditions. The lime slurry was dosed by peristaltic pump (Blue White Model A1N30F-6T) with maximum capacity of 1.25 gph (4.73 lph). This pump is capable of delivering lime up to 346 mg/L when operating at flux of 15 gfd and 120 mg/L when operating at 45 gfd. Operating flux of 15 gfd was selected to be the starting flux, based on previous pilot testing using other coagulants such ferric chloride, ferric sulfate and alum. Trisep recommends the following sustainable flux rates: For municipal secondary effluent: 15 to 18 gfd; municipal drinking water: 25 gfd; landfill leachate (with chemical precipitation): 15 gfd. The flux will eventually increase to 30, and 45 gfd. Lime slurry was dosed to achieve pH of 8.3, 9.4, 10.6, and 11.2 at the lime reactor overflow or discharge to the membrane tank. The various phases of testing were performed at the following schedule:

<u>Day</u>	<u>Flux (gfd)</u>	<u>pH</u>
1	15	Feed water pH
2	15	Feed water pH
3	30	Feed water pH
4	30	Feed water pH
5	45	Feed water pH
6	45	Feed water pH
7	15	8.3
8	15	8.3
9	30	8.3
10	30	8.3

11	45	8.3
12	45	8.3
13	15	9.4
14	15	9.4
15	30	9.4
16	30	9.4
17	45	9.4
18	45	9.4
19	15	10.6
20	15	10.6
21	30	10.6
22	30	10.6
23	45	10.6
24	45	10.6
25	15	11.2
26	15	11.2
27	30	11.2
28	30	11.2
29	45	11.2
30	45	11.2

The flux was set by controlling the flow through the filtrate pump through the adjustment of the filtrate control valve. During the test the trans-membrane pressure (TMP) was monitored through a digital pressure indicator connected to a pressure transmitter installed at the manifold between the UF membrane filtrate discharge and the suction of the filtrate pump.

The pilot testing log will include the following information: Date and time, actual flow rate reading, total flow (from flow totalizer), pH, temperature, raw water and filtrate turbidity, TMP or UF pump suction line pressure located at the same level as the water in the UF reactor tank. The net flow in each segment of test can be determined and used as basis of calculating the average permeability.

The flow is indicated by a SIGNET Model 8550 Flow Transmitter with digital flow indicator and totalizer, receiving signal from a SIGNET Model 515 flow sensor. The pH is indicated by a SIGNET Model 8750 pH transmitter with digital pH and temperature indicator, receiving signal from a SIGNET Model 2754 pH probe. The TMP is measured by local mounted EFFECTOR pressure transmitter/ indicator. The turbidity is continuously monitored by HACH Model 1720D Low Range Process Turbidimeter, provided with sample connections to allow turbidity measurement of either the raw water or the filtrate.

The permeability was plotted against elapsed time. The permeability was calculated as flux (in gfd) divided by the trans membrane pressure (psi). The permeability has a unit of gfd/psi. The TMP values were also plotted against time.

Composite samples of feed and filtrate were taken daily and were analyzed for pH, alkalinity, calcium and magnesium hardness, conductivity, turbidity, and total organic carbon (TOC). Sample of the water in the membrane reactor was also taken for suspended solids analysis.

3.5 Analytical Procedures

The analysis of the water samples were made following the EPA Methods and Standard Methods for the Examination of Water and Wastewater (APHA, AWWA, WEF, 1995). The water samples taken during the test were sent to Severn Trent Laboratories, Inc. (STLI) in Tampa for analysis. STLI is EPA certified laboratory. Chemical analyses were also conducted on site using Hach test kits for verification and calibration of instruments.

Water analysis was also conducted in the nearby CSD bottler's laboratory, for comparison. Analysis was also done at the American Water Chemicals facilities.

3.5.1 pH and Temperature

pH and temperature were directly measured using the installed pH analyzer (Signet 8750 ProcessPro pH Transmitter) with immersed probe (Signet 2754 pH probe). The immersed pH probe was calibrated with pH buffer kit (Signet PN 3-0700.390). The pH of the water samples were measured using portable pH meter (Hach SensION 1 Portable pH meter). The probe of the portable pH meter was calibrated using pH 4.01 and pH 10.0 buffer solutions (Hach PN#22834-49 and PN#22836-49).

3.5.2 Alkalinity

Alkalinity was measured using SM18 2320 B.

3.5.3 Calcium and Magnesium Hardness

The calcium and magnesium hardness were measured using EPA SW846-6010B Inductive Coupled Plasma – Atomic Emission Spectrometry- Total Recoverable.

3.5.4 Turbidity

Turbidity was measured using a portable turbidimeter (Hach Model 2100 Series) calibrated with <0.1, 1, 20, 100 and 800 NTU stabilized formazin standards (Hach Calibration kit PN#26594-05) and EPA Method 180.1. Turbidity was also measured directly from the HACH Model 1720D Low Range Process Turbidimeter installed in the pilot unit.

3.5.5 Total Suspended Solids

Total Suspended Solids was analyzed using EPA Method 160.2 . The suspended solids analyzed was the calcium carbonate precipitate in the UF reactor tank.

3.5.6 Total Organic Carbon

Total organic carbon was analyzed using EPA Method 415.1

Chapter Four

Results and Discussions

The pilot testing was conducted at Doosan Hydro Technology, Inc. facilities in Tampa, Florida. The pilot testing was divided into three phases. The first phase was to stabilize the flows and calibrate control valves and instruments. The first phase started on October 15, 2005, and was supposed to last one week, however it was extended by one more week, due to mechanical and instrument problems. The second phase was performed at varying flux and pH conditions. It started on October 29, 2005 and lasted four weeks. The objective of the third phase was to simulate the operation in a CSD Bottler Plant, based on the data obtained from the second phase. The third phase started on December 2, 2005 and ended on January 5, 2006. The source of feed water during the test was city of water supply.

4.1 Initial Operating Conditions Without Chemical Addition

The purpose of running the pilot unit at different flux levels, without the addition of chemicals, is to determine the flow characteristics of the unit and to calibrate the instruments. Based on the UF membrane area of 178 ft², the filtrate flow rates of 1.9, 3.7 and 5.6 gpm corresponded to flux values of approximately 15, 30 and 45 gfd. During the initial test run, it was noticed that display on the pressure indicator in the

suction line of the UF permeate pump was giving reading on increments of 0.5 psig and has to be replaced with a pressure indicator to provide reading down to 0.1 psig. pH and temperature were continuously displayed. The pressure measured on the UF membrane filtrate discharge and suction of the UF permeate pump pipework is the trans-membrane pressure. The location of the pressure sensor was in the same level as the water level in the UF reactor tank as recommended by the membrane manufacturer. This eliminated the need for correcting for the difference in hydraulic heads. The vacuum pressure reading can be considered as the trans-membrane pressure. Controlling the filtrate flow with the manual ball valve at the discharge of the UF pump was difficult, especially at lower flow, and it was replaced with a more accurate globe valve. After the flow and pressure readings were stabilized, the pilot unit was operated with varying flows of 1.9 to 5.6 gpm. The back flushing was set every 15 minutes for duration of 30 seconds. It was expected that the TMP will increase prior to back flushing. During the initial run at 1.9 gpm, the TMP remained at -0.5 psi, before and after back flushing throughout the 2 days of operation. At the flow of 3.7 gpm, the TMP stayed consistently at -1.1 psi after back flushing, and the pressure before back flushing was -1.5 psi. When operating at 5.6 gpm, the TMP after back flushing was -1.7 psi and increased to -2.0 psi before back flushing. Water samples were taken for analysis. Raw water analysis is shown in Table 4. The average pH of the feed water is 7.3 and the water temperature ranges from 20 to 25°C. Chlorine was not dosed during back flushing and during PFE.

Table 4
Raw Water Analysis

	10/15/05	11/12/05	12/10/05
pH	7.31	7.5	7.3
Alkalinity, mg/L CaCO ₃	76	70	75
TOC, mg/L	3.8	4.0	3.6
Ca, mg/L CaCO ₃	65	60	62
Mg, mg/L CaCO ₃	4.2	4.5	4.2
Turbidity, NTU	0.1	0.1	0.1

4.2 Operation at Varying Flux and pH

The second phase of the pilot testing was the addition of lime to achieve operating pH values of 8.3, 9.4, 10.6 and 11.2, at flows of 1.9, 3.7 and 5.6 gpm (or flux of 15, 30 and 45 gfd). The pilot unit was operated continuously for 2 days for each flow condition. The pH was set to the desired operating pH and the chemical feed pump automatically dosed the required lime solution. The average TMP values before and after back flushing are shown in the Table 5. The flux and permeability values at different operating conditions are shown in Table 6. The Permeability Profile at various operating conditions is shown in Figure 8. The permeability values range from 50% to 85% of the clean water permeability for SpiraSep UF membrane, which is 35 gfd/ psi. Figure 9 shows the TMP profile during the test. It can be observed, TMPs tends to increase with increasing flow (or flux) and operating pH.

Table 5
Average Vacuum Pressures or TMP Values in psi Before and After
UF Back Flushing at Various Flux Values

pH	Flux Values		
	15 gfd	30 gfd	15 gfd
7.3	-0.4/-0.5 psi	-1.1/-1.3 psi	-1.8/-2.0 psi
8.3	-0.6/-0.8 psi	-1.2/-1.5 psi	-1.9/-2.2 psi
9.4	-0.6/-0.8 psi	-1.3/-1.8 psi	-2.2/-2.6 psi
10.6	-0.7/-1.0 psi	-1.6/-2.2 psi	-2.4/-2.8 psi
11.2	-0.6/-1.0 psi	-1.7/-2.3 psi	-2.6/-3.4 psi

Note: After BF/Before BF

Table 6
Flux vs. Permeability at Various Operating pH

pH	Flux		
	15 gfd	30 gfd	45 gfd
7.3	29.6	26.29	26.12
8.3	25.81	25.31	23.95
9.4	25.14	23.03	20.43
10.6	23.0	19.32	19.21
11.2	21.4	19.05	18.17

Note: Permeability is gfd/psi

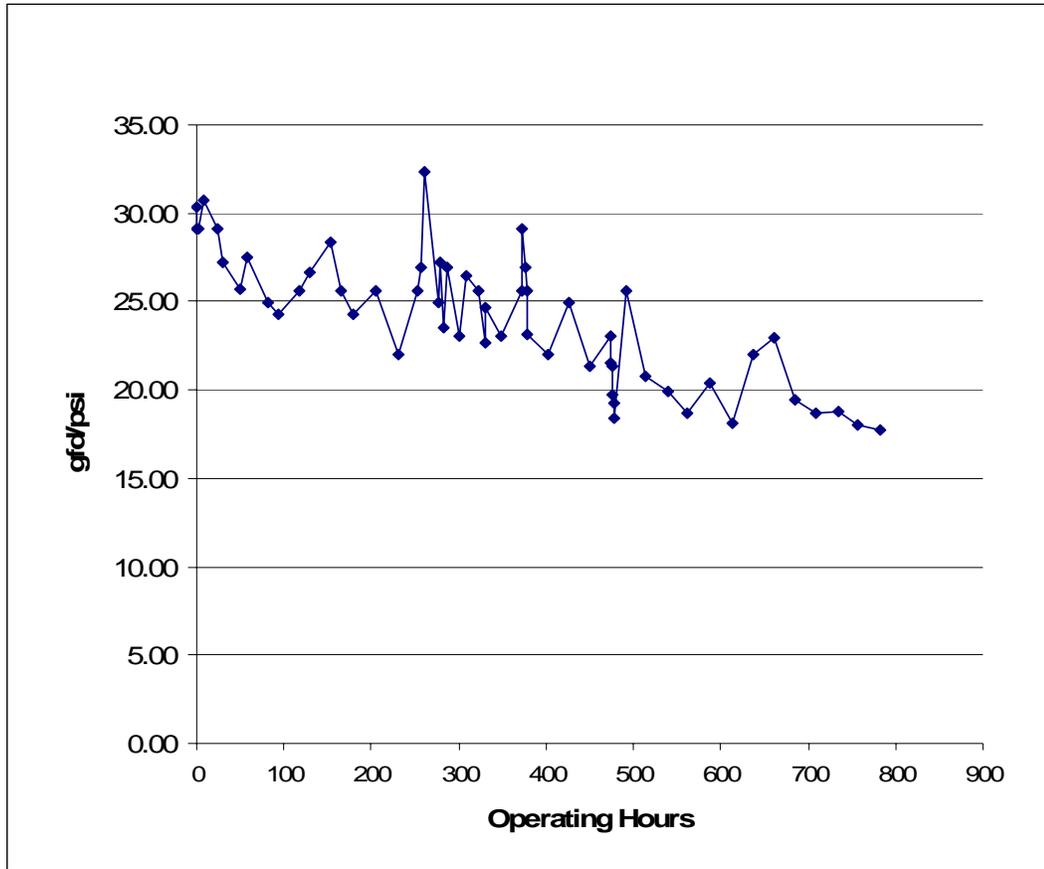


Figure 8 Permeability Profile at Various Operating Conditions

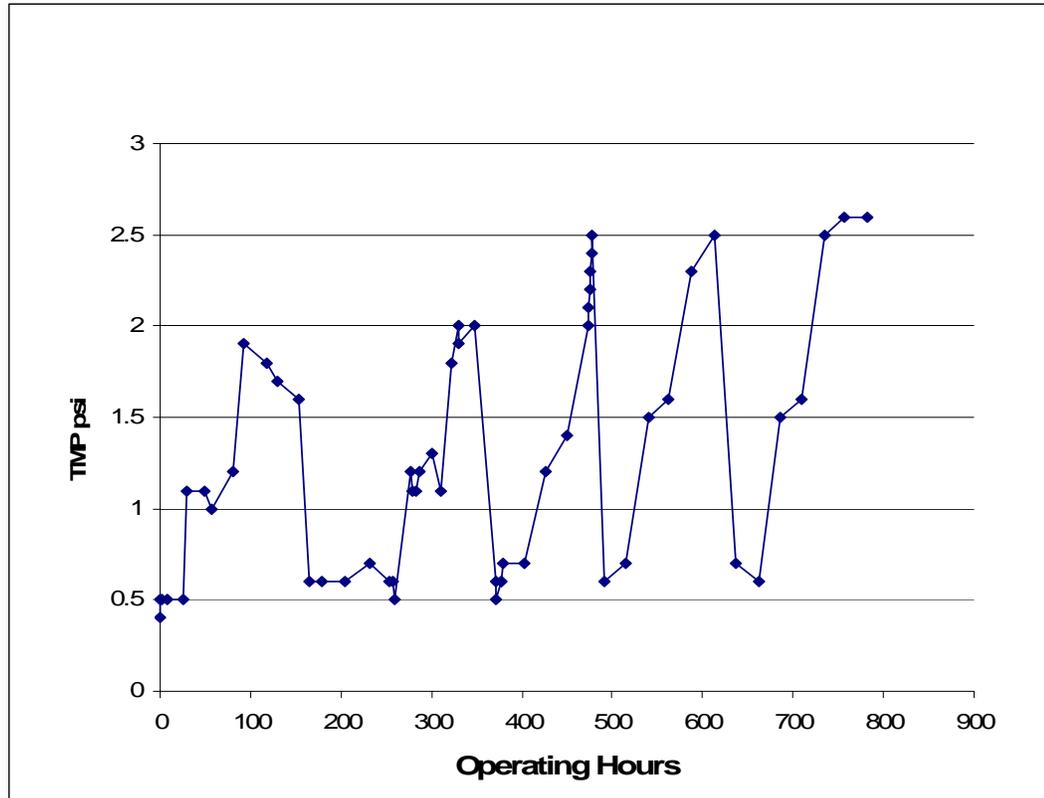


Figure 9 TMP Profile at Various Operating Conditions

Composite raw water and filtered water samples were analyzed for Ca, Mg, Alkalinity, pH, turbidity, and TOC. Grab water sample from the membrane reactor was also taken for total suspended solids analysis. The results of the water analysis are shown in Table 7.

An analysis of water sample was also conducted by the CSD Bottler and shown in Table 8. Note that there is difference between the operating pH value and the pH of the Filtrate analyzed in the laboratory. The pH of the filtrate was expected to be lower due to the effect of aeration in the UF tank which tends to strip the CO₂ or add CO₂ from the air. Aeration has stabilizing effect on the filtrate. During the test the amount of lime in each run was not monitored, however every time a batch was prepared, the quantity was recorded.

Table 7
Analysis of Water Samples at Various Operating Conditions

Operating pH	7.3	7.3	8.3	9.4	10.6	11.2
pH of the Sample	7.31	7.31	8.06	9.2	10.3	10.8
Type of Water	Raw	Filtrate	Filtrate	Filtrate	Filtrate	Filtrate
Alkalinity, mg/L CaCO ₃	76	76	62	30	36	36
TOC, mg/L	3.8	3.8	3.7	3.5	3.3	3.2
Ca, mg/L as CaCO ₃	65	65	57	38	41	56
Mg, mg/L as CaCO ₃	4.6	4.6	4.5	4.3	3.9	2.4
Turbidity NTU	0.1	0.05	0.05	0.05	0.05	0.05

From the tables above, it can be noted that there is a significant reduction of alkalinity and hardness, whereas at pH 10.6, the alkalinity and hardness increased. At pH 10.6 and above, the increase in alkalinity and calcium was due to the lime addition. The magnesium concentration continues to drop as the pH went up as expected.

The dilute sludge that accumulates at the bottom of the membrane reactor tank is manually drained, when the unit is stopped. During backwashing, the water in the membrane reactor overflows to lime reactor tank. The concentration of the suspended solids in the membrane reactor is shown in Table 9. It was observed that there was slight change in the sludge concentration when operating pH changed as shown in Table 9.

The concentrated sludge that accumulated at the bottom of the lime reactor tank was pumped out using another rotary flexible impeller pump rated at 0.25 gpm. Usually, 1/3 of the sludge in the conical section of the lime reactor tank was drained when the volume of sludge reaches the top of the conical section.

4.3 Operation at CSD Bottler Plant Conditions

The next phase of the test was to simulate the operation in an actual CSD bottler plant condition. Operation at pH 9.4 to 9.8 was chosen because the results in the previous tests satisfied the water quality requirement of the CSD bottler using the same source water as used in this test, although their actual operating pH was slightly higher. The resulting alkalinity level was favorable to their operation. The flux selection of 30 gfd (or flow of 3.7 gpm) was based on the following factors: economics, competing UF membrane's operating flux, test results from the second phase of the test, and guideline of the membrane manufacturer. The test also predicted the intervals between cleaning and estimated the consumption of lime. Water samples were taken and analyzed. The amount of lime used was also monitored. The test lasted for over 30 days. Figure 10 shows the permeability profile and Figure 11 shows the TMP profile throughout the duration of the test period. On the 18th day of test the TMP has almost doubled and the permeability dropped to down to 50% from the first day value. Based on experience, when this condition occurs, it is necessary to chemically clean the UF membrane. The cleaning was made as per the CIP procedure described Section 3.2.3. After cleaning the TMP and permeability values were

restored to the first day values. The operation of the pilot unit was continued for another 10 days after cleaning. The TMP and permeability profile after cleaning is similar to the initial profile. The analysis of the filtrate by STLI and the CSD bottler are shown in Tables 7 and 8.

Table 8
Analysis of the Filtrate by CSD Bottler

Operating pH	7.3	9.45	9.6	9.8
pH (Lab)	7.3	8.49	9.65	9.14
Phenolphthalei n Alkalinity, mg/L CaCO ₃	4.2	9.6	23.3	14.8
Methyl Orange Alkalinity, mg/L CaCO ₃	88.1	35.8	33.7	26.8

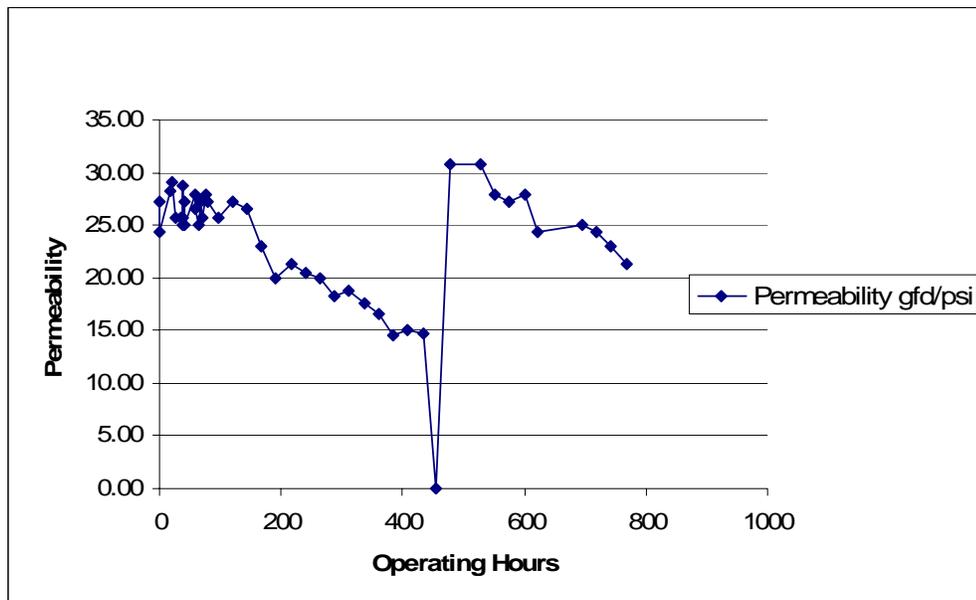


Figure 10 Permeability Profile at CSD Bottler Operating Conditions

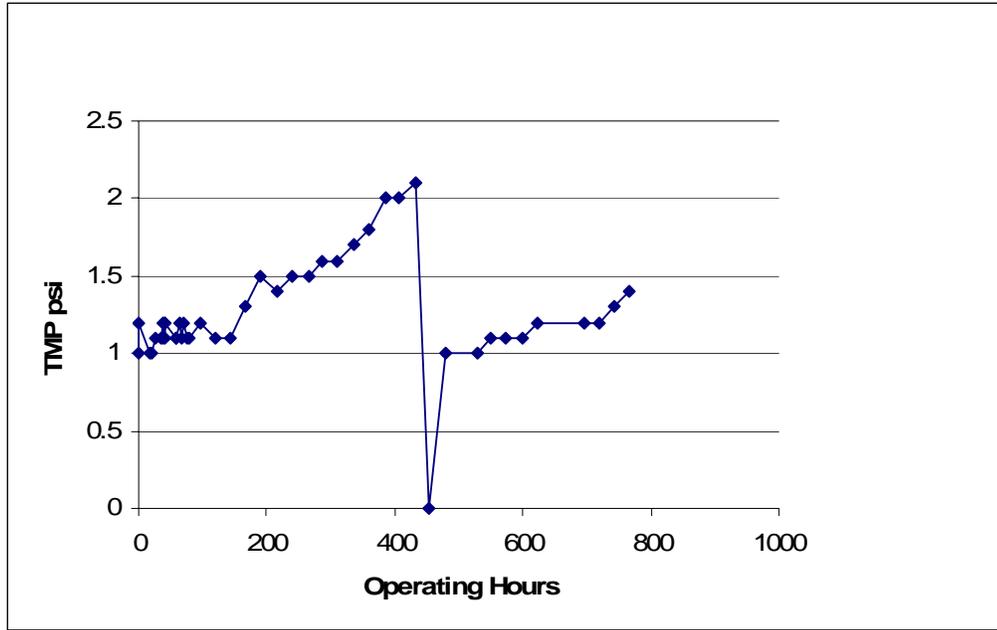


Figure 11 TMP Profile at CSD Bottler Operating Conditions

The concentration of suspended solids in the membrane reactor tank was maintained at 600 to 700 mg/L range. Backflushing seemed to maintain constant solids concentration in the membrane reactor. During backflushing, the excess water flowed back to the lime reactor tank, carrying suspended solids, and the backwash water diluted the water in membrane reactor. The sludge from the membrane and lime reactor tanks were drained as described in Section 4.2.

Table 9
Average Suspended Solids Concentrations in the Membrane Reactor

Operating pH	7.3	8.3	9.4	10.6	11.2
Suspended Solids conc., mg/L	10	580	600	600	680

Chapter Five Summary and Conclusions

5.1 Alkalinity Reduction

Alkalinity reduction to less than 50 mg/L or to the preferred level of 20 to 30 mg/L and maintenance of the desired Phenolphthalein Alkalinity and Methyl Orange Alkalinity ($2 \times \text{P alk} - \text{MO alk} = 2 \text{ to } 7$) can be achieved continuously in the lime softening UF unit with relatively simpler control, operation and maintenance compared to conventional lime softening process. The lime softening UF unit can be started in a matter of minutes, unlike the conventional lime softening which requires hours or days to build up of the sludge blanket before stable operation is achieved. The lime dosage during the third phase of test (operating pH=9.8) was 70 mg/L, based on raw water alkalinity concentration of 76 mg/L and pH of 7.3 and the filtrate alkalinity and pH are 26.8 mg/L and 9.18 respectively. The theoretical or calculated dosage using the Rothberg, Tamburini, and Windsor model was 65 mg/L. The lime dosage of the CSD bottler was in the range of 120 to 130 mg/L operating at pH of 9.8 to 10.2 with ferric chloride addition.

5.2 UF Filtrate Turbidity

The turbidity of the filtrate was consistently observed to be in the range of 0.04 to 0.05 NTU throughout the duration of the test. The filtrate turbidity was not affected by

the incoming feed water turbidity. When the pilot unit was operated without the lime addition, the feed water and filtrate turbidity were 0.1 NTU and 0.05 NTU, respectively. The suspended solids concentration in the membrane reactor tank throughout the test was in the range of 580 to 650 mg/L. Table 9 shows the average suspended solids concentration in the membrane reactor.

5.3 Trans-membrane Pressure (TMP) vs. pH and Flux

The increase in flux results to corresponding increase in TMP, however as the operating pH increases, the rate of TMP increases as shown in Figures 13 and 15.

5.4 Permeability

The operating the pH vs. permeability profile shown in Figure12 indicates, the permeability decreases with increasing operating pH. The TMP vs. flux profile shown in Figure 14 , indicate permeability decrease with increasing flux. The decline in permeability during the second phase of the test was due to the increase in operating pH. The starting and ending average permeability values were 31.25 gfd/psi and 17.53 gfd/psi. The prolonged operation without CIP had not impacted the permeability, because when the third phase of the test started, the starting average permeability during the first 2 days of operation was 26.93 gfd/psi, which is comparable to 25.5 gfd/psi when the operation started in second phase of the test at pH 9.4.

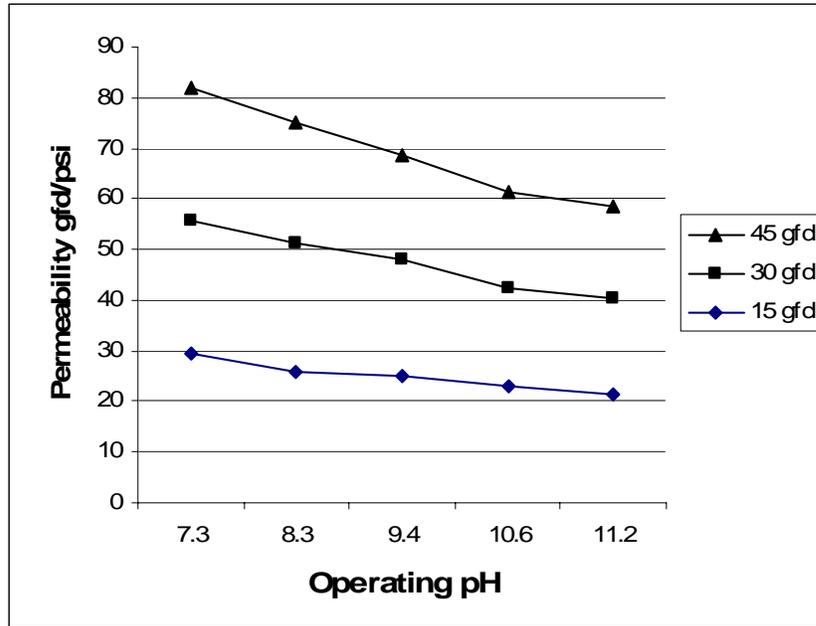


Figure 12 Permeability vs. Operating pH at Various Flux Rates

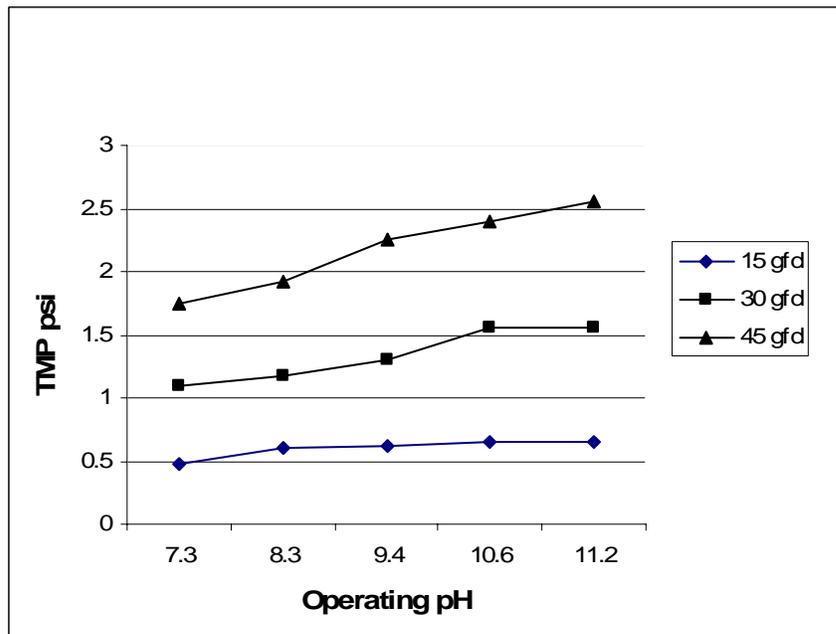


Figure 13 TMP vs. Operating pH at Various Flux Rates

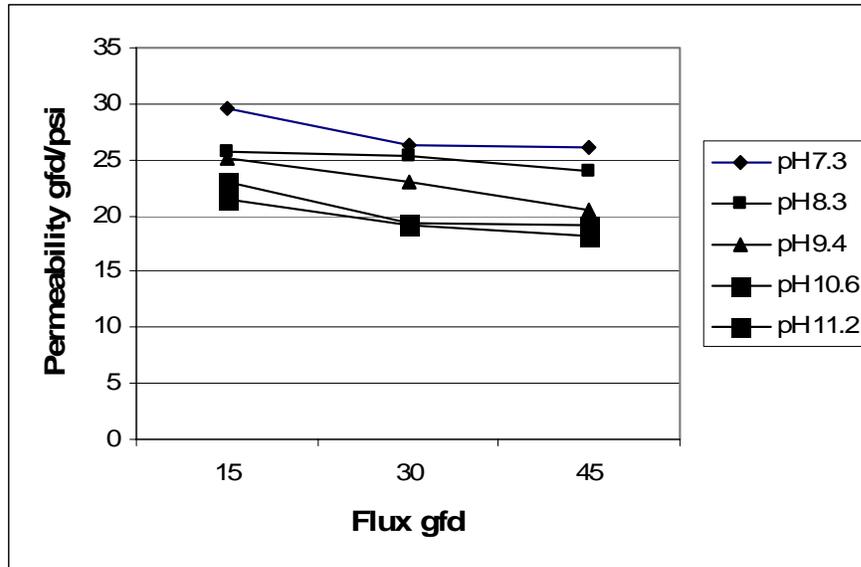


Figure 14 Permeability vs. Flux at Various Operating pH

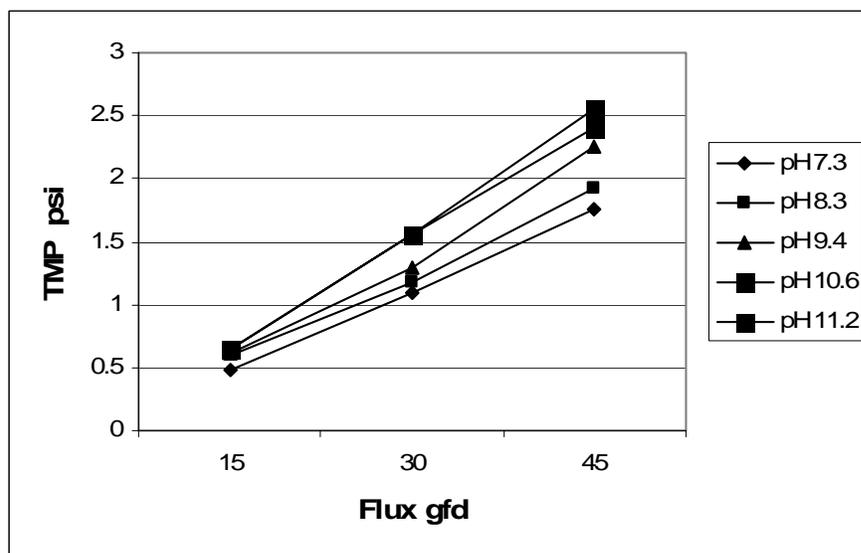


Figure 15 TMP vs. Flux Various Operating pH

5.5 Total Organic Carbon (TOC)

The data in Table 7 indicate that there was no reduction in TOC when the pilot unit was operated without lime addition. With the addition of lime, there was a slight reduction of TOC. The reduction in TOC ranged from 2.6% to 15.8%, when the pilot unit was operated at various pH values.

5.6 Hardness Reduction

Table 7 indicates the reduction in Ca and Mg hardness which was expected as a result of the increase in operating pH. The reduction of hardness is secondary concern in CSD bottling operations. It is assumed that alkalinity reduction will reduce hardness.

5.7 Operating Flux

The operating flux of 30 gfd was initially selected because most of the ultrafiltration membranes used in treating municipal operate at this flux value, although Trisep recommendation is 25 gfd for treating municipal water supply, when dosing coagulants (such as ferric chloride or sulfate, alum and polyaluminum chloride). It was assumed that lime will behave like the other coagulants although there were concerns of excessive fouling and scaling. The results of this research confirmed that the immersed SpiraSep UF membrane can achieve the treatment objectives when operated at flux of 30 gfd, and fed with lime treated water at pH 9.8, with suspended solids concentration of 600 mg/L. The cleaning of the membrane or CIP was initiated when the TMP value was doubled, which correspond to about 50% of clean membrane permeability. The CIP was conducted after 19 days of operation, noting that the pilot unit has been in operation for over 30 days in the first and second phases before the third phase started. The third phase of the test also confirmed the following: the cleaning procedures and chemicals mentioned in Section 3.2.3 effectively restored the membrane to its starting TMP and permeability; by extrapolating the permeability and TMP profiles the expected next cleaning will be after 48 days. This corresponds to 30 days cleaning interval.

5.8 Chlorination

During the entire duration of test, chlorine was not added to the back flush water or in the PFE. The residual chlorine in the feed water ranged from 0.2 to 0.7 mg/L. Chlorine was dosed only during CIP and when the unit was stopped longer than 24 hours.

5.9 Benefits of the Lime Softening Ultrafiltration (LSUF) Process to CSD Bottler

The benefits of the Lime Softening Ultrafiltration Process to CSD bottler, based on the results of this study can be summarized in the following:

- There is considerable economic benefit when the conventional treatment processes comprising of chlorination, lime softening, clarification, and filtration, is replaced with LSUF comprising of a single equipment with smaller footprint. With less equipment, operation and maintenance will be simpler.
- The LSUF process requires shorter time for start-up, unlike conventional lime softening which requires time to build up sludge, stabilize the flow and attain the desired treated water quality.
- The LSUF process produces less sludge and dirty backwash water. It can be operated at relatively lower pH and with no addition of ferric chloride which significantly reduced the volume of sludge. The water during backflush operation can be returned back to the system. The water wasted is the water that goes with the waste sludge, which is minimal.

- Continuous chlorination of raw water can be eliminated, reducing the formation of the THMs.
- Process control in LSUF reduced to adjustment of pH and flows.
The process is less sensitive to temperature.
- In LSUF process, the sludge removal is simplified because there is no sludge blanket to maintain.
- The ultrafiltration process provides physical barrier for microorganism and particles, minimizing the contamination in the down stream processes.
- Existing lime softening plants can be retrofitted and their rated capacity can be increased with just the addition of the UF system processes.

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Appendices

Appendix A Pilot Unit Equipment Description

1.0 Pilot Plant Systems Parameters

The pilot plant consist of the lime reactor and the UF system. The lime reactor was designed to suit the requirement of this research. The UF system is a full scale commercial unit with one (1) UF element.

1.1 Lime Reactor

Retention time:	30 minutes
Flash Mixing Chamber Retention:	30 – 60 seconds

1.2 Pilot Plant Process Flows

Plant Capacity (Effluent):	0 – 7 gpm
Membrane Flux Range:	0 – 60 gfd
System Recovery:	90%

1.3 Aeration

Aeration Flow Rate:	6.2 scfm
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1.4 Membrane Backwash

Frequency:	15 minutes
Duration:	30 seconds
Back Flush Flow Rate:	7.5 gpm
Back Flush Water Volume Used per Backwash:	5 – 6 gallons
Back Flush NaOCl Dosage Concentration:	10 mg/L

1.5 Periodic Flux Enhancement (PFE)

PFE Back Flush Flow Rate:	1.8 gpm
PFE Water Volume Used per PFE:	35 gallons
NaOCl PFE Frequency:	24 – 48 hours
Citric Acid PFE Frequency:	3 days
PFE Back Flush Length:	10minutes
PFE Static Soak Length:	10 minutes

Appendix A: (Continued)

NaOCl PFE Dosage Concentration: 100 mg/L
Citric Acid PFE Dosage Concentration: 0.1%

1.6 Clean-In-Place (CIP)

CIP Backwash Flow Rate: 1.8 gpm
CIP Tank Volume per Manifold: 75 gallons
NaOCl CIP Cleaning Frequency: 3 months
CIP Duration: 4 – 8 hours
NaOCl CIP Dosage Concentration: 2,000 mg/L
NaOCl CIP Concentration: 0.1%
Citric Acid CIP Cleaning Frequency: 3 months
Citric Acid CIP Concentration: 1.0%

2.0 Equipment Specifications

2.1 Lime Reactor

Type: Cylindrical with conical bottom
Capacity: 200 gallons
Materials of Construction: PE
Mixer: 1/20 hp

2.2 Ultrafiltration Membrane

Model: SpiraSep 900
Chemistry: PES
Quantity: One (1)
Element Diameter: 9.38 inches
Element Length: 42 inches

2.3 Aeration Manifold

Material: Schedule 40 PVC
Size: 1 inch Schedule 40 PVC

Appendix A: (Continued)

2.4 Membrane Tank

Quantity: One (1)
Material: PVC
Height: 60 inches

Water Level: 54 inches
Diameter: 18 inches
Effective Volume: 60 gallons

2.5 Filtrate Storage Tank

Quantity: One (1)
Material: PE
Volume: 75 gallons

2.6 Filtrate Pump

Quantity: One (1)
Pump Type: Self-priming centrifugal
Model: Flotec FP5162
Construction: Noryl wetted parts
Process Piping: SCH 40 PVC
Control: Manual Throttle Valve
Capacity: 10 gpm @ 20 feet suction lift
Pump Power: 0.75 hp
Power: 115/230 VAC, 1 phase, 60Hz

2.7 Backwash Pump

Quantity: One (1)
Pump Type: Centrifugal
Model: American Stainless SSPC1
Construction: 316 SS wetted parts
Process Piping: SCH 80 PVC, 316
Control: VFD
Capacity: 10 gpm @ 10.0 psi discharge
Pump Power: 1.0 hp
Power: 230/460 VAC, 3 phase, 60Hz

Appendix A: (Continued)

2.8 Blower

Quantity:	One (1)
Quantity per Train:	One (1)
Blower Type:	Regenerative (oil-less)
Model:	Ghast or equivalent
Construction:	Carbon Steel
Process Piping:	Galvanized Steel
Control:	Manual Throttle Valve
Capacity:	10 scfm @ 2.5 psi discharge pressure
Blower Power:	1.0 hp
Power:	230/460 VAC, 3 phase, 60Hz

2.9 Low Capacity Chlorine Metering Pump

Quantity:	One (1)
Pump Type:	Positive Displacement
Model:	LMI
Wetted Ends:	Polypropylene with PVC
Diaphragm:	PTFE
Balls:	Ceramic
Capacity:	0.2 gpd
Controller:	Manual
Power:	115/230 VAC, 1 phase, 60Hz

2.10 High Capacity Chlorine Metering Pump

Quantity:	One (1)
Pump Type:	Positive Displacement
Model:	LMI
Wetted Ends:	Polypropylene and PVC
Diaphragm:	PTFE
Balls:	Ceramic
Capacity:	14 gpd
Controller:	Manual
Power:	115/230 VAC, 1 phase, 60Hz

Appendix A: (Continued)

2.11 Sodium Hydroxide Metering Pump

Quantity:	One (1)
Pump Type:	Positive Displacement
Model:	LMI
Wetted Ends:	Polypropylene and PVC
Diaphragm:	PTFE
Balls:	Ceramic
Capacity:	10 gpd
Controller:	Manual
Power:	115/230 VAC, 1 phase, 60Hz

2.12 Citric Acid Metering Pump

Quantity:	One (1)
Pump Type:	Positive Displacement
Model:	LMI
Wetted Ends:	Polypropylene and PVC
Diaphragm:	PTFE
Balls:	Ceramic
Capacity:	10 gpd
Controller:	Manual
Power:	115/230 VAC, 1 phase, 60Hz

2.13 PLC/Control Panel

Quantity:	One (1)
Model:	Automation Direct DL 06
Power Input:	230/460 VAC, 3 phase, 60 Hz
Enclosure:	NEMA 12, Carbon Steel
Operator Interface:	LCD with push buttons

2.13 Instrumentation

Level Switches:	4, NOC Float Switches
Pressure Gauges:	2, Ashcroft or equivalent
Quantity Rotameters:	4, Blue-White or equivalent
Air Rotameters:	1, Blue-White or equivalent
Temperature Gauge:	1, Cole Palmer or equivalent

Appendix A: (Continued)

	pH Meter :	1, SIGNET
	Turbidimeter	1, HACH
2.14	Piping and Automated Valves	
	Automated Ball Valves:	Four (4)
	Manual Globe Valve:	Four (4)
	Manual Ball Valves:	Four (4)
	Piping Material:	SCH 40 PVC

Appendix B

SpiraSep Trans-membrane Pressure (TMP) Measurements

1.0 TMP Measurement

The TMP of the SpiraSep system can be calculated by the following equation:

$$\text{TMP} = P_{\text{vac}} + H_{\text{tank}} - H_{\text{ct}} \quad (1.1)$$

where

$$P_{\text{vac}} = \text{vacuum pressure}$$

$$H_{\text{tank}} = \text{hydrostatic pressure in tank}$$

$$H_{\text{ct}} = \text{hydrostatic pressure in membrane core tube}$$

Since $H_{\text{tank}} = H_{\text{ct}}$, the TMP is equal to the vacuum pressure. Equation (1.1) now becomes:

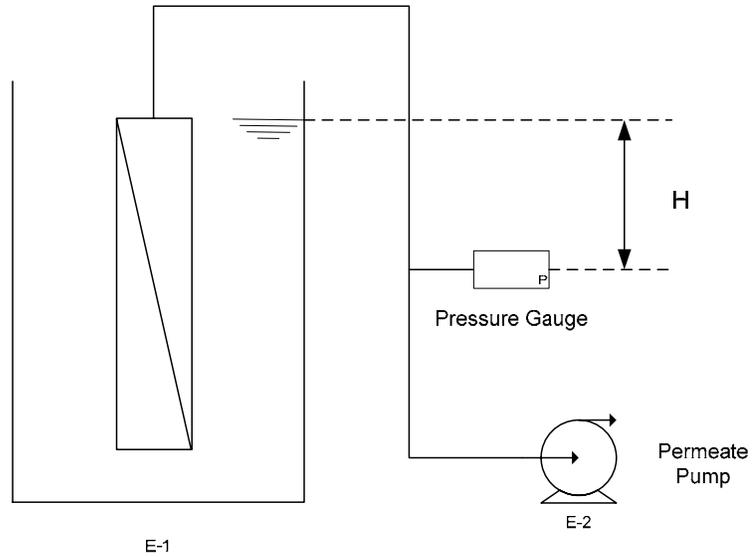
$$\text{TMP} = P_{\text{vac}} \quad (1.2)$$

2.0 Pressure Gauge Location

The height of the pressure gauge location should even with the water level inside the membrane tank, as this will indicate the true trans-membrane pressure. It is important to account for any hydrostatic pressure losses/gains in the suction pipe when measuring TMP. Although the hydrostatic pressures inside the membrane tank and element core tube cancel each other out, the hydrostatic pressures in the suction line leaving the element must be accounted for. Below are several different scenarios on TMP measurement based on gauge/sensor location.

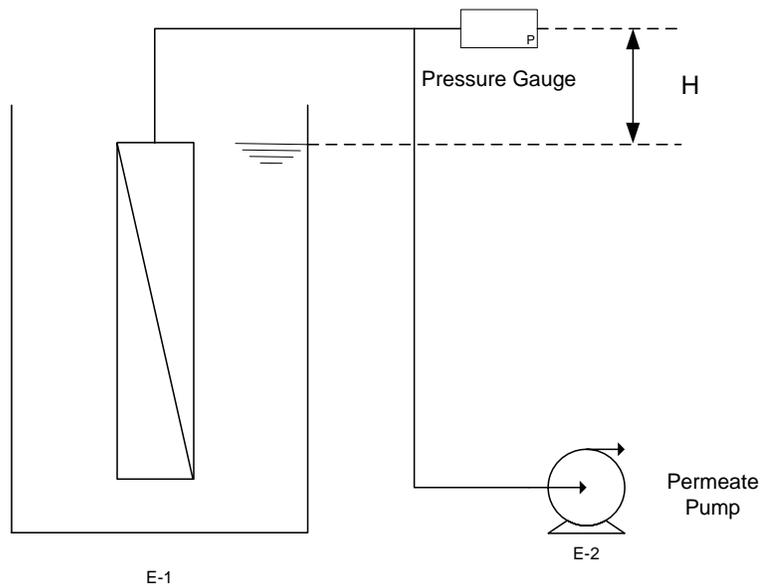
Appendix B: (Continued)

Scenario 1



$TMP = P_{vac} - H$, where P_{vac} is the pressure measured by the gauge/sensor. For example, if the pressure measured in scenario 1 by the pressure sensor is -1.5 psi and H is equal to 24 inches, then the TMP is equal to -2.37 psi (-1.5 minus 0.87).

Scenario 2



Appendix B: (Continued)

$TMP = P_{vac} + H$, where P_{vac} is the pressure measured by the gauge/sensor. For example, if the pressure measured by the pressure sensor is -2.0 psi and H is equal to 12 inches, then the TMP is -1.57 psi (-2.0 plus .43).