<u>Final Report</u>

Application of Ceramic Membrane Modules For Combined Sewer Overflow Management: Engineering Process Considerations

By

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Abstract

The purpose of bench scale research is to develop and understand microfiltration process parameters as applied to “dilute sewage” and to use this information as a guide for field scaled pilot plant activities. The extent and breadth of inquiry and number of independent variables that can be investigated is necessarily limited at the field scale. Bench scaled preliminary research is found valuable for exploratory purposes, and as a guide for subsequent larger scaled field activities.

This report is divided into four sections: the first having an introduction and general comments and literature review, the second showing results of bench-scaled research, the third focusing on field studies at ALCOSAN, and the fourth containing recommended next steps to take.

Application of cross flow microfiltration with and without backpulsing was evaluated at the bench scale for the treatment of dilute primary sewage effluent simulating combined sewer overflow wastewater. Four alpha alumina ceramic membranes of various pores sizes (0.2 - 5.0 µm) were tested to understand the impact of cross flow velocity and transmembrane pressure on the permeate water quality and flux rate. The 0.2 and 0.8 µm membranes produced a permeate water quality that is likely to be suitable for surface water discharge. The combination of permeate chemical and biological water quality and long-term flux rates suggest that a 0.2 µm membrane would be the most appropriate membrane for the treatment of combined sewer overflow wastewater within sewersheds.

A pilot scale investigation followed and built upon results from the bench work. The pilot work was undertaken at the Allegheny County Sanitary Authority (ALCOSAN) for approximately 12 months to evaluate the feasibility of using cross-flow microfiltration for the treatment of primary sewage effluent simulating combined and sanitary sewer overflows. Commercial sized ceramic membranes of various pores sizes (0.05 - 1.4 μm) were tested using ALCOSAN primary effluent wastewater to understand the impact of cross flow velocity, transmembrane pressure, and feed suspended solids on permeate water quality and permeate flux rate. A 0.2 μm membrane operated with a 1.8 m/s cross flow velocity, a transmembrane pressure below 2.1 bar and a back pulse frequency of 60 seconds showed the best performance among the conditions evaluated in this study. The 0.2 μm membrane consistently met water quality objectives for fecal coliforms, *E Coli*, E*nterococci*, BOD_5 and suspended solids independent of the feed concentration, suggesting that direct discharge to surface water may be feasible. Feed suspended solids concentration and temperature influenced membrane permeate flux. Membrane cleaning with alkaline sodium hypochlorite solution is recommended as the first step followed by nitric acid cleaning if needed.

***Keywords**:* Combined Sewer Overflow, Sanitary Sewer Overflow, Ceramic Membrane Filtration, Microfiltration, Primary Effluent, Pathogenic Indicator Organisms, BOD, Cross Flow Velocity, Transmembrane Pressure

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Section 1

Introduction

Combined sewer overflows (CSO) occur in 772 communities and sanitary sewer overflows (SSO) occur at least 40,000 times a year in the United States [US EPA, 2001a, 2001b]. These overflows contribute high levels of suspended solids, pathogenic microorganisms, oxygen-demanding compounds, and other pollutants into the receiving stream that pose risks to human health, to aquatic life and habitat, and to the use and enjoyment of waterways [US EPA, 1986, 2001a]. The temporal level of pollution coupled with regulatory pressure is challenging communities to find feasible management strategies and treatment alternatives [US EPA, 1999]. With the recent growing awareness of membrane filtration and the declining cost of membranes, the use of membrane microfiltration within the sewer shed may be a viable management and treatment option for many communities. Advantages of microfiltration membranes include the ability to reject bacteria, silts, cysts and spores, reduced land requirements compared to storage basins, and the potential for mobile treatment [Gan 1999, Till et al., 1998, Cardew & Le 1998].

CSO control is difficult due to the site-specific variability in the volume, frequency, and characteristics of the diluted sewage-stormwater matrix (US EPA, 2001). Some locations have CSOs with virtually every rainfall, while other locations have CSOs infrequently (US EPA, 2001). The rivers surrounding the greater Pittsburgh, PA region provide evidence for bacterial contamination following storm events. In one recent sampling survey during the summer of 2001, bacterial levels as high as 50,000 fecal coliforms cfu/100 mL, 39,000 Escherichia coli (E Coli) cfu/100mL, and 1,060 Enterococci cfu/100 mL were detected in Allegheny river (USGS, 2001).

The Pennsylvania Department of Environmental Protection regulates surface water quality in the Pittsburgh region under PA Code Title 25, Subpart C, Article II, Chapter 93.7a. The regulations require that the monthly geometric mean of fecal coliforms during the warm weather months of May through September should be less than 200 cfu/100 mL and no more than 10% of the samples can be greater than 400 cfu/100 mL. For the remainder of the year, the maximum fecal coliforms level should be less than a monthly geometric mean of 2,000 cfu/100 mL.

The United States Environmental Protection Agency (EPA) suggested that States adopt new bacterial standards based on the 1986 EPA's Ambient Water Quality Criteria for bacteria that go beyond fecal coliforms and additionally embrace a monthly geometric mean of 126 cfu/100 mL for E. Coli and 33 cfu/100 mL for Enterococci, with a monthly maximum dependent on the use of the water body (Slack et al., 2000, US EPA, 1986).

The level of pollution coupled with regulatory pressure is challenging communities to find a feasible treatment alternative for CSO discharges (US EPA, 1999b). Some reduction of bacteria can be achieved through solids removal by sedimentation, flotation and filtration with disinfection being the most common method (US EPA, 1999a). Chlorine disinfection may not be widely acceptable for combined sewer overflows due to residual disinfectant toxicity for the receiving waters, difficulty in regulating the addition of the disinfectant, wide variety of bacterial compositions and concentrations, and high suspended solids concentrations of the CSO wastewater (US EPA, 1999a). Chlorine alternatives such as ozonation, ultraviolet radiation, peracetic acid and electron beam irradiation are now being considered for the treatment of CSOs (US EPA, 1999a).

The total cost of the facilities in the United States that are required to provide combined sewer overflow treatment (*minimally consisting of primary sedimentation, chlorine disinfection, and dechlorination)* was estimated at $44.7 billion (1996 U.S. dollars) (US EPA, 1994). A capital cost of $1.1 million (1997 U.S dollars) and an annualized cost of $386 thousand (1997 U.S. dollars) for chlorine dioxide disinfection of CSO wastewater at a flow rate of 71 m3/s was estimated based on the pilot scale studies (US EPA, 1999a). The capital cost for ozone disinfection for the same CSO wastewater flow rate was estimated at $23 million (1997 U.S. dollars) with an annualized cost of $2.9 million (1997 U.S. dollars) (US EPA, 1999a). Accordingly, research leading to the development of other alternative technologies for future CSO application by municipalities is prudent and timely. Some advantages of membrane filtration include bacteria removals with no chemical requirement, reduced land requirements compared to storage basins and the potential for mobile treatment. (Gan 1999, Till et al., 1998).

Porous microfiltration membranes are used for liquid/solid separation to selectively remove undesired pollutants. The wastewater, referred herein as the feed, is driven through the membrane by an applied pressure. The treated water that passes through a membrane is referred to as permeate, while the water that is rejected by the membrane is referred to as the retentate. During the initial stages of operations while the membrane is still clean, the undesired particulate pollutants are rejected by the size of the membrane pores. After the initial stages of operation, the mechanism changes. The particles that accumulate near the membrane surface eventually form a cake layer or fouling layer, which assists in pollutant removal. The fouling layer decreases the permeate flow rate and is the key factor governing the design and application of microfiltration systems in environmental engineering practice.

The cross-flow configuration was selected for this study due to the ability to treat wastewater with high levels of particulates in a continuous fashion. During cross-flow microfiltration, the wastewater flows parallel to the membrane surface scraping away particles and thus reducing the impact of the fouling layer. In addition, a "backpulse" was consistently used in this study to further reduce the adverse impact of high solids concentration in the feed on permeate flux. Backpulsing is the short duration, periodic redirection of water flow from the permeate side of the membrane to the feed side of the membrane accomplished by supplying a greater pressure on the permeate side of the membrane. The redirected water flows through the pores fracturing the fouling layer. This results in particles being carried away from the membrane surface with the retentate. Recent research using ceramic cross-flow microfiltration systems studying the impact of backpulsing on a primary sewage effluent demonstrated a higher long term permeate flow rate [Gan, 1999, Sondhi 2000, Bendick (2003)].

Key operational and design parameters for cross-flow microfiltration systems include the membrane pore size, permeate flux rate, cross flow velocity, transmembrane pressure and operating temperature. The membrane pores represent a barrier for pollutants and a smaller pore size may be expected to provide better permeate quality. However, smaller pore sizes will reduce the permeate flow rate at a given transmembrane pressure. The permeate flow rate is also impacted by the cross-flow velocity, transmembrane pressure and operating temperature.

An investigation using cross-flow ceramic membranes with a mean pore size between 0.22 – 1.3 μm for the treatment of a primary sewage effluent demonstrated large initial (clean membrane) variations in the permeate flow rate, with the variations diminishing significantly as fouling layer is established during normal operation [Gan, 1999]. The similar steady state flux rates even at larger pore sizes were attributed to mass transfer limitations both at the membrane surface, and within the pore structure [Gan, 1999]. For example, investigation of cross flow microfiltration for beer clarification demonstrated greater permeate flux rate for a 0.5 μm membrane than for a 1.3 μm membrane [Gan et al, 1997]. As with the prior study, the major reason for the larger pore size providing for lower permeate flow rates was attributed to in-pore fouling [Gan et al, 1997]. These results suggest that the particles that accumulate on the membrane surface and within the membrane may be more critical to the permeate flow rate than the membrane pore size.

Ultimately, the permeate flow rate determines the required membrane surface area for a given installation and design flow rate. The flux is defined as the permeate flow rate per unit surface area of the membrane and is calculated as follows:

$$J = \frac{Q_p}{A_S} \tag{1}$$

where J = flux (L/hr-m^2), Q_p = permeate flow rate (L/hr), A_s = membrane surface area available for filtration (m^2).

During cross flow microfiltration, the permeate flux rate is initially very high. As the fouling layer develops, a rapid decrease in the observed flux rate occurs followed by a more gradual decrease towards a relatively constant flux rate. The constant flux rate, referred to as the steady state flux rate, should be used for process design.

The cross flow velocity is the rate at which the feed water flows across the membrane surface and is calculated as follows:

$$V = \frac{Q_b}{A_c} \tag{2}$$

where V = cross flow velocity (m/s), Q_b = feed flow rate within a membrane module (m^3/s) and A_c = cross sectional area available for flow (m^2).

The cross flow velocity is a critical design parameter for cross-flow microfiltration systems. The selection of a preferred velocity depends on the trade off between an improved flux rate and an increase in pumping costs. Previous studies demonstrated a 15% flux improvement with an increase in cross-flow velocity from 2 to 6 m/s during microfiltration of a primary sewage effluent [Gan and Allen, 1999]. Similarly, cross-flow microfiltration of a primary and secondary effluent at velocities from 0.9 to 5.7 m/s demonstrated an improved flux rate [Judd et al, 2000].

In cross-flow ceramic microfiltration, the untreated wastewater flows through a tube or a channel with interior walls coated with an inert substance producing a membrane surface with a well defined pore structure. The micro porous membrane is conceptually similar in function to a traditional filter, except that the defined pores are designed to be in the range of 0.01 to 10 μm. (Baker, 2000). As an initial approximation, Darcy's Law can model the flow through a membrane as a porous media, with flux is proportional to the pressure drop across the membrane and inversely proportional to the resistance (Al-Malack et al., 1997). The permeate flux through a membrane can be expressed by the following equation:

$$J = \frac{\Delta P}{(\eta * R_t)} \tag{3}$$

where, J is the flux (m/s), ΔP is the transmembrane pressure (N/m^2), η is the viscosity of the feed water ($N/s\text{-}m^2$), and R_t is the total resistance to flow (1/m).

The flux is the flow of permeate water per unit surface area of the membrane and is a function of the physical characteristics of the membrane, the mode of operation and the operating conditions.

The transmembrane pressure is the driving force for membrane filtration. The transmembrane pressure is the difference in pressure from the feed side of the membrane to the permeate side of a membrane. The transmembrane pressure, ΔP, for a cross flow membrane is calculated as:

$$\Delta P=[(P_i+P_o)/2]-P_p \tag{4}$$

where, ΔP = transmembrane pressure (bar), P_i = inlet pressure to the membrane module (bar), P_o= outlet pressure from the membrane module (bar) and P_p = permeate pressure (bar).

The transmembrane pressure is also an important design parameter for cross flow microfiltration systems as an increase in transmembrane pressure is expected to increase the flux rate.

The total resistance to flow (R_t) of the membrane includes the resistance of the membrane itself (R_m) and the resistance of the fouling layer (R_f), and can be expressed as:

$$R_t = R_m + R_f \tag{5}$$

The intrinsic resistance of the membrane (R_m) is a specific physical property that remains constant over time. The resistance of the fouling layer (R_f) is negligible for a clean membrane and will increase with time as particles accumulate inside the membrane and on the membrane surface. Using clean water (0.2 μm filtered deionized water) and the flux at the beginning of a laboratory filtration experiment, J_o, the resistance of the membrane can be calculated as:

$$R_m = \frac{\Delta P}{(\eta * J_o)} \tag{6}$$

The resistance of the fouling layer, which causes a decrease in permeate flux during wastewater filtration can be calculated as:

$$R_f = \frac{\Delta P}{(\eta * J)} - R_m \tag{7}$$

The selection of a preferred transmembrane pressure depends on the trade off between an improved flux rate and an increase in pumping costs. At very low transmembrane pressures, the impact of the fouling layer is not as significant because the particles can be swept away by the feed water. As the transmembrane pressure increases, the particles are compressed on the membrane surface and the effect of the fouling layer on transport resistance increases, thereby limiting the permeate flux rate. At high transmembrane pressures, the particles may become so concentrated at the membrane surface that a gel layer that forms becomes the significant barrier in controlling permeate flux rates.

An investigation into the cross-flow microfiltration of primary sewage effluent demonstrated a 41% steady state flux increase for an increase in transmembrane pressure from 0.5 to 2.5 bar [Gan and Allen, 1999]. However, the flux increase with increase in transmembrane pressure from 0.5 to 1 bar was 25%, while the last incremental increase in pressure from 2.0 to 2.5 bar improved the permeate flux by less than 1%. This lack of commensurate increase in permeate flux with an increase in pressure demonstrates the concept of a limiting flux at elevated transmembrane pressures.

The viscosity of the fluid decreases with an increase in temperature. Therefore, as suggested by Darcy's Law, an increase in temperature will create higher permeate flux rates. For this study, the standard viscosity correction factor shown below [Lorch, 1987] was used to compare flux rates obtained at different temperatures.

$$J_{20} = J_T \times 1.03^{(20-T)} \tag{8}$$

where J_{20} = permeate flux at 20^{o}C (L/hr-m^2), J_T = permeate flux at actual temperature (L/hr-m^2) and T = actual temperature (^{o}C).

Section 2

Bench Scale Cross-Flow Microfiltration of Combined Sewer Overflow

MATERIALS & METHODS-BENCH UNIT

Objective

The experimental investigation was conducted to determine the membrane pore size capable of reducing pathogenic indicator bacteria to negligible levels, to better understand the relationship between the fouling layer and the permeate flux rates and water quality, to verify the effect of backpulse on permeate flux and water quality, and to determine the impact of the pore size on membrane performance during extended operation consistent with combined sewer overflow event duration (typically less than three days). In addition, information gathered at the bench scale is used to guide the experimental design for the subsequent larger field- scale pilot unit.

Experimental Approach

A bench-scale cross-flow ceramic membrane filtration unit, Membralox® Model 1T1-70 was used at the University of Pittsburgh. The apparatus (Figure 1) consists of a ¾ HP centrifugal pump, a 15-L feed tank, an in-line flow meter, a ceramic test module housing one membrane, a temperature gauge, an automatic backpulse device, eight process control ball valves, and three pressure gauges to monitor the inlet, outlet and permeate pressure. The filtration system can achieve a cross flow velocity of 1.6 m/s to 8.2 m/s and an inlet pressure to the membrane housing of 0 bar to 3.8 bar. The backpulse device uses 5.5 to 8.3 bar of oil-free, dried, filtered nitrogen gas. The device sends pressurized nitrogen gas to a piston on the permeate side of the membrane where a small amount of clean permeate water is stored in a reservoir. The piston compression creates higher pressure on the permeate side of the membrane, thereby redirecting the flow from the permeate side of the membrane to the feed side of the membrane. The frequency and duration of the backpulse can be controlled independently. The experiments were performed with a transmembrane pressure between 1.0 and 1.8 bar, a cross flow velocity of 6.56 m/s, a backpulse frequency of 60-94 seconds and a backpulse duration of 0.5 seconds.

Membranes

The four bench scale Membralox® T1-70 alpha alumina membranes with mean pore size of 0.2, 0.8, 2.0 and 5.0 μm were evaluated. Each tubular membrane is 250 mm in length, 7 mm in diameter and has 55 cm^2 of available surface area. The membranes are capable of withstanding a pressure limit of 7.9 bar, a temperature limit of 225 ^oC and a pH range of 0-14. The membrane influent and permeate were analyzed for bacteriological parameters and for chemical oxygen demand (COD) at various times during filtration tests according to the Standard Methods for the Examination of Water and Wastewater (APHA, 1995).

CSO Wastewater:

Primary Sewage Effluent (PSE) from Allegheny County Sanitary Authority (ALCOSAN), Pittsburgh, PA, was used to simulate the wastewater produced during a CSO event. The PSE was collected just after storm events and was transported to the laboratory immediately before the filtration experiments. PSE was selected for its ease in collection, its ability to provide a consistent feed water quality and its ability to accurately represent CSO wastewater. The key parameters of the primary effluent are summarized and compared to typical CSO wastewater in Table 1. A comparison demonstrates that the primary effluent adequately simulates both the gross parameter concentration and the bacterial levels expected of CSO wastewater. The primary effluent did contain less suspended solids than typical combined sewer overflow wastewater. However, the beginning of a CSO event will contain high solids and bacterial loadings, but the concentration will decrease as the storm event continues (US EPA, 1999b).

Membrane Cleaning Procedures:

The membranes were chemically cleaned and the system was disinfected between filtration experiments. The chemical cleaning solution is municipal tap water with 1,500 mg/l sodium hypochlorite (from common household bleach) raised to a pH greater than 11 S.U. by the addition of sodium hydroxide. The cleaning solution was processed through the system for two hours with the permeate valves closed. The system and feed tank were then drained and clean water permeability tests were conducted. The results from the clean water permeability tests were compared to the clean water flux rates using a 0.2 μm filtered deionized water. The clean water permeability test on cleaned membranes verified that the solids buildup and fouling layer had been removed using this procedure.

In addition, further disinfection of the entire bench system (piping, pumps, etc.) between test experiments is needed due to bacterial contamination on the shell side of the membrane caused by the passage of bacteria through larger pore size membranes. The system was disinfected by recirculating a 1,000 mg/L sodium hypochlorite solution at a pH above 11 S.U. for 30 minutes without a membrane in the membrane housing module and with the permeate valve open.

RESULTS AND DISCUSSION – BENCH UNIT

Impact of Pore Size on Permeate Quality

The initial experiments were performed for one to two hours for each of the four membranes to determine the achievable permeate water quality. Specific emphasis was placed on minimizing the permeate passage of pathogenic indicator bacteria (fecal coliforms, *enterococcus* and *e-coli*) as key selection criteria for the membranes that should be investigated further in the field. The results from these initial experiments are presented in Table 2. In general, the permeate water quality improved as the mean pore size of the membrane decreased (Figure 2). The membranes with a pore size greater than 0.2 μm produced a variable bacterial permeate water quality, while the 0.2 μm membrane consistently produced permeate with no or minimally detectable bacteria.

Impact of Fouling Layer on Permeate Quality

The variation in bacterial levels depicted in Table 2 can be attributed to the buildup of the fouling layer on the membrane surface during the filtration run. As the fouling layer develops, the permeate flux decreases due to the increase in the resistance of the fouling layer. As can be seen in Figure 3, the resistance of the fouling layer increases rapidly and then levels off at approximately 160 x 10^{-10} 1/m. The formation of the fouling layer benefits permeate quality for the 2.0 μm and 0.8 μm membranes, while the 0.2 μm membrane consistently produced a permeate water quality with no detectable fecal coliforms bacteria (Figure 4). These results are supported by previous research in the literature that demonstrated an initial breakthrough of fecal coliform bacteria, with an increase in bacterial rejection as the membrane reaches steady-state operation (Till et al., 1998).

Impact of Backpulse Operations on Rates of Flux

The use of a backpulse has been demonstrated to reduce membrane fouling and to maintain an elevated flux during filtration of a primary sewage effluent (Gan, 1999). A series of short-term testing was performed to confirm the benefit of the backpulse on permeate flux rates, an example of which is shown on Figure 5. For this example, the 0.8 μm membrane was operated with a backpulse frequency of 30 seconds and duration of 0.5 seconds for the first hour. During this period, membrane filtration approached a steady state at a flux of 175 L/hr-m^2. The backpulse was then switched off and a 63% decrease in flux rate was observed within 15 minutes. The backpulse was then turned on and the membrane was operated for another 45 minutes. The restarted backpulse could not completely restore the permeate flux to the original rate of 175 L/hr-m^2, but it is clear that it is effective in maintaining higher permeate flux. The data suggest that backpulsing should continue for the entire duration of a microfiltration operation. This is best illustrated on Figure 6 where microfiltration runs on the order of 50 hours were conducted with small flux declines.

Ultimately, the incremental costs of back pulsing must be balanced against the benefits of prolonged filtration runs and higher permeate flux rates. Table 3 summarizes the influence of backpulsing frequency on the 24 hour flux rates for both the 0.2 and 0.8 μm membranes. Two experiments were conducted under identical operating conditions for 24 hour period with a backpulse frequency of 60 and 94 seconds and a backpulse duration of 0.5 seconds, a cross flow velocity of 6.56 m/s, a transmembrane pressure of 1.2 bar, and at ambient temperature of 24 (+/- 2) ^{o}C. The increase in back pulsing frequency from once every 94 seconds to once every 60 seconds resulted in an 18.3 percent increase in permeate flux for the 0.2 μm membrane and a similar 13.1 percent increase for the 0.8 μm membrane.

As suggested by table 4, back pulsing has an adverse effect on the permeate bacterial quality for larger pore-sized membranes. Short-term (up to two hours) experiments were conducted with and without backpulse for the 2.0, 0.8 and 0.2 μm membranes. The membranes were operated under identical conditions with a transmembrane pressure of 1.2 bar, a cross flow velocity of 6.56 m/s and within a temperature range of 21-25^{o}C. The frequency of the backpulse was controlled at 94 sec with 0.5 sec duration. The use of the backpulse facilitated a greater passage of bacteria (measured as fecal coliforms) for the 2.0 and 0.8 μm membranes, while the 0.2 μm membrane consistently produced permeate with no detectable fecal coliform bacteria. It appears that the backpulse disruption of the fouling layer on larger pore-sized membranes allows for some bacterial passage from primary effluent which is of no apparent consequence for the 0.2 micron sized pore membranes. .

Longer-Term Performance

To better simulate an actual CSO event and to understand long-term performance of ceramic microfiltration systems, 48-hour filtration experiments were conducted using 0.8 and 0.2 μm membranes operated at a transmembrane pressure of 1.2 bar, a cross flow velocity of 6.56 m/s, a backpulse frequency of 94 s, a backpulse duration of 0.5 sec, and within a temperature range of 22-28^{0}C. The 0.8 μm membrane initially produced a permeate flux rate of 836 L/hr-m^2 compared to the initial permeate flux rate of 507 L/hr-m^2 for the 0.2 μm membrane. As can be seen in Figure 6, the 0.2 μm produced a slightly greater permeate flux rate than the 0.8 μm membrane in the later stages of the experiment; the final flux rates being 202 L/hr-m^2 and 181 L/hr-m^2 for the 0.2 μm and 0.8 μm membrane, respectively. The slightly lower permeate flux rate for the larger pore size membrane is believed to be due to the difference in the fouling mechanism; smaller pore sizes reject many of the smaller particles that can enter the larger pores and create more severe internal fouling. Furthermore, continuation of the back pulsing program for the duration of the experiment resulted in a fairly constant flux rate. Additional information and details of experimentation can be found in the work of Bendick (2003).

SUMMARY AND CONCLUSIONS – BENCH UNIT

This study demonstrated that dilute (COD ~ 100 mg/l) primary sewage effluent treatment with ceramic microfiltration membranes produced a permeate containing minimal levels of suspended solids, fecal coliform, *enterococcus,* and *e-coli* and reduced BOD, and COD. The 0.2 μm membrane appears to be a virtual barrier to bacteria, while the 0.8 μm membrane allows for some bacterial breakthrough. Over time, the 0.2 μm membrane produced a slightly greater permeate flux rate than the 0.8 μm membrane. Such behavior is believed to occur due to severe internal fouling of the larger pore size membrane.

The backpulse was effective in maintaining an elevated permeate flux, but an increase in bacterial breakthrough is observed for membranes greater than 0.2 μm. An increase in backpulse frequency facilitated greater permeate flux rates over a prolonged time period.

The combination of permeate water quality and long-term flux rates suggest that a 0.2 μm membrane would be the most appropriate ceramic membrane for the treatment of combined sewer overflow under the experimental conditions used in this study.

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Table 1. Comparison of primary effluent and combined sewer overflow wastewater

	BOD_5 (mg/)	C-BOD (mg/)	COD (mg/)	TSS (mg/)	Total N (mg/)	NH_3-N (mg/)	Total P (mg/)	Fecal Coliforms[3] (cfu/100 mL)
CSO Range[1]								
	25-100	**-**	**-**	**150-400**	**3-24**	**-**	**1-10**	**10^5-10^7**
Primary Effluent[2]								
Min.	**-**	**23**	**44**	**24**	**-**	**2**	**1**	**10^6**
Max.	**-**	**70**	**218**	**184**	**-**	**13**	**3**	**10^6**
Avg.	**-**	**40**	**125**	**55**	**-**	**7**	**2**	**10^6**

[1] US EPA, 2001, [2] ALCOSAN Monitoring Data, [3] University of Pittsburgh Lab Data

Table 2. Primary effluent and permeate water quality

Primary Effluent				
	COD (mg/L)	Fecal Coliforms (cfu/100 ml)	E Coli [1] (cfu/100 ml)	Enterococci [1] (cfu/100 ml)
	61-108	1,660,000 - 2,165,797	1,215,000	133,000

Permeate				
Pore Size (μm)	COD (mg/L)	Fecal Coliforms (cfu/100 ml)	E Coli [1] (cfu/100 ml)	Enterococci [1] (cfu/100 ml)
5.0	48 – 80	750,000 - 1,240,000	905,000	108,000
2.0	45 – 52	4,100 - 46,500	115,000	1,950
0.8	32 – 42	16 – 450	30	ND
0.2	24 – 31	ND	ND	ND

[1] E Coli and Enterococci data consist of a single sample set

Table 3. Impact of backpulse frequency on permeate flux

	Flux (L/hr-m2)	
Backpulse	Pore Size (μm)	
Frequency (s)	0.2	0.8
60	273	229
94	223	199

Table 4. Impact of backpulse on permeate water quality

Pore Size (μm)	Backpulse (on/off)	Fecal Coliforms cfu/100 mL
2.0	on	6,212
	off	4,100
0.8	on	116
	off	16
0.2	on	ND
	off	ND

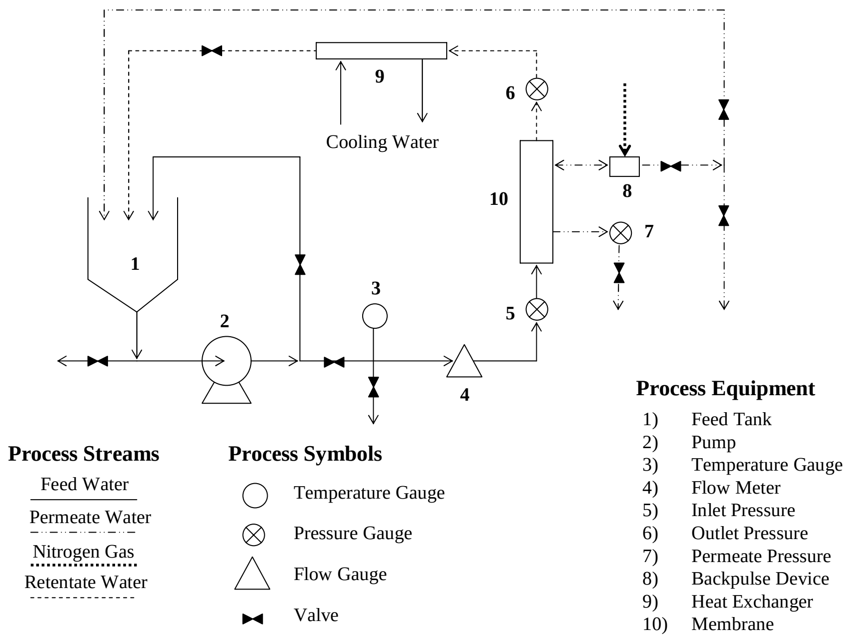

Figure 1. Schematic of the cross-flow microfiltration system

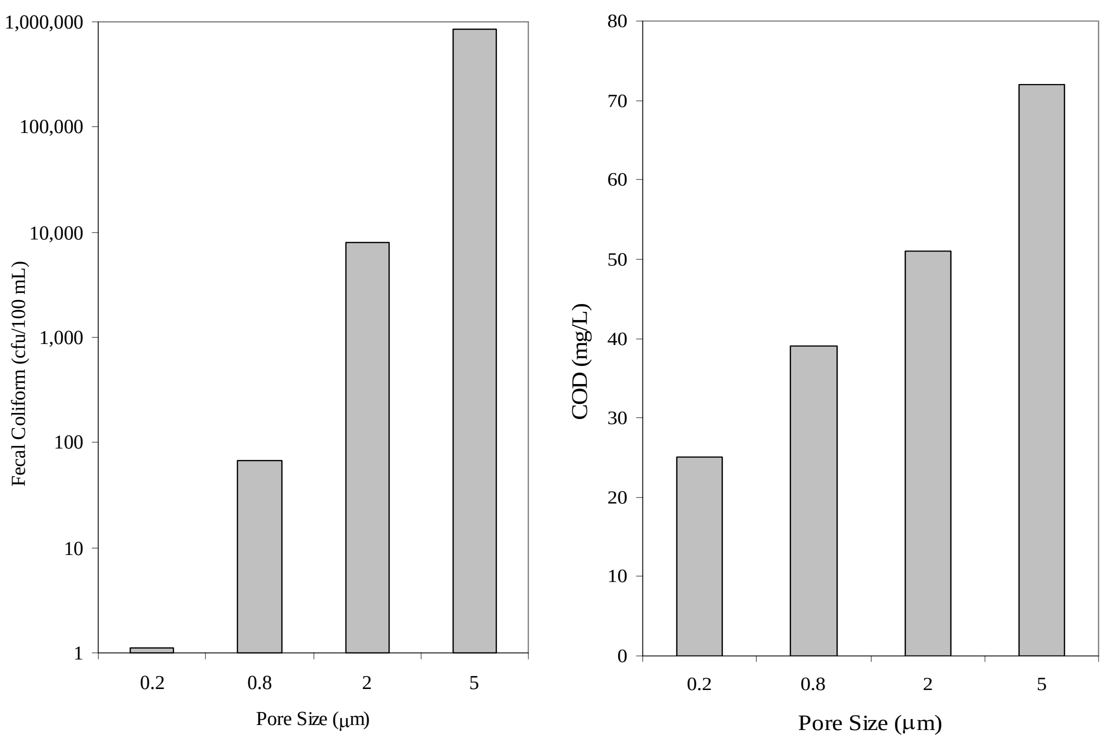

Figure 2. Impact of pore size on permeate water quality

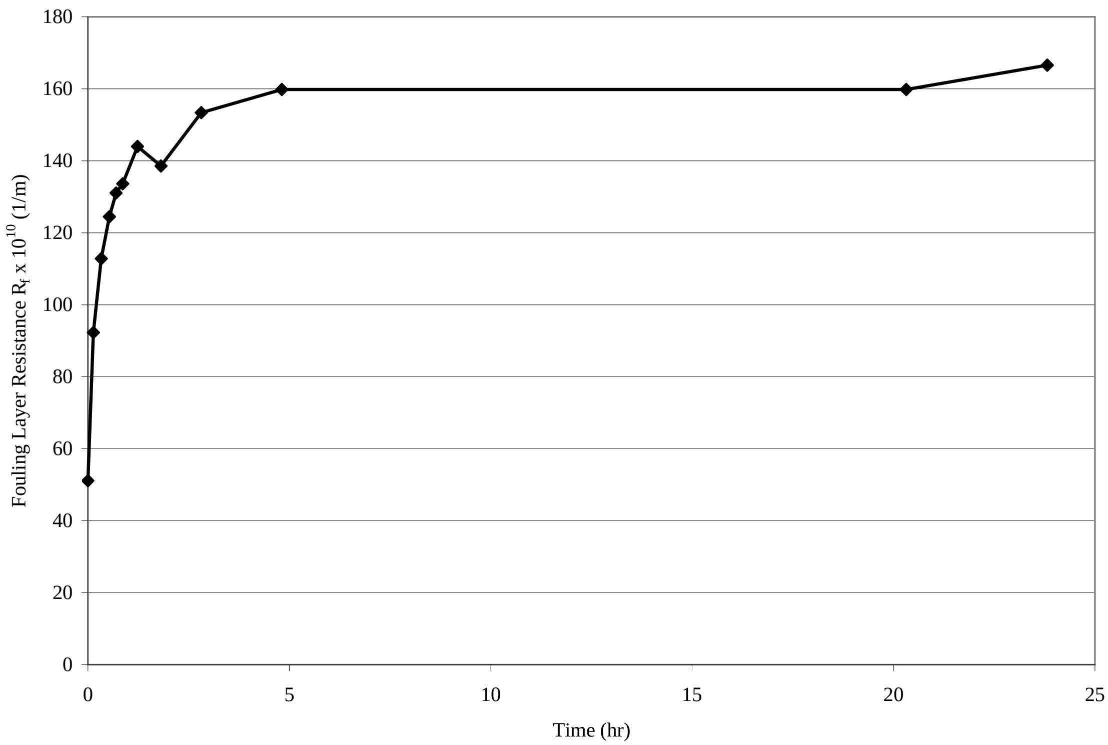

Figure 3. Increase in the fouling layer resistance with time

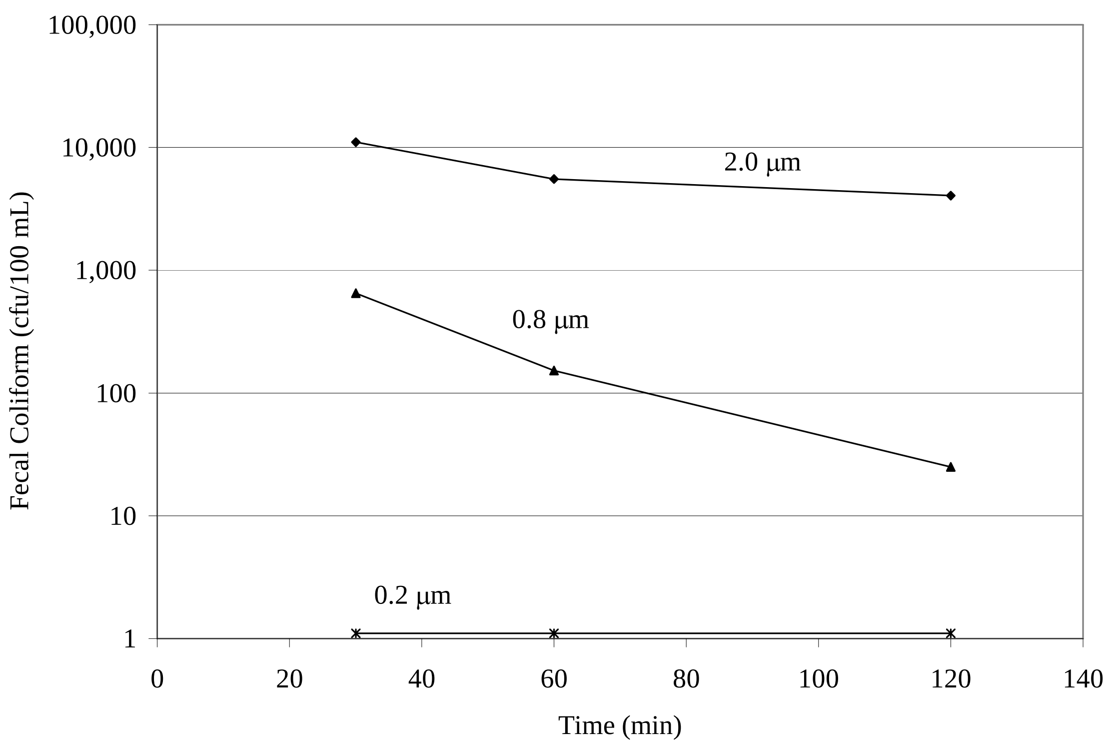

Figure 4. Impact of operating time on permeate quality

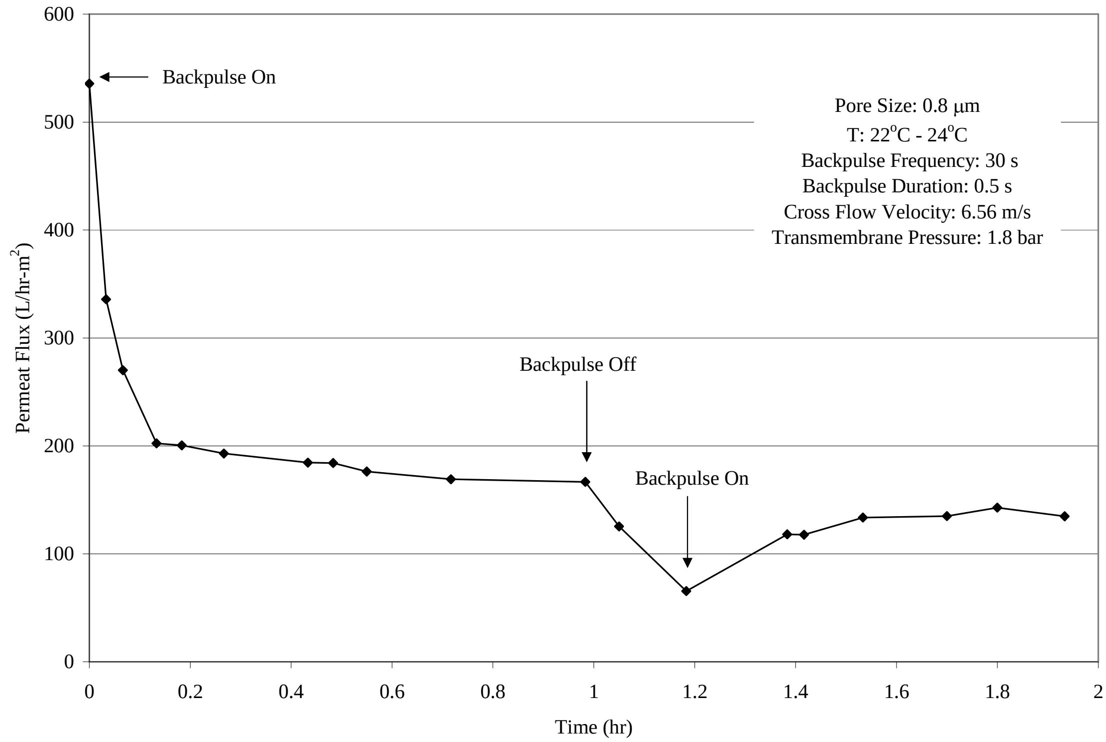

Figure 5. Impact of backpulse on flux rate

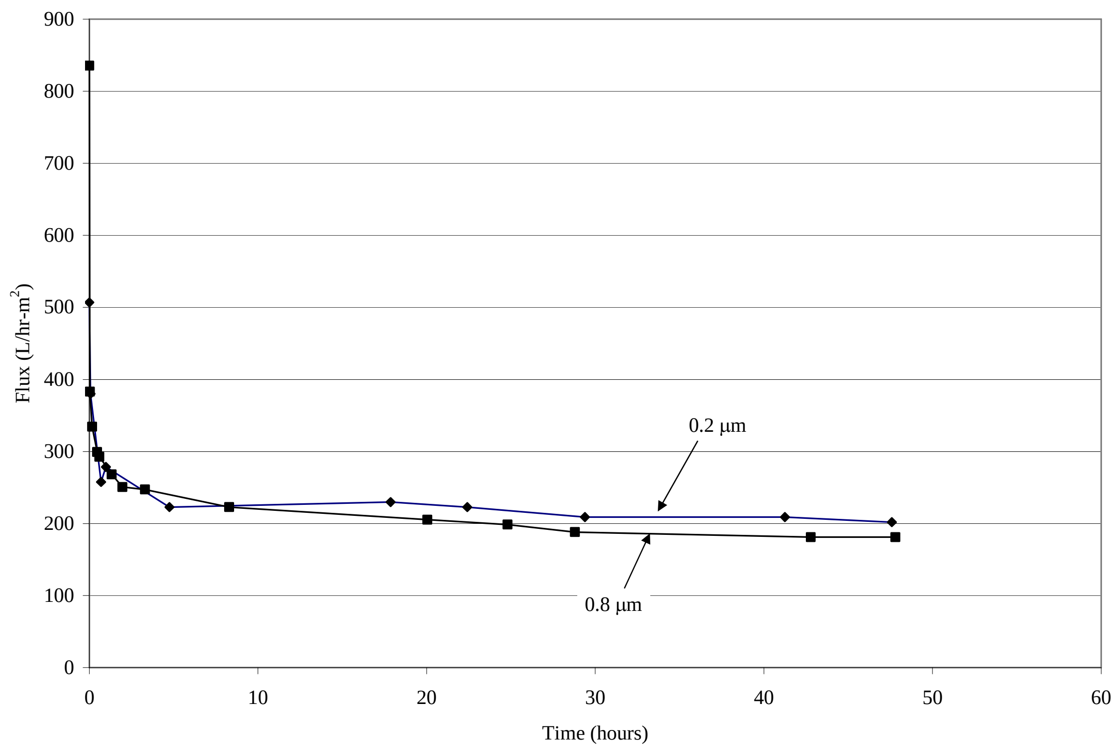

Figure 6. Long-term flux rate

Section 3

PILOT SCALE Cross-Flow Microfiltration of Combined Sewer Overflow at ALCOSAN

MATERIALS & METHODS-PILOT UNIT

Pilot-scale testing was conducted at the Allegheny County Sanitary Authority (ALCOSAN), Pittsburgh, PA, using fresh primary sewage effluent (PSE) to simulate combined and sanitary sewer overflows. The field system was a commercial cross-flow filtration unit, USFilter Silverback® Model 150 microfiltration system, which is typically used to treat industrial alkaline cleaners. Modifications to the commercial unit included a variable frequency pump drive to allow for the independent study of various transmembrane pressures and cross flow velocities on the permeate flux rate and computerized instrumentation and data recording.

The modified system (Figure 7) consists of a 3-HP suction pump, a settling tank, a process tank, a small permeate reservoir, an in-line flow meter with electronic output, a module housing one commercial-sized ceramic membrane, an automatic backpulse device, eight process control ball valves, two solenoid valves for the backpulse air and the feed wastewater, two floats to monitor the tank water level, and three electronic pressure transducers to monitor the inlet, outlet and permeate pressures. The filtration system is capable of operating at a cross flow velocity of 0.5 m/s to 5 m/s and an inlet pressure to the membrane housing of up to 3.5 bar. The backpulse device uses 4.1 bar of compressed air supplied from an external air compressor. The device sends pressurized air to the solenoid valve on the permeate side of the membrane where a small amount of clean permeate water is stored in a reservoir. The solenoid valve is programmed to open at periodic intervals to create a greater pressure on the permeate side of the membrane, thereby redirecting the flow from the permeate side of the membrane to the feed side of the membrane. The frequency and duration of the backpulse are controlled independently.

Permeate flow measurements were taken both electronically and using a graduated cylinder and stopwatch. The temperature in the process tank was measured using a handheld thermometer. City water running through a commercial garden hose placed in the process tank was used to maintain relatively constant temperature (± 5^{o}C). In addition, the unit was instrumented to continuously monitor pressures and flow rates with electronic sensors controlled by a National Instruments Field Point I/O module and Lab View software.

Experiments were conducted with five Membralox® alpha alumina membranes with mean pore size of 0.05, 0.2, 0.5, 0.8 and 1.4 μm. Each tubular membrane is 1.020 m in length and consists of 19 channels that are 6 mm in diameter. Total surface area of each membrane module is 0.36 m^2. The ceramic membranes are durable and capable of withstanding a pressure limit of 7.9 bar, a temperature limit of 220 ^{o}C and a pH range of 0-14.

Laboratory Analysis

The ALCOSAN laboratory conducted all water quality analysis with minimum holding times. The samples were analyzed for fecal coliforms (Standard Method 9222D), *E Coli* (EPA Method 9223B), *Enterococci* (EPA Method 9223B), 5-day biochemical oxygen demand (BOD) (Standard Method 5210B), chemical oxygen demand (COD)

(EPA Method 410.4), suspended solids (SS) (Standard Method 2540D) and ammonia nitrogen (NH_3-N) (EPA Method 350.2). The *E Coli* and *Enterococci* analysis used a defined substrate methodology. Samples are mixed with specific media and are incubated at specific temperatures. If the indicator organisms are present, they will cleave the chromofluorogenic substrate with a specific enzyme, producing a chromogen/fluorogen that will produce a distinct color/fluorescence (Yakub et al., 2002). Quantification was performed using a standard MPN algorithm.

Membrane Cleaning

Routine chemical cleaning and disinfection of the membranes and the entire pilot system was done between filtration experiments. The standard chemical cleaning solution was prepared by diluting household bleach (~6% NaOCl) with municipal tap water to produce a solution containing 1,500 mg/l sodium hypochlorite and if needed, the pH in the feed tank was adjusted above 11 with NaOH pellets. This cleaning solution was processed through the system with the permeate valves closed. The temperature was approximately 10^{o}C at the beginning of the cleaning process. Cleaning proved unsuccessful at lower temperatures and it was decided to allow the solution to recirculate overnight so that a cleaning temperature of approximately 45^{o}C could be attained due to heat supplied from the pump recirculating the cleaning solution through the membrane and feed tank. The sodium hypochlorite solution was effective in cleaning the membrane most of the time. However, the increase in the number of cleaning cycles and the increase in the concentration of suspended solids in the feed required a more aggressive cleaning solution. In that case, a 1% nitric acid solution was recirculated overnight to reach a cleaning temperature of 45^{o}C.

The pilot system was disinfected between experiments to eliminate contamination on the permeate side of the membrane caused by the passage of bacteria through larger pore size membranes used in this study. The system was disinfected by recirculating a 1,000 mg/L sodium hypochlorite solution at a pH above 11 for 30 minutes without a membrane in the housing module and with the permeate valve open.

Membranes were stored in a 1,500 mg/L sodium hypochlorite solution with a pH above 11 when not in use. Storage in this solution aided the cleaning process and prevented biological growth.

RESULTS AND DISCUSSION-PILOT UNIT

Prior to wide-spread implementation of ceramic cross-flow microfiltration for combined or sanitary sewer overflow treatment, the technology needs to demonstrate consistent removal of pollutants to satisfactory levels, with permeate quality suitable for reclamation or discharge to surface waters. There are currently no specific water quality or treatment standards that specifically apply to combined sewer overflows in Pennsylvania; however, secondary treatment standards are applicable to sanitary sewer overflows. In addition, a literature review revealed several applicable water quality standards such as Pennsylvania Department of Environmental Protection (PADEP) surface water quality standards, surface water quality criteria suggested by the US EPA, Pennsylvania Department of Health standards and Ohio River Valley Sanitation Commission standards. Such standards vary from State to State.

For this study, the development of a set of water quality treatment objectives was based on all PA regulations and standards that were combined with a selection of the most stringent level for any given parameter. The "benchmark" parameters and desired treatment objectives for this study are: fecal coliforms (200 CFU/100 mL), *E Coli* (133 CFU/100 mL), *enterococci* (26 CFU/100 mL), suspended solids (30 mg/L) and biochemical oxygen demand (30 mg/L).

Impact of Pore Size on Permeate Water Quality and Flux Rate

Five membranes with a mean pore size below 2.0 μm (0.05, 0.2, 0.5, 0.8 and 1.4 μm) were screened based on the permeate water quality and flux rates to select preferred pore sizes for further evaluation. Membranes were operated with a transmembrane pressure of 0.7 bar, a cross flow velocity of 2.7 m/s and a backpulse frequency of 60 seconds. Permeate was discharged to the drain and fresh primary effluent filled the process tank. Each experiment in this phase of the study was performed in duplicate for a period of 6 hours and all flux rates were corrected to 20^{o}C using the Equation 8. The feed and permeate water quality was sampled after 0.5 and 5.5 hours of operation. The membranes were chemically cleaned with alkaline sodium hypochlorite solution prior to each experiment to avoid contamination from prior experiments.

Water quality results from all experiments are summarized in Table 5. These results indicate that all membranes evaluated in this study readily exceeded the benchmark water quality objectives for pathogenic bacterial indicator organisms, BOD_5 and suspended solids. All membranes, with the exception of the 0.8 μm pore size membrane, demonstrated non-detectable levels of indicator bacteria in the permeate. Higher bacterial removal by the 1.4 μm membrane when compared with 0.8 μm membrane was surprising and the experiment with 1.4 μm membrane was repeated. The 1.4 μm membrane was chemically cleaned and operated as before, but this time for a period of 24 hours. The permeate water quality was analyzed after 0.5 and 24 hours and again, no indicator bacteria were detected thus confirming the initial observation.

In addition to monitoring the water quality, the flux rates were monitored throughout each experiment. As expected, the initial permeate flux rate was very high and was followed by a rapid decrease until about 15 minutes into the experimental run, when a gradual decrease was observed. For the purposes of this demonstration, a continuous operating period of 6 hours was considered sufficient to reach operational “steady state” as the change in the flux rate was bellow 1% within the last hour of operation. Therefore, the data collected after 6 hours of continuous operation are used to select membranes for further evaluation.

Figure 8 shows the relationship between the 6-hour steady state flux rate and membrane pore size. Numbers shown in parentheses represent the suspended solids concentration in the feed for each experiment. The data show an increase in steady state flux rate with an increase in pore size in from 0.05 μm to 0.2 μm, while similar steady state flux rates were observed for all membranes with pore sizes greater than or equal to 0.2 μm. These results suggest that for pore sizes greater than 0.2 μm, key factor governing the flux rate is not the pore size but the fouling layer that forms on the membrane surface or inside the membrane pores. The 0.2 μm membrane was selected as the preferred membrane for further evaluation. This was based on observations that the steady state flux rates for membranes with pore sizes greater than 0.2 μm exhibited essentially the same steady state flux rates as the 0.2 μm membrane, while having a greater potential for the passage of pathogenic indicator bacteria.

Table 5 includes the feed and permeate water quality results from all long term experiments for fecal coliforms, *E-Coli*, *enterococci*, BOD_5, COD, SS and NH_3-N. Of the 150 samples taken for bacterial analysis, only 2 permeate samples, both from the 0.8 micron membrane, had detectable levels of pathogenic indicator bacteria, which suggests that microfiltration has the potential to meet secondary treatment standards for disinfection without the use of chemicals. Similar conclusions may be reached when examining permeate quality with respect to suspended solids (1 mg/L) and BOD_5 (18 mg/L). The pilot plant data showed that the 0.2 μm membrane produced an average SS removal of 99% and an average BOD removal of 83% for the average influent BOD of only 103 mg/L.

Impact of Cross Flow Velocity and Transmembrane Pressure on Steady State Permeate Flux Rate

An experimental test matrix for the 0.2 μm membrane was developed to study the influence of cross flow velocity and transmembrane pressure on permeate flux rate. The matrix consisted of three transmembrane pressures (0.7, 1.4 and 2.1 bar) and four or five cross flow velocities (0.5, 0.9, 1.8, 2.7 and 3.7 m/s). Each experiment was performed for a period of 6 hours with backpulse frequency of 60 seconds and with the permeate recycled to the process tank to maintain constant feed concentration. The membrane was chemically cleaned with alkaline sodium hypochlorite solution prior to each experiment to ensure consistent initial conditions. The data presented in this section are always based on the flux rate corrected to 20^{o}C. Feed and permeate water quality samples were taken after 0.5 and 5.5 hours of operation.

The steady state flux rate as a function of the cross flow velocity for each transmembrane pressure is shown in Figure 9. Values in parentheses are actual suspended solids concentrations in the feed. These results suggest that the permeate flux rate increases with an increase in cross flow velocity until the velocity reaches 1.8 m/s. Beyond 1.8 m/s, the increase in the flux rate is negligible, thereby further suggesting that the resistances to mass transfer are controlled by mechanisms associated with removable surface foulants and not strictly pore size. Experimental results of this study suggest that the preferred cross flow velocity for process engineering purposes using these ceramic membranes is 1.8 m/s.

The steady state flux rate as a function of transmembrane pressure for each cross flow velocity is shown in Figure 10. The selection of a preferred transmembrane pressure is not as straightforward as the selection of a preferred cross flow velocity. The data on Figure 10 indicate that a slight increase in steady state flux rate is achieved at higher transmembrane pressures for cross flow velocities of 1.8 m/s or greater. The data collected at 1.4 and 2.1 bar for low cross flow velocities showed no increase in flux with an increase in transmembrane pressure. Such behavior indicates that the scouring effect at cross flow velocities below 1.0 m/s may not be sufficient to overcome adhesion and compaction of solids on the membrane surface that would is induced by these pressures. Overall, the data suggests that low transmembrane pressures (<2.1 bar) will be adequate, especially for lower cross flow velocities.

Impact of Suspended Solids in the Feed on Steady State Permeate Flux Rate

Multistage membrane filtration designs are often used to achieve higher overall clean water recovery rates. In multistage systems, the retentate from the first stage is feed water to the second stage. This pattern is followed and all clean water permeates may be collected together for discharge or reuse. However, in such designs, the suspended solids concentration of feed water to each subsequent stage is increased. In order to develop a system that can handle a variety of influent suspended solids concentrations, the relationship between the feed suspended solids and the steady state flux rate needs to be understood. In order to better understand the relationship between feed suspended solids and flux rates, an experiment, using the 0.2 μm membrane, was conducted over a 5-day period and evaluated the steady state flux rate at four suspended solids concentrations.

In order to better understand the relationship between feed suspended solids and flux rates, a series of experiments with the 0.2 μm membrane was performed with a transmembrane pressure of 0.7 bar, a cross flow velocity of 1.9 m/s and a backpulse frequency of 60 seconds to quantify this relationship.

The process tank was filled with primary sewage effluent and permeate was always returned to the feed tank to ensure constant suspended solids in the feed throughout each experiment. Higher suspended solids concentrations in the feed tank for subsequent tests was accomplished by discharging the permeate from the system, thereby effectively lowering the volume of water while retaining most of the solids in the feed tank (the permeate routinely has less than 1.0 mg/L of suspended solids). If needed, the tank was refilled with fresh primary effluent and the concentration procedure continued until a desired suspended solids concentration in the feed tank was achieved. The steady state flux rate from that experiment along with the results from this experiment is given in Table 6. All flux rates and data presented within this section are corrected to 20^{o}C. As expected, Table 6 suggests that the steady state flux rate decreased as the feed suspended solids concentration is increased. However, the influence was less pronounced as the concentration became greater than about 239 mg/L and reached a relatively constant flux value of about 105 L/hr-m^2 for all concentration greater than about 500 mg/L. In all cases, the permeate flow rate remained constant for each 24 hour experimental condition.

Impact of Temperature on Steady State Flux Rates

Darcy's Law suggests that as the temperature of a fluid increases, the viscosity of the fluid decreases leading higher flux rates. A series of experiments at six different temperatures over the range of $12^{o}C$ to $26^{o}C$ was conducted using the 0.2 μm membrane to evaluate and confirm the field utility of the empirical temperature correction factor used shown as equation 4.

The process feed tank was drained and refilled with fresh primary effluent at a starting temperature of $12^{o}C$. The membrane was chemically cleaned with alkaline sodium hypochlorite solution prior to the experiment to ensure accurate flux data. After filling the feed tank with primary sewage, the membrane unit was operated for a 6 hour period with a transmembrane pressure of 1.4 bar, a cross flow velocity of 1.9 m/s, a backpulse frequency of 60 seconds with the permeate recycled to the feed tank to ensure a constant feed concentration. The cooling water was shut off (Figure 7), and the temperature in the feed tank gradually increased to $26^{o}C$ as a consequence of the energy supplied by the recirculation pump. Flux rates were monitored as the temperature increased. The data show that as the temperature increased from $12^{o}C$ to $26^{o}C$, the flux increased by nearly 30%, from 82 to 116 L/hr-m^2. Application of equation 8 to the data showed a correlation coefficient of 0.99 suggesting that the temperature correction factors used in this equation are reasonable for design.

Impact of Operating Mode on the Permeate Flux Rate

Two common modes of operation for cross flow microfiltration systems are "constant pressure" and "varying pressure" operation. Constant pressure operation controls only the feed pressure while keeping the permeate at atmospheric pressure. This mode of operation allows for the treatment of as much wastewater as can pass through the membrane under this constant driving force. Under these conditions, the permeate flux rate is initially very high, followed by a rapid decrease and then a gradual decrease towards a constant flux rate.

Varying pressure operation attempts to keep the permeate flux rate constant throughout the filtration run. This is achieved by maintaining a constant inlet and outlet pressure while varying the permeate pressure. When the permeate pressure decreases to atmospheric, this mode of operation essentially becomes constant pressure mode.

During constant pressure operation, it is believed that the initial rapid flow of water through the membrane forces particles into the membrane pores, thereby creating more severe internal fouling. It is believed that keeping the initial flux rate constant would create less severe fouling by allowing fewer particles to enter into the membrane pores.

To simulate a wet weather combined or sanitary sewer overflow event and to evaluate the different modes of operation, a three-day (72-hour) experiment was conducted for each of the two operating modes using the 0.2 μm membrane. The constant pressure experiment was conducted with a 1.4 bar transmembrane pressure, a 1.8 m/s cross flow velocity and a backpulse frequency of 60 seconds. The varying pressure experiment attempted to maintain a constant flux rate of 150 to 160 L/hr-m^2 and was conducted with a feed pressure of 1.4 bar, a varying permeate pressure, a 1.8 m/s cross flow velocity and a backpulse frequency of 60 seconds. The membranes were chemically cleaned with alkaline sodium hypochlorite solution prior to each experiment to ensure identical initial conditions. For each experiment, the permeate was discharged to the drain and fresh primary effluent was fed to the process tank. All flux rates are corrected to 20^{o}C.

The results from these tests, summarized in Figure 11, show that both modes of operation produced similar flux rates after 1-hour of operation. The permeate pressure in the varying pressure operation was lowered to atmospheric after only two hours in an effort to maintain a permeate flux rate of 160 L/hr-m^2. The varying pressure mode resulted in higher permeate flux rate for the first 24 hours, but not after 48 hours. It may be concluded that the long-term performance of this membrane system is not greatly influenced by the mode of operation employed. Table 3, which includes feed and permeate BOD and COD values during these experiments, clearly illustrates that operation in either mode can produce permeates that consistently meet conventional BOD quality requirements for secondary treatment.

Membrane Maintenance and Cleaning

Chemical cleaning of membranes is critical for the operation and maintenance of any membrane facility. The purpose of chemical cleaning is to reduce or remove overall membrane foul ants to restore flux rates to initially clean conditions. Chemical cleaning of membranes in this study was done between filtration experiments and the cleaning effectiveness was determined based on the initial flux measurements with city water under identical operating conditions: transmembrane pressure of 1.4 bar, cross flow velocity of 2.7 m/s and backpulse frequency of 60 seconds. During the entire pilot study, the 0.2 μm membrane was cleaned 25 times. In each case, the initial clean water flux after cleaning was measured and compared to the initial value achieved in the first test. The percent of the initial clean water flux that was restored after each cleaning cycle is shown in Figure 12. Throughout the initial fifteen experiments, the alkaline NaOCl solution returned the clean water flux to well above 80% of the original value with many values being over 90% of the original value.

Experiments 15–21 are identified as the experiments conducted at elevated feed suspended solids. As can be seen on Figure 12, the alkaline sodium hypochlorite cleaning procedure proved to be less effective for those experiments and the clean water flux rates decreased to 50-75% of the original value. Visual inspection of the membrane after the 21st cleaning cycle revealed the presence of a brown-color on both the feed and permeates side of the membrane. The membrane was then cleaned with a 1% nitric acid solution, which restored the permeability to its original value. Following this aggressive cleaning with nitric acid, cleaning with the alkaline NaOCl solution after experiments with low suspended solids in the feed was effective again. Consequently, it is suggested that an appropriate cleaning procedure for field applications would be to first use the alkaline sodium hypochlorite, while a dilute nitric acid solution may be used as a second step in the cleaning protocol if necessary.

SUMMARY AND CONCLUSIONS-PILOT UNIT

Results of the bench study indicate that all microfiltration membranes with pore sizes from 0.05 to 1.4 μm are capable of separating pathogen bacterial indicator organisms (fecal coliforms, *E Coli*, *enterococci)* to virtually non-detectable levels, while reducing BOD_5 and suspended solids below secondary treatment standards. These outcomes suggest that membrane filtration of combined and sanitary sewer overflow waters may produce permeate quality suitable for direct discharge to surface waters.

All membranes with pore sizes greater than 0.05 μm produced similar steady state flux rates. It is therefore suggested that 0.2 μm membranes be employed as membranes with larger pores demonstrated no benefit in steady state flux, while having an increased potential for bacterial passage into the permeate.

Permeate flux rates were considered to be optimal (up to 165 L/m^2-hr at 20^{o}C) for a cross flow velocity of 1.8 m/s, suggesting that higher pumping rates for recirculation through membranes may not be justified at transmembrane pressures of up to 2.1 bar. These results also indicate that the scouring effect at low cross flow velocities (< 1.8 m/s) may not be sufficient to overcome the adhesion and compaction of the solids on the membrane surface and that field systems should not be operated at low cross flow velocities or extremely high transmembrane pressures.

An increase in the feed suspended solids concentration will adversely affect the permeate flux rate. Results indicate that the steady state flux rate decreased until the feed suspended solids concentration reached between 300 to 500 mg/L. The steady state permeate flux appears to remain constant at approximately 105 L/hr-m^2 for all feed concentration greater than 500 mg/L. At feed suspended solids concentrations of less than about 300 mg/L, the “varying pressure” mode of operation produced marginally higher flux rates as compared to the “constant pressure” mode during the initial 48 hours of operation. The flux rates then remained constant and similar for the subsequent 24 hours of operation. At suspended solids concentrations greater than 300 mg/L virtually no difference in flux between the varying and constant pressure mode of operation was observed.

A periodic chemical cleaning with alkaline sodium hypochlorite cleaning solution (1,500 mg/L NaOCl) was effective in recovering over 90% of the initial clean water permeability for experiments with feed suspended solids concentrations typical of diluted sewage. Cleaning effectiveness decreased as the feed suspended solids concentration increased. A more aggressive cleaning with a dilute nitric acid (1%) solution restored the initial clean water flux to its original rate. Suggested cleaning procedures in the field would be to first use the alkaline sodium hypochlorite solution, followed by the nitric acid solution as a more aggressive cleaning protocol when necessary.

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Table 5. Average Water Quality Results at Various Pore Sizes

Pore Size (μm)	Fecal Coliforms (CFU per 100 mL)	*E.Coli* (MPN per 100 mL)	*Enterococci* (MPN per 100 mL)	BOD (mg/L)	SS (mg/L)	COD (mg/L)	NH_3-N (mg/L)
Feed[1]							
	7.9×10^6	2.9×10^6	0.2×10^6	103	141	216	9
Permeate							
0.05[2]	< 2	< 1	< 1	18.3	0.6	58.6	5.60
0.2[3]	< 2	< 1	< 1	18.0	1.0	76.0	7.00
0.5[2]	< 2	< 1	< 1	14.3	0.3	34.8	5.00
0.8 [2]	8	8	< 4	16.7	0.4	61.8	5.30
1.4 [2]	< 2	< 1	< 1	18.3	0.4	52.0	6.70
Water Quality Objectives							
Monthly Average	200 [4]	126 [5]	33 [5]	30 [6]	30 [6]	-	-

[1] Average Result from all experiments – Sample Size: 50 [2] Sample Size: 4 [3] Average from all experiments – Sample Size: 50, [4] PA Code Title 25 Chapter 93.7, [5] US EPA 1986 [6] PA Code Title 25 Chapter 92.2

Table 6. Impact of Suspended Solids on Steady State Flux – 0.2 μm

SS (mg/L)	56	239	450	490	640
Steady State Flux (L/hr-m^2)	153	125	110	105	105

Table 7. Variation in Permeate BOD and COD During Constant and Varying Pressure Operation of the Membrane System

Time (hr)			**0**	**24**	**48**	**72**
Constant Pressure Operation	**Flux (L/hr-m^2)**		**176**	**130**	**107**	**104**
	BOD (mg/L)	**Feed**	**112**	**209**	**326**	**292**
		Permeate	**29**	**11**	**10**	**3**
	COD (mg/L)	**Feed**	**179**	**376**	**494**	**628**
		Permeate	**79**	**57**	**64**	**80**
Varying Pressure Operation	**Flux (L/hr-m^2)**		**171**	**120**	**105**	**104**
	BOD (mg/L)	**Feed**	**90**	**118**	**203**	**240**
		Permeate	**28**	**11**	**11**	**6**
	COD (mg/L)	**Feed**	**224**	**267**	**428**	**589**
		Permeate	**92**	**52**	**118**	**48**

Figure 7. Pilot Scale Cross-Flow Microfiltration Process Flow Diagram

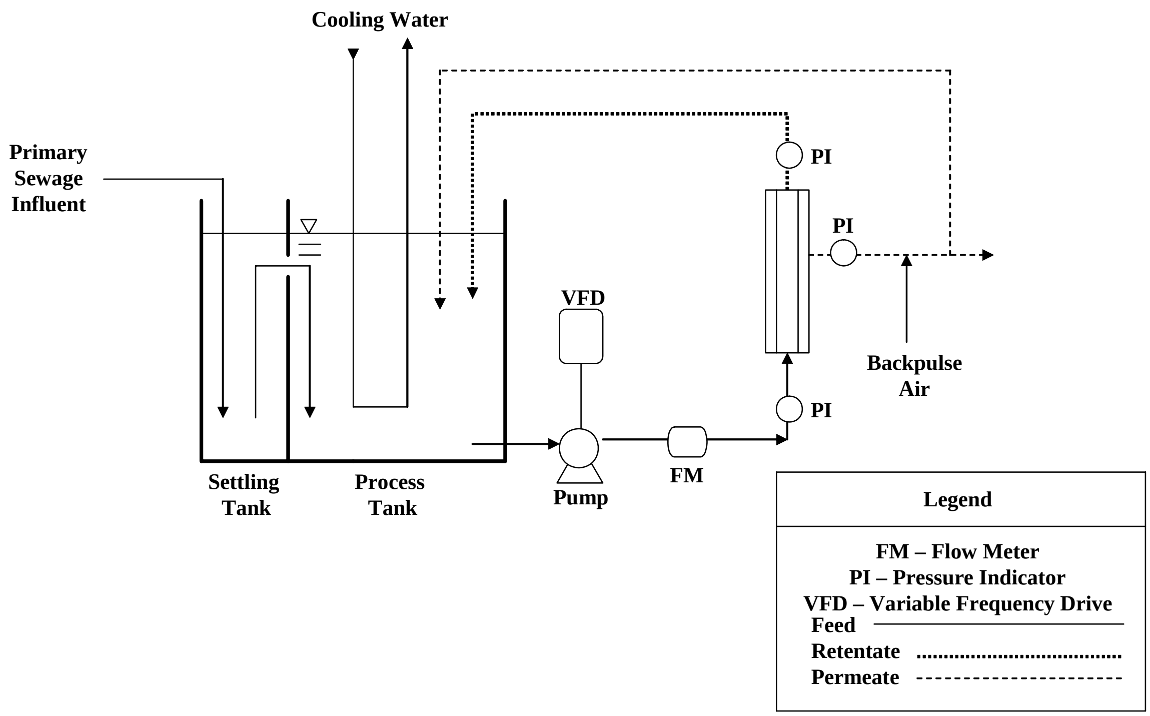


Figure 8. Steady State Flux versus Membrane Pore Size

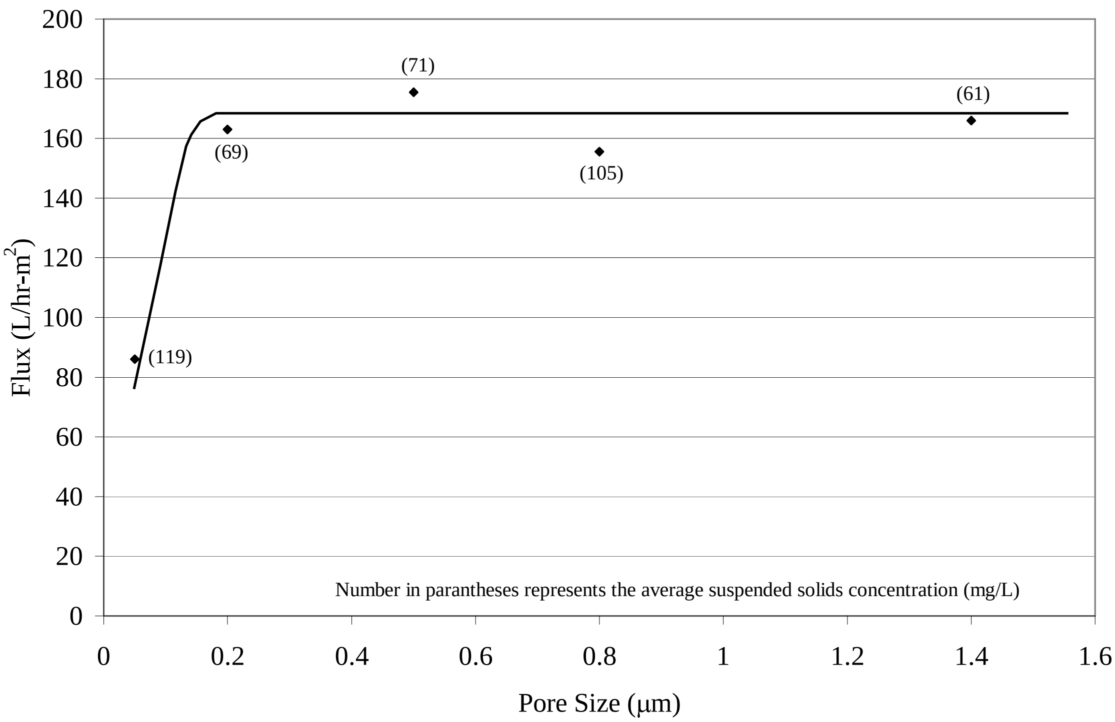


Figure 9. Impact of Cross Flow Velocity on Steady State Flux – 0.2 μm Membrane

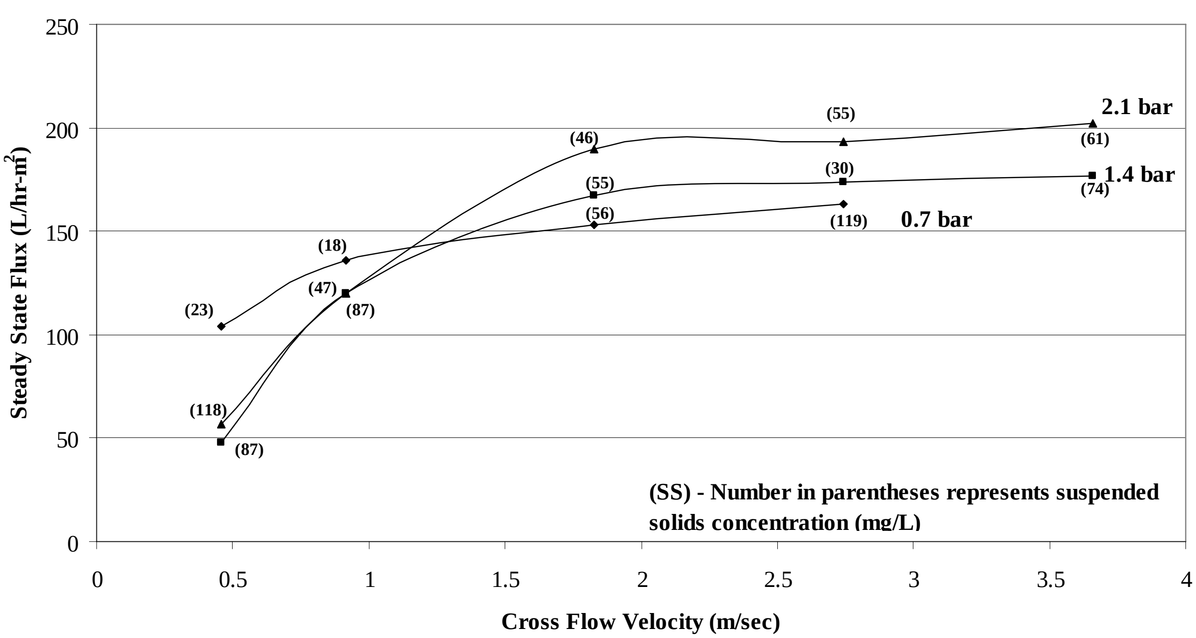


Figure 10

Impact of Cross Flow Velocity and Transmembrane Pressure on 6-Hour Steady State Permeate Flux 0.2 μm Membrane Feed Wastewater - Primary Sewage Effluent Backpulse frequency - 60 sec, T - 20°C

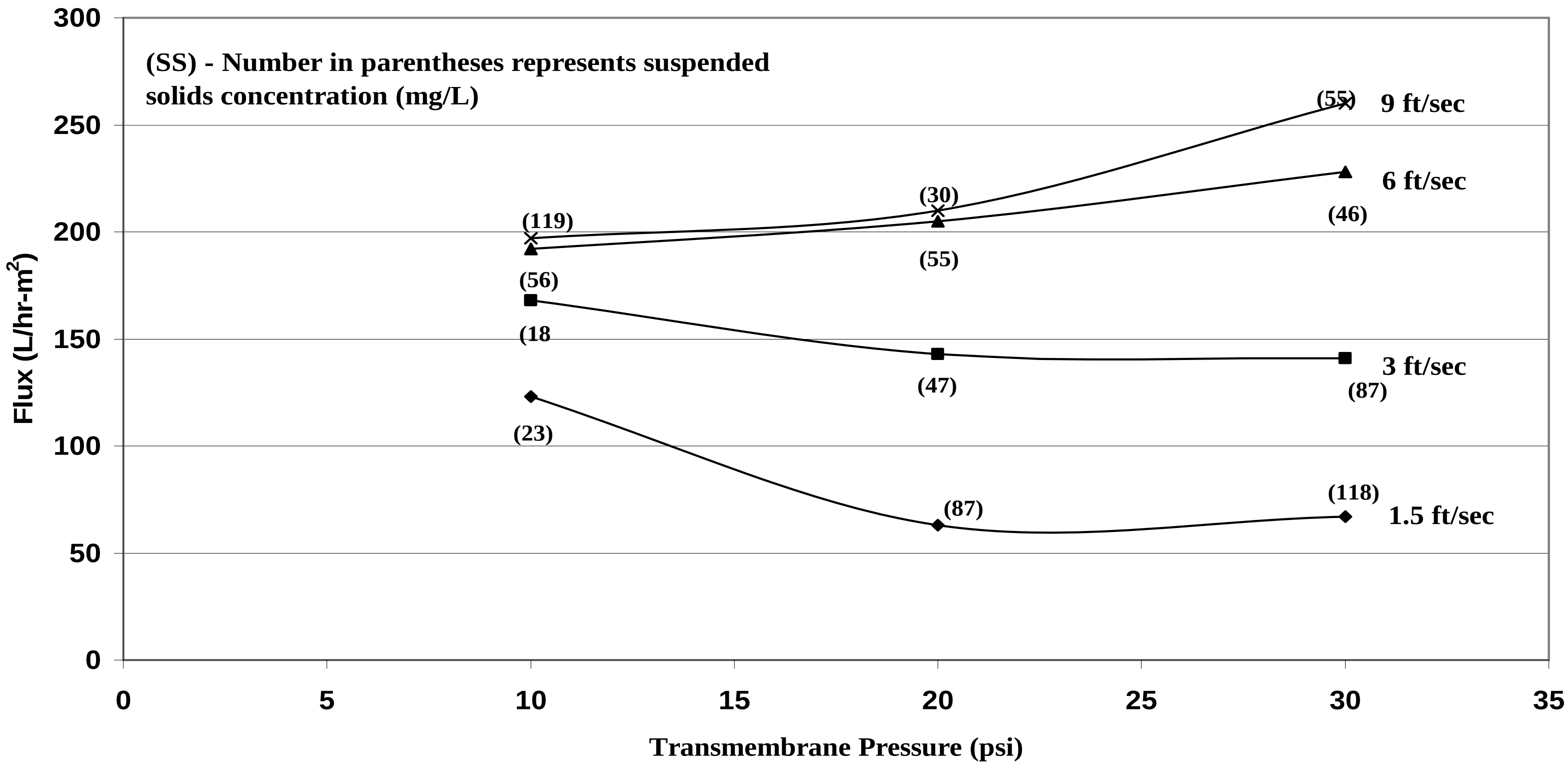


Figure 11. Impact of Operating Mode on Flux – 0.2 μm Membrane

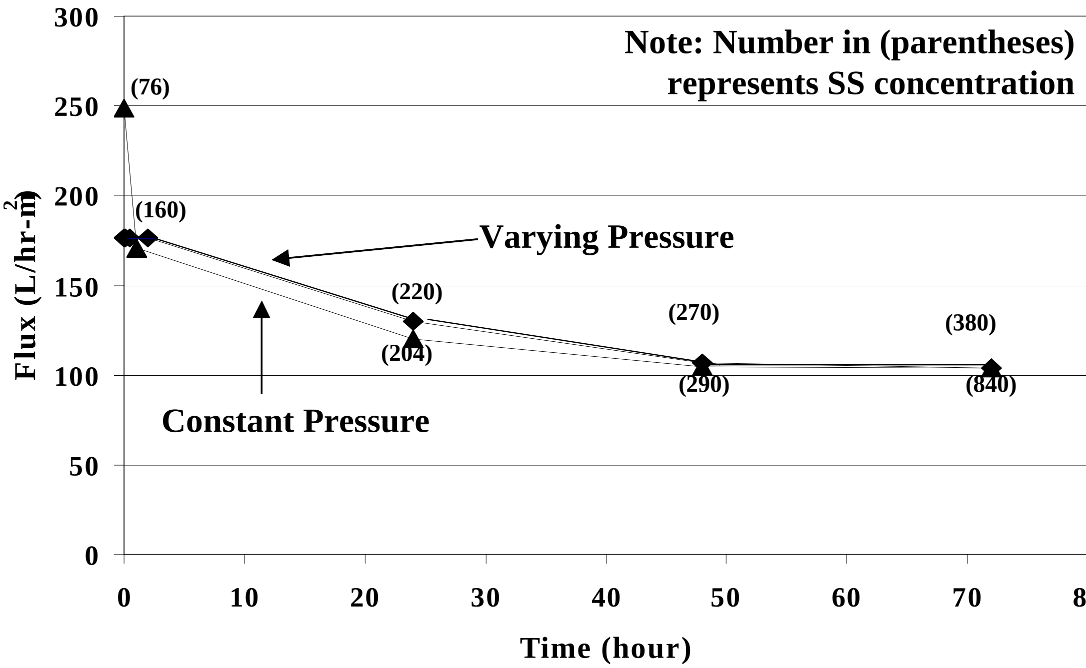
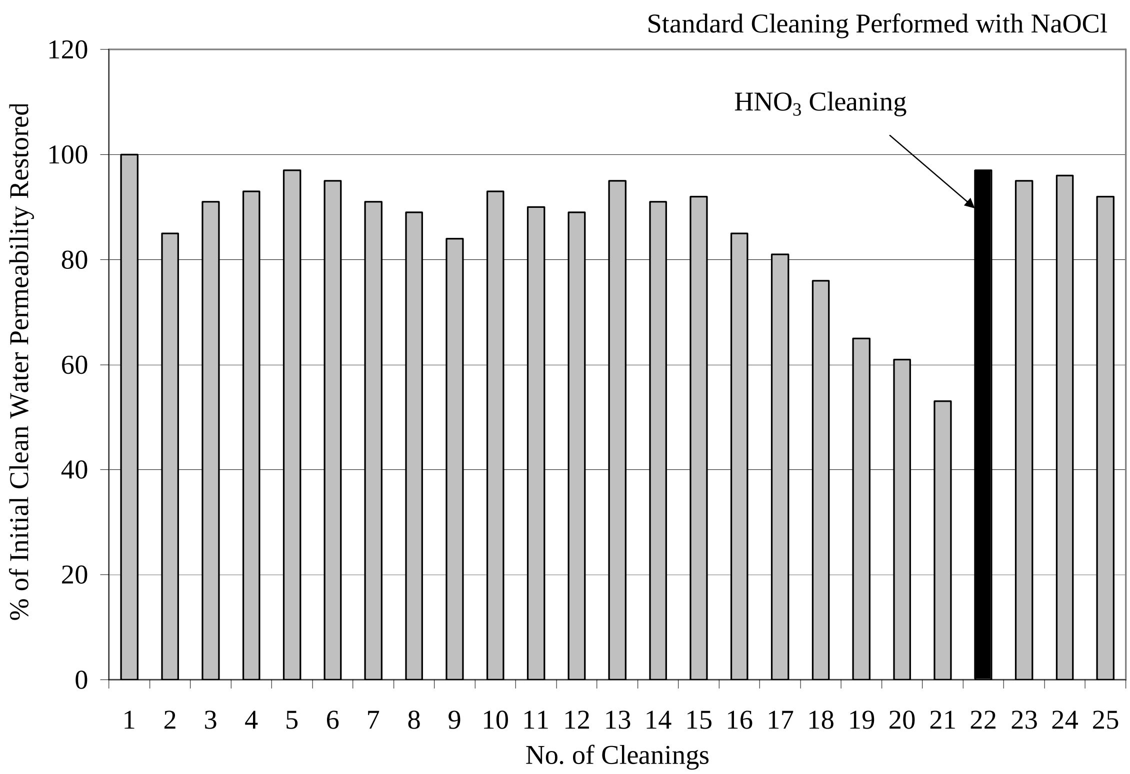


Figure 12. Cleaning Effectiveness – 0.2 μm Membrane



Section 4

Recommended Next Steps to Take

1. The next step should be focused at development of an engineering design complete with rational economic projections. The data contained herein are useful for the development of an engineering design for cross-flow membrane filtration of combined sewer overflows. Accordingly, it would be appropriate to conduct longer duration piloting of both ceramic and cross-flow polymeric membrane systems to better determine operating comparative costs and the influences of seasonal variations.

2. There is no scientific reason to select ceramic membranes over polymeric membranes for application to the management of wet weather flows. The choice should be based on economics for a given location and given set of acceptable vendors. At present, ceramic membranes are more costly ($/square foot), and vendor experience is not as well advanced in the environmental engineering market place as it is for polymeric membranes. This is likely to change in the future. Ceramic membranes, however are inherently more robust and not prone to fracture, tearing, physical abrasion or chemical damage as are polymeric membranes; consequently ceramic membranes are not as likely to need scheduled replacement. Scheduled replacement is a factor when considering polymeric membrane systems.

   There are a significant number of vendors of polymeric membranes in the environmental marketplace. Currently (2005), there are a significant number of environmental engineering technical papers and conference presentations involving application of polymeric systems (*mainly "hollow fiber" systems*) as replacements for secondary clarifiers in "membrane biological reactors". This represents a relatively new market for polymeric membrane vendors who traditionally have found application for their products in the industrial manufacturing sector. Polymeric membranes have a comparatively shorter service life than ceramic membranes and, as outlined above, replacement costs must be considered when doing a life-cycle economic evaluation.

3. Deployment of membrane filtration for management of wet-weather flows may be best accomplished by considering this as "satellite treatment" to be located in multiple modules within the upper reaches of the sewer shed. In this fashion, excess flows may be removed from the sewer shed and either discharged to surface waters, or reclaimed and recycled for alternate uses. Considerations should be given by permitting authorities to allow membrane-filtered CSO water to be directly discharged to surface water.

a. Data presented above suggests that membrane filtration can accomplish enteric pathogenic bacteria elimination without the use of chemicals.

b. The data presented above shows that all suspended organic matter is removed from CSO waters suggesting the permeate (discharge) from membranes can effectively meet the same effluent discharge standards as if it went through a conventional secondary treatment process in a POTW. The data shows that only soluble BOD passes through the membrane filter which is often less than 30 mg/L.

4. Some decisions need to be made and several pieces of design information are required before full-scale systems can be deployed. These include, but are not limited to the following:

   a. Hydrological information for outfalls must be obtained, and decisions must be made as to the degree of CSO control required.

      i. How many wet-weather overflow events are to be allowed? For example, shall the design be based on the 90th percentile of average 24-hour flows (allowing statistically 3 wet-weather discharges per month during maximum flow months [3 out of 30 days]?)
      ii. What shall the <u>averaging time period</u> be for determination of a wet-weather event and existence of an excess wet-weather flow [24 hours? 8 hours? 1 hour?) Note that a physical membrane module must be sized for a given maximum flux rate (gallons per day per square foot of membrane surface) and flows in excess of the design rate must be temporarily stored so that they can later be processed by the membrane. Thus, a "number" is required to determine membrane surface area requirements.

   b. Membrane systems must have some device in-front of them to remove stones and other large objects that can cause blockages. While devices such as "*Swirl Separators*" may be satisfactory, consideration should be given to the utilization of diversion tanks that can temporarily store flows that are in excess of membrane maximum-design throughputs. In addition, such tanks can serve "double-duty" if accommodated with large-particle[1] collection devices and oil skimmers. In this fashion, an economic optimum for a given site can be made between expenditures on membrane surface areas vs. expenditures on storage tank volume.

[1] Particles that are large enough to settle as discrete particles in accord with Stokes law (or "type-1 settling")

5. With the above considerations taken into account, external funding should be developed and made available to provide for:

    a. A pilot/prototype demonstration comparing both polymeric and ceramic membrane commercially available filtration equipment for CSO management, and

    b. A plan to collect engineering information sufficient for a detailed design and cost estimate for polymeric and ceramic membrane filtration at one or more CSO outfalls.

    c. A plan to implement full scale design using commercially available equipment for either polymeric or ceramic membrane filtration systems for CSO control at one or more outfalls within the sewer shed.