NOVEL MEMBRANE PROCESS WITH RAPID BACKPULSING FOR WATER TREATMENT

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A novel membrane process which employs rapid backpulsing to reduce fouling was investigated for the treatment or pretreatment of water streams containing particulates, microorganisms, and/or oil droplets. Various microfiltration membranes and operating conditions were tested for use with yeast suspensions, bacteria1 suspensions, oily wastewaters, and bentonite suspensions. In the absence of backpulsing, in all cases, fouling caused severe flux decline. Rapid backpulsing resulted in nearly a IO-fold improvement in flux for yeast suspensions at the optimal backpulsing frequency. For bacterial suspensions, varying the backpulse duration and frequency resulted in more than a 1 O-fold improvement in the net flux. Rapid backpulsing of bentonite suspensions and oily wastewater also yielded approximately IO-fold improvements over the values recorded in the absence of backpulsiig. An economic analysis shows that membrane filtration with backpulsiig is expected to yield water pretreatment costs approximately one-half of those obtained with conventional flocculation/filtration.

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April 1997

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1. Executive Summary

A novel membrane process which employs rapid backpulsing to reduce fouling was investigated for the treatment or pretreatment of water streams containing particulates, microorganisms, and/or oil droplets. Various microfiltration membranes and operating conditions were tested for use with yeast suspensions, bacterial suspensions, oily wastewaters, and bentonite suspensions; fouled membranes and flux-decline data were analyzed to identify fouling mechanisms; and the rapid backpulsing process was first modeled, then implemented, and then analyzed economically. In the absence of backpulsing, in all cases, fouling caused severe flux decline, generally to levels below 100 $L/m^2{\cdot}h$ (2.8×10^{-5} m/s).[1] The microbial and particulate suspensions produced only external membrane fouling, whereas the oily waste water fouled first the internal and then the external ceramic membranes. Rapid backpulsing resulted in nearly a IO-fold improvement in flux to 680 $L/m^2{\cdot}h$ (1.9×10^{-4} m/s) for yeast suspensions at the optimal backpulsing frequency, in good agreement with predictions of the theory. For bacterial suspensions, varying the backpulse duration and frequency resulted in more than a 1 O-fold improvement in the net flux, to a value of 140 $L/m^2{\cdot}h$ (3.8×10^{-5} m/s). Rapid backpulsing of bentonite suspensions and oily wastewater has yielded net fluxes as high as 2220 $L/m^2{\cdot}h$ (6.4×10^{-4} m/s) and 1260 $L/m^2{\cdot}h$ (3.5×10^{-4} m/s), respectively, also representing approximately lo-fold improvements over the values recorded in the absence of backpulsing. An economic analysis shows that membrane filtration with backpulsing is expected to yield water pretreatment costs approximately one-half of those obtained with conventional flocculation/filtration.

2. Background and Introduction

This project focused on membrane process development for the treatment of water streams containing particulates, microorganisms, and/or dispersed oil. For fresh waters, these contaminants must be removed prior to release of the water to the environment or prior to use of the treated water in domestic and agricultural applications. For salt waters, these contaminants must be removed as a pretreatment step prior to membrane desalting.

Billions of gallons of wastewaters containing oils and particulates are produced each year by metallurgical plants, ships, petroleum and gas operations, industrial washing operations, and other processes (Wahl et al., 1979). Traditional technologies, such as gravity separators, air or gas flotation, chemical flocculation, plate coalescers, and hydrocyclones, are generally able to produce effluents containing as little as 30 ppm dispersed oil and particulates (Vandermeulen and Hrudey, 1987; Powell, 1992). However, these treatment technologies perform poorly on chemically stabilized suspensions and emulsions, very small particles and droplets (G-10 μm in diameter), and soluble components. Moreover, effluents with less than 10 ppm impurities are desired, because of the potential toxic effects of the contaminants and their tendency to foul reverse-osmosis membranes and downstream processing equipment.

Microfiltration and ultrafiltration membranes are able to remove particulates, microorganisms and oils from water, if the membrane material and pore sizes are chosen appropriately. However, they are subject to fouling, which often reduces the permeate flux (volume of water passing through

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[1] Note that 1 $L/m^2{\cdot}h$ = 0.589 $gal/ft^2{\cdot}day$ = 2.8×10^{-7} m/s.

the membrane per surface area per time) below acceptable levels. The completed study investigated the use of rapid backpulsing for controlling membrane fouling and improving the permeate flux; this process has previously shown promise in biotechnology applications (Rodgers and Sparks, 1992; Wenten, 1995; Redkar and Davis, 1995). In rapid backpulsing, the transmembrane pressure is reversed for a few tenths of a second once every few seconds. This results in a hydraulic cleaning of the membrane by forcing permeate back through the membrane in the reverse direction. The process is similar to the common industrial process of backflushing, except that it occurs on a much more rapid time scale. As a result, foulants are removed by backpulsing shortly after they are deposited, and in some cases fouling may be entirely prevented by backpulsing.

3. Conclusions and Recommendations

The use of rapid backpulsing has proved to be extremely successful for the water streams investigated. Net fluxes achieved with backpulsing under optimum conditions are as high as 1000-2000 L/m^2 ($2.8\text{-}5.6 \times 10^{-4}$ m/s), whereas those without backpulsing are an order of magnitude lower. Shorter backpulse durations generally improve the performance, since these minimize the loss of permeate during reverse filtration. Typical durations of backpulses are 0.1-0.5 s. For each suspension and each backpulse duration, there is an optimum backpulse frequency. Higher backpulse frequencies lead to too much negative permeate flow during backpulsing, whereas lower backpulse frequencies lead to too much fouling and flux decline during forward filtration. Typical optimum forward filtration times between backpulses are l-10 s, and the optimum time increases with decreasing feed concentration.

An economic analysis shows that crossflow microfiltration in the absence of backpulsing is not competitive for water pretreatment, except for relatively small applications. In contrast, crossflow microfiltration with backpulsing is economically competitive, costing approximately one-half of the cost for conventional pretreatment.

Commercial development of membrane filtration with rapid backpulsing makes sense economically for water treatment and pretreatment. However, further fundamental study is needed to more fully identify optimum operating conditions and maximum net fluxes for model suspensions and streams representative of practical applications. A particular challenge will be the treatment of streams containing adhesive foulants, for which modifications of membrane surface chemistry may be required in combination with rapid backpulsing.

4. Work Accomplished

A modified work schedule is given in Table 1. This schedule was updated from that originally proposed in order to include bentonite suspensions. With one exception, all of the listed tasks have been substantially completed. Task 4 was not undertaken, due to the recommendation by the Bureau of Reclamation to add studies of bentonite suspensions and to deemphasize studies of oily wastewaters.

Table 1. Modified Schedule of Work

Oct.–Dec. 1995: Testing of Membranes and Operating Conditions

1. Identify best membranes for oily wastewater
2. Identify best membranes for bacterial suspensions
3. Test different oil compositions and rejection of soluble contaminants
4. Test effects of pH, salt, and emulsifiers
5. Identify best operating conditions for oily wastewater
6. Identify best operating conditions for bacterial suspensions
7. Submit First Quarterly Report

Jan.-Mar. 1996: Fundamental Analysis of Membrane Fouling

8. Fit resistance curves to internal and external fouling models
9. Evaluate scanning electron micrographs of fouled membranes
10. Measure mass of foulant deposit for various conditions
11. Test cleaning methods for foulant removal and flux recovery
12. Identify key fouling mechanisms
13. Determine if fouling layers are adhesive or able to flow
14. Submit Second Quarterly Report

Apr.–June 1996: Studies of Rapid Backpulsing

15. Modify experimental apparatus for rapid backpulsing
16. Perform initial backpulsing experiments with oily wastewater
17. Perform initial backpulsing experiments with bacterial suspensions
18. Identify membranes and operating conditions for bentonite suspensions
19. Perform initial backpulsing experiments with bentonite suspensions
20. Determine parameter values for the backpulsing model
21. Submit Third Quarterly Report

July–Sept. 1996: Optimization of Rapid Backpulsing

22. Use backpulsing model to predict optimal backpulsing conditions
23. Perform optimal backpulsing experiments with oily wastewater
24. Perform optimal backpulsing experiments with bacterial suspensions
25. Perform optimal backpulsing experiments with bentonite suspensions
26. Compare results to theory and to goals for permeate flux and quality
27. Perform economic analysis
28. Submit Final Report

4.1 Waters Containing Oil Droplets

Two alpha-alumina ceramic membranes (0.2- and 0.8-μm pore sizes) and a surface-modified polyacrylonitrile (PAN) membrane (0.1-μm pore size) were tested with an oily water containing various concentrations of heavy crude oil droplets of 2-10 μm diameters (Mueller et al., 1997). Despite significant fouling and flux decline (Figure 1), the membranes always produced a high-quality permeate containing less than 6 ppm oil (Table 2). Increased oil concentrations in the feed decreased the long-term flux, whereas the crossflow rate, transmembrane pressure, and temperature had relatively little effect on the long-term flux. Typical long-term flux values for membranes at 250 ppm oil in the feed are approximately 30-40 $L/m^2 \cdot h$ ($8\text{-}11 \times 10^{-6}$ m/s).

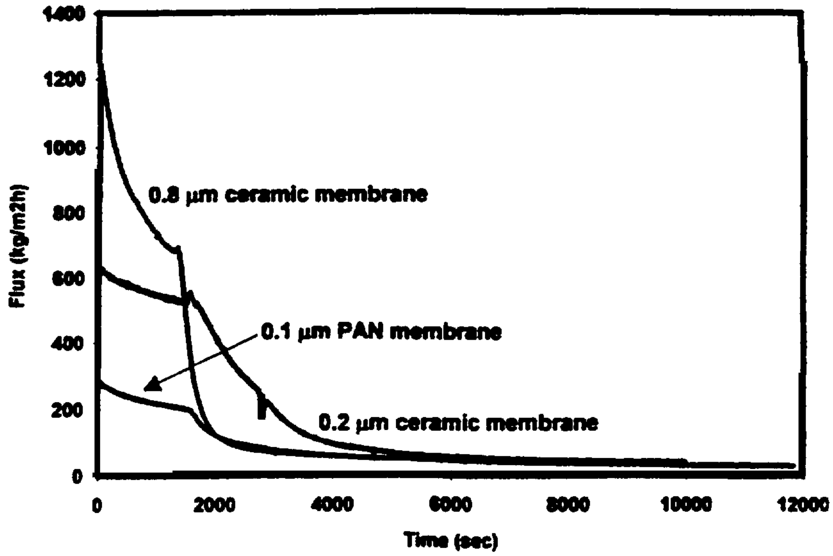

Figure 1. Mass flux of permeate versus time for crossflow microfiltration of 250 ppm heavy crude oil in water at 40°C, with a transmembrane pressure of 10 psi (69 kPa) and an average feed velocity of 0.24 m/s, for three different tubular membranes.

Table 2. Steady-state permeate flux and concentration of total oil and grease in the permeate for cross microfiltration of wastewater contaminated with oil at 40°C

The report values are the mean plus and minus the 90% confidence intervals for three repetitions [1 psi = 6.9 kPa, 1 mL/min = 1.7×10^{-8} m^3/s, 1 $L/m^2 \cdot h$ = 2.8×10^{-7} m/s]

Membrane	Pressure (psi)	Feed Rate (mL/min)	Feed Concentration (ppm)	Permeate Concentration (ppm)	Permeate Flux ($L/m^2 \cdot h$)
Ceramic	10	550	250	3.7 ± 1.8	42 ± 18
(0.2 μm)	20	550	250	5.1 ± 0.8	21 ± 3
	10	2100	250	4.6 ± 0.2	32 ± 12
	10	550	1000	5.8 ± 0.9	25 ± 5
Polymeric	10	4900	250	1.8 ± 0.6	34 ± 2
(0.1 pm)	20	5000	250	0.5 ± 0.2	32 ± 3
	10	4900	1000	0.9 ± 0.3	7 ± 2
Ceramic	10	550	250	0.4 ± 0.1	33 ± 6
(0.8 pm)	20	550	250	0.6 ± 0.3	40 ± 22
	10	2100	250	1.8i0.5	46 ± 6
	10	550	1000	1.5i0.2	26 ± 11

The fouling mechanisms were identified with the aid of resistance models, in which the shape of resistance-versus-time curves indicates whether membrane fouling is internal (concave-up) or external (concavedown). Both the 0.2- and 0.8-μm ceramic membranes exhibited internal fouling followed by external fouling, whereas external fouling characterized the behavior of the 0.1 -μm PAN membrane from the beginning of filtration (Figure 2). Examination of the external fouling layer showed a very thin, hydrophobic oil layer adsorbed to the membrane surface. This oil layer made the membrane surface hydrophobic, as demonstrated by increased water-contact angles. The oil layer proved resistant to removal by hydrodynamic (shear) methods, so it had to be extracted using tetrachloroethylene. Based on the results of IR analysis, the average thickness of oil at the end of the experiment was estimated at 61 μm for the 0.2-μm ceramic membrane and 30 pm for the 0.1-μm PAN membrane. These measurements are in good agreement with the predicted thicknesses from a simple mass balance in which it is assumed that all of the rejected oil is retained on the membrane and does not flow to the filter exit.

The membranes were also tested using backpulsing. Figure 3 gives the flux versus time for two different backpulsing cycles using the ceramic membrane with a 0.8-μm nominal pore size. The net flux was maintained nearly constant for the duration of each experiment, and a value of 320 $L/m^2 \cdot h$ (8.9×10^{-5} m/s) was achieved for a backpulse duration of 2 s and a forward filtration time between backpulses of 60 s. The corresponding long-term flux in the absence of backpulsing is less than 100 $L/m^2 \cdot h$ (2.8×10^{-5} m/s). Figure 4 shows a plot of the net flux with backpulsing versus the time of forward filtration between backpulses of duration 0.5 s, for a dilute feed stream of 60 ppm oil in water, using a ceramic tubular membrane with a 0.2-μm nominal pore size. A long-term flux of 70 $L/m^2 \cdot h$ (1.9×10^{-5} m/s) was observed in the absence of backpulsing, whereas an optimal net flux of 1260 $L/m^2 \cdot h$ (3.5×10^{-4} m/s) was obtained for a forward filtration time of 3 s between backpulses. For all cases, the total hydrocarbon content in the permeate was only 2-3 ppm.

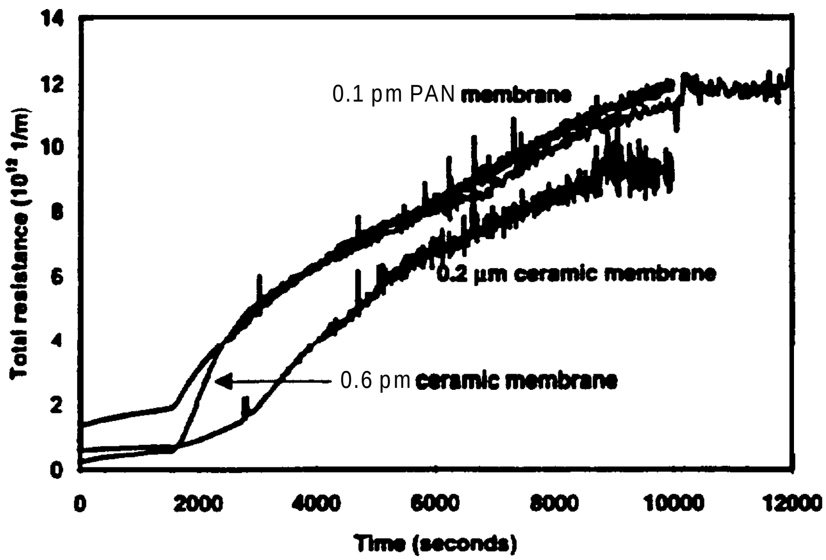

Figure 2. Total resistance of a fouled membrane versus time for crossflow microfiltration of 250 ppm heavy crude oil in water at 40°C, with a transmembrane pressure of 10 psi (69 kPa) and an average feed velocity of 0.24 m/s, for three different tubular membranes.

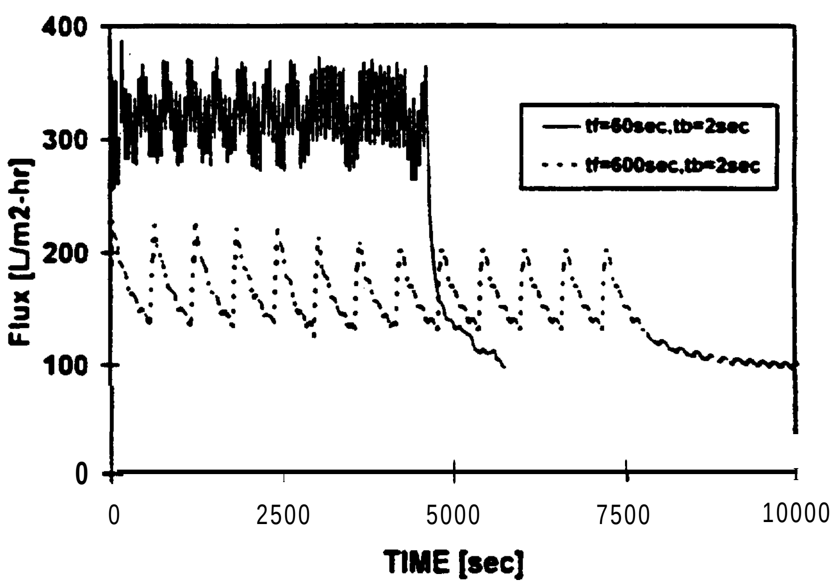

Figure 3. Forward flux versus time during backpulsing of 400 ppm light crude oil in water at 20°C, with a forward transmembrane pressure of 15 psi (103 kPa), backpulse duration of 2 s, and an **average** feed velocity of 4 m/s, using a 0.8-μm tubular ceramic membrane. For the cycle with a forward filtration duration of 60 s, the reverse transmembrane pressure was 10 psi; for the cycle with a forward filtration duration of 600 s, the reverse transmembrane pressure was 5 psi.

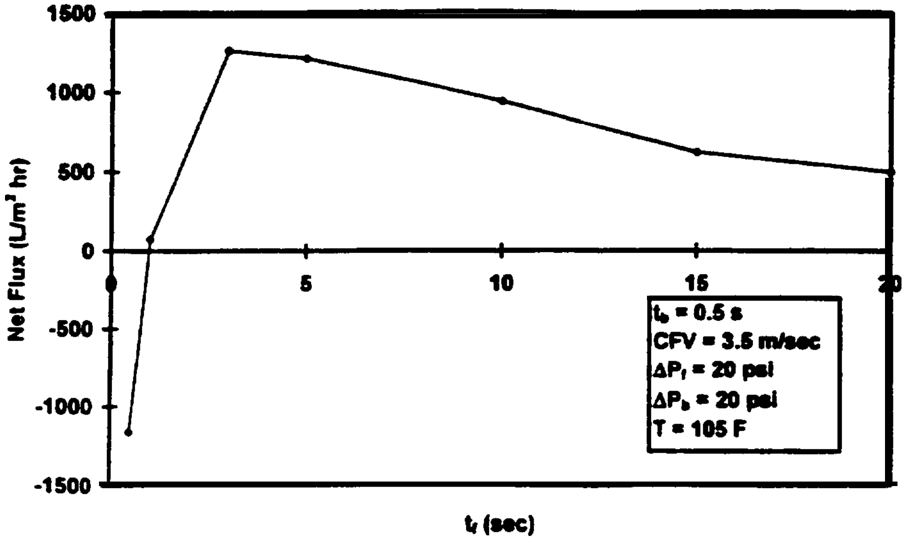

Figure 4. Net flux with backpulsing, versus duration of forward filtration, for oily wastewater (60 ppm heavy crude oil in the feed) feeding through a 0.2-μm ceramic tubular membrane. Backpulse duration 0.5 s; forward and reverse transmembrane pressures equal at 20 psi (138 kPa); average feed velocity 3.5 m/s.

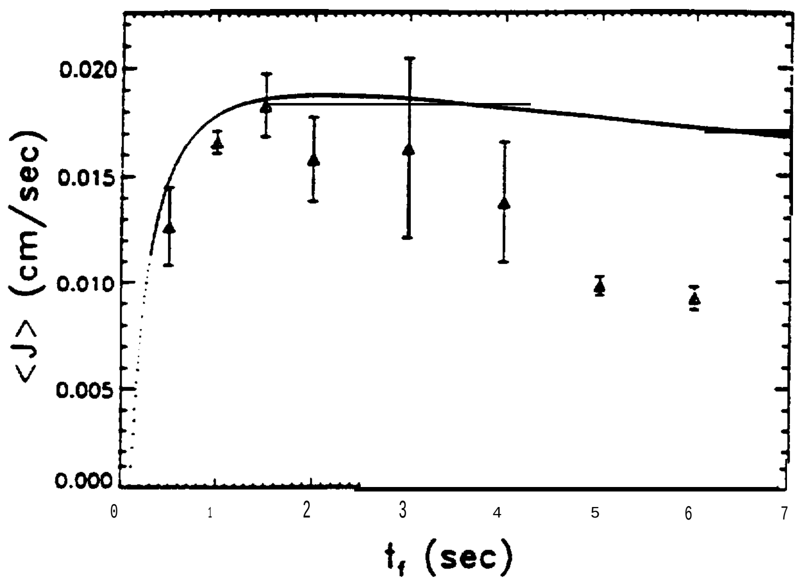

Figure 5. Net flux with backpulsing, versus duration of forward filtration for a yeast suspension (0.78% cells by volume in the feed). Backpulse duration 0.1 s; forward and reverse transmembrane pressures equal at 5 psi (34 kPa). The symbols are averages from three repeated experiments, the error bars are plus and minus one standard deviation, and the curve shows the theoretical flux (Redkar et al., 1996).

4.2 Waters Containing Microorganisms

The experiments with microorganisms were initially performed using Saccharomyces cerevisiae yeast cells in water, since yeast is readily available in powdered form which can be rehydrated for use as needed. In addition, our laboratory has had considerable experience with microfiltration of yeast (Redkar and Davis, 1993, 1995). Since yeast cells are large (4-5 μm in diameter) relative to membrane pore sizes, only external fouling was observed. Rapid backpulsing experiments used yeast suspended in deionized water, a flat-sheet crossflow microfiltration module, and cellulose acetate membranes with a 0.07-μm average pore diameter. The optimum forward filtration times were found to be 1.5, 3, and 5 seconds, respectively, for backpulse durations of 0.1, 0.2, and 0.3 seconds (Table 3, Figure 5). Both theory and experiment gave net fluxes with backpulsing of about 85% of the clean membrane flux (2.2×10^{-4} m/s = 790 L/m²·h), whereas the long-term flux in the absence of backpulsing is nearly an order of magnitude lower (2.6×10^{-5} m/s = 94 L/m²·h). The data fall below the theory at longer forward filtration durations, because the backpulse duration is too short to remove all of the yeast deposit.

Table 3. Net permeate flux at various backpulse durations for crossflow microfiltration of a yeast suspension

[Yeast cells 0.78% by volume; forward and reverse transmembrane pressures equal at 5 psi (34 kPa). Forward filtration duration was optimal for each backpulse duration. The clean membrane water flux is 790 $L/m^2 \cdot h$, whereas the long-term flux in the absence of backpulsing is 94 $L/m^2 \cdot h$ ($1\ L/m^2 \cdot h = 2.8 \times 10^{-7}$ m/s)]

Backpulse duration t_b (s)	Theoretical		Observed	
	Permeate flux ($L/m^2 \cdot h$)	Fotward-flow duration t_f (s)	Permeate flux ($L/m^2 \cdot h$)	Forward-flow duration t_f (s)
0.1	680	2.1	650	1.5
0.2	650	3.1	680	3.0
0.3	610	3.8	680	5.0

Later experiments were performed using *Escherichia coli* bacterial cells in water and a flat-sheet cellulose acetate membrane with a 0.22-μm nominal pore size. The pure water flux for a clean membrane at 10 psi (69 kPa) transmembrane pressure was 4700 $L/m^2 \cdot h$ (1.3×10^{-3} m/s). In the absence of backpulsing, a suspension of 0.01 g bacteria/g (wet weight basis) caused a severe flux decline, dropping to a long-term value of only 12 $L/m^2 \cdot h$ (3.2×10^{-6} m/s) (Figure 6). However, a backpulse duration of 1 s resulted in a maximum net flux of 50 $L/m^2 \cdot h$ (1.4×10^{-5} m/s) (Figure 7). By using more rapid backpulsing, the maximum net flux was raised to 140 $L/m^2 \cdot h$ (3.8×10^{-5} m/s) for a backpulse duration of 0.1 s (Figure 8).

43 Waters Containing Particulates

Crossflow microfiltration is commonly used as a pretreatment step to remove particulates from wastewater prior to nanofiltration or reverse osmosis. A review of the literature on this process led to a study of the effectiveness of combining crossflow microfiltration with rapid backpulsing as a pretreatment strategy using model aqueous suspensions of bentonite particles. The bentonite particle size distribution, as measured by a Coulter multisizer, is shown in Figure 9. Most of the particles have diameters in the range of l-10 μm.

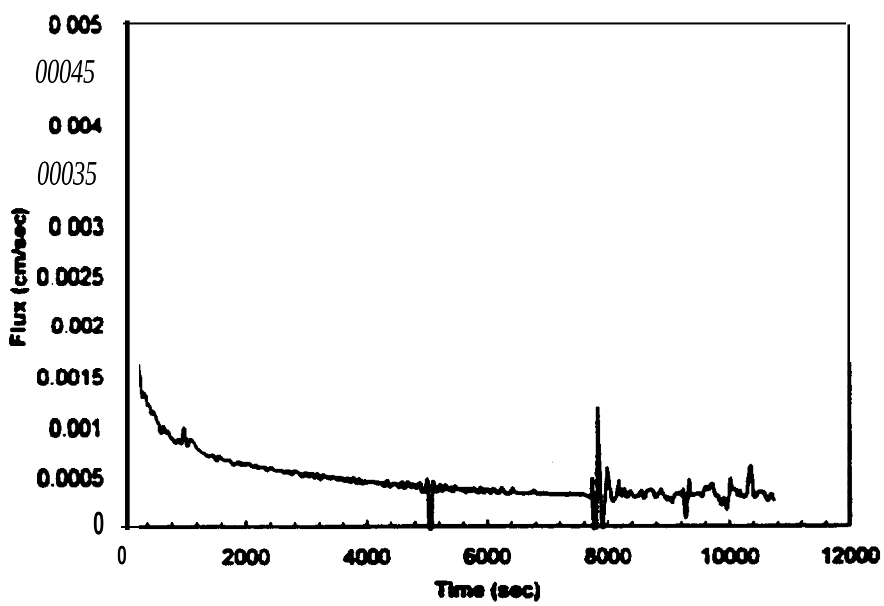

Figure 6. Permeate flux versus time for crossflow microfiltration without backpulsing of a 1% (wet weight) bacterial suspension in water at a transmembrane pressure of 10 psi (69 kPa) and an average feed velocity of 0.24 m/s, using a0.2-μmcellulose-acetate flat-sheet membrane; the initial flux of 0.13 cm/s is off the scale.

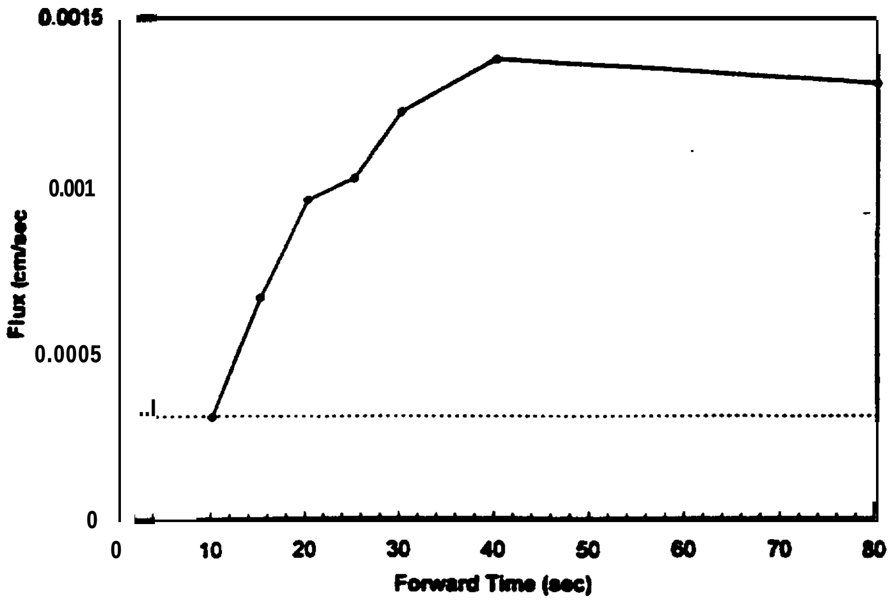

Figure 7. Net permeate flux with backpulsing versus duration of forward filtration for crossflow microfiltration of a 1% (wet weight) bacterial suspension in water at forward and reverse transmembrane pressures of 10 psi (69 **kPa)**, backpulse duration of 1 s, and average feed velocity of 0.24 m/s, using a **0.2-μm** cellulose-acetate flat-sheet membrane.

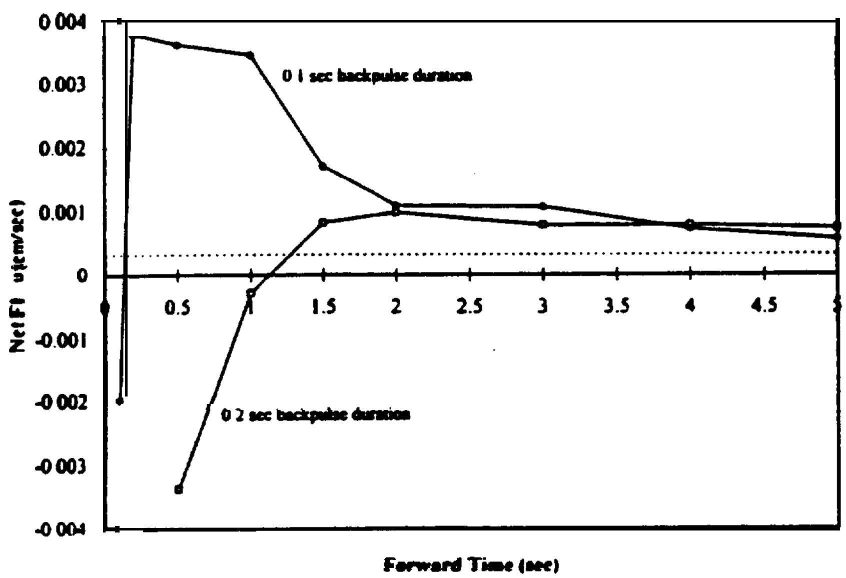

Figure 8. Net permeate flux with backpulsing versus duration of forward filtration for crossflow microfiltration of a 1% (wet weight) bacterial suspension in water at **forward** and reverse transmembrane pressures of 10 psi (69 **kPa)**, backpulse durations of 0.1 s and 0.2 s, and an average feed velocity of 0.24 m/s, using a **0.2-μm** cellulose-acetate **flat-**sheet membrane.

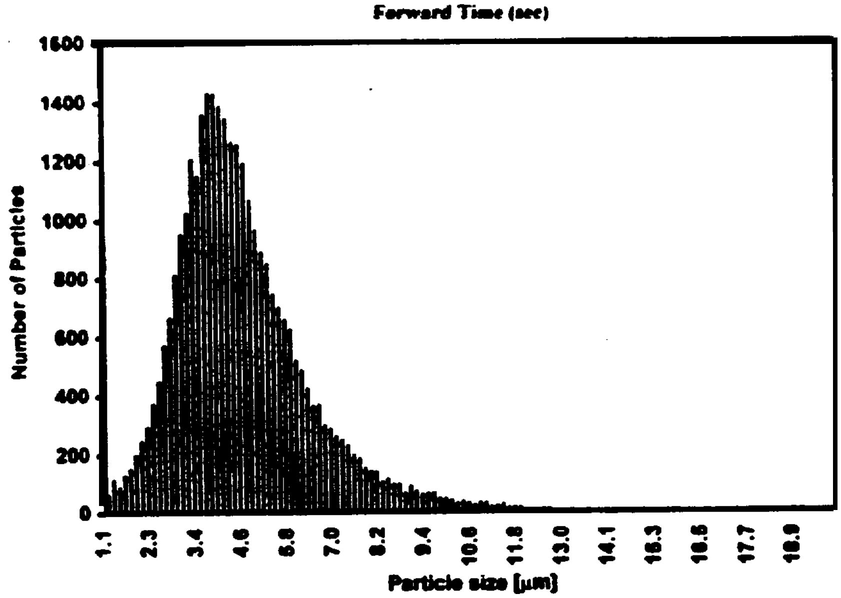

Figure 9. Size distribution, expressed as particle count versus particle diameter, of bentonite particles suspended in water.

Microfiltration was performed on aqueous bentonite suspensions using a tubular ceramic membrane with a 0.8-μm nominal pore diameter. Because of the large pore size, a very high water flux of 2,900 L/m^2·h (8.2×10^{-4} m/s) occurred at 20 psi (138 kPa) transmembrane pressure, but the permeate flux for a 0.2 g/L bentonite suspension quickly declined to 440 L/m^2·h (1.2×10^{-4} m/s) in the absence of backpulsing (Figure 10). Using backpulses with a duration of 0.5 s led to a dramatic improvement in the net flux (Figure 10), and a maximum net flux of 2300 L/m^2·h (6.4×10^{-4} m/s) was achieved at an optimum forward filtration duration of 5 s between backpulses (Figure 11). Interestingly, increasing the backpulse pressure to 20 psi (138 kPa) led to lower net fluxes (Figure 12), with an optimum flux of only 970 L/m^2·h (2.7×10^{-4} m/s). Presumably, the decrease was due to an increased loss of permeate during backpulsing. Additional results are shown in Figure 13 for a shorter backpulse duration of 0.2 s, which gave a slightly higher optimal flux (1300 L/m^2·h = 3.6×10^{-4} m/s). At a lower feed concentration of 0.04 g/L, the optimal net flux improved to 2220 L/m^2·h (6.2×10^{-4} m/s), which is nearly a 10-fold increase in the long-term flux in the absence of backpulsing and almost equal to the clean-water flux of 2300 L/m^2·h (6.4×10^{-4} m/s). For the more dilute suspension, the optimal forward filtration between backpulses is higher (Figure 14), due to the slower buildup of the fouling layer. In all cases the permeate was very clean, with a turbidity index of less than 2 NTU.

4.4 Optimization Theory of Rapid Backpulsing

Rapid backpulsing to reduce membrane fouling during crossflow microfiltration and ultrafiltration was studied theoretically by solving the convection-diffusion equation for concentration polarization and depolarization during cyclic operation with transmembrane pressure reversal. For a fixed duration of reverse filtration, there is a critical duration of forward filtration which must not be exceeded if the formation of a cake or gel layer on the membrane surface is to be avoided. The theory also predicts an optimum duration of forward filtration which maximizes the net flux, since backpulsing at too high a frequency does not allow for adequate permeate collection during forward filtration relative to that lost during reverse filtration, whereas backpulsing at too low a frequency results in significant flux decline due to cake or gel buildup during each period of

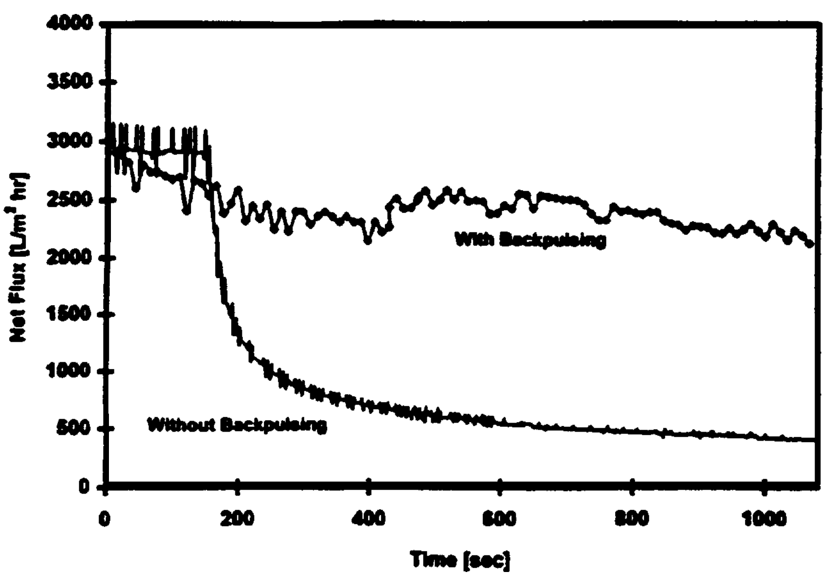

Figure 10. Permeate flux versus time for crossflow microfiltration of 0.2-g/L bentonite particles suspended in tap water at 27°C, with a transmembrane pressure of 20 psi (138 kPa) and an average feed velocity of 2.6 m/s, using a 0.8-μm ceramic tubular membrane with and without backpulsing. The backpulsing experiment involved a backpulse duration of 0.5 s, a 6-s interval between backpulses, and a reverse transmembrane pressure of 8 psi (55 kPa).

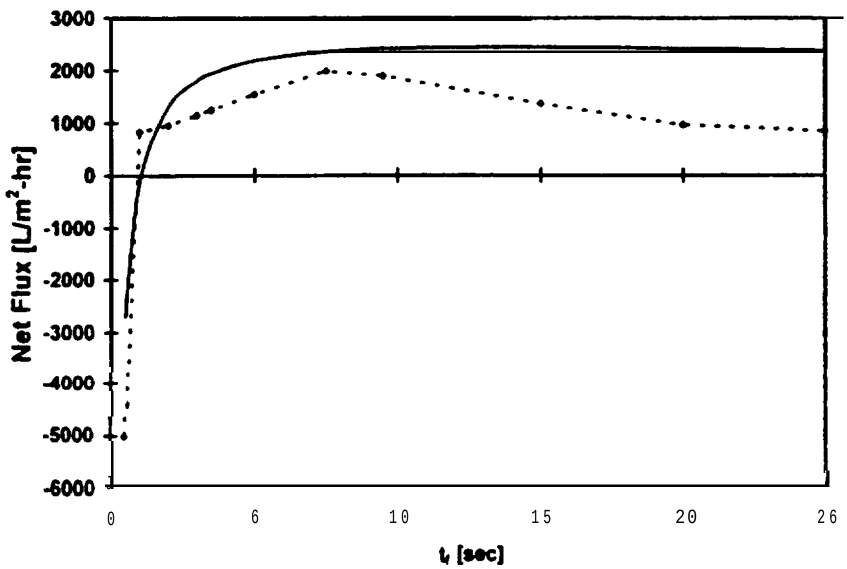

Figure 11. Net permeate flux versus the duration of forward filtration between backpulses for crossflow microfiltration of a 0.2-g/L bentonite suspension in water at forward and reverse trans-membrane pressure of 20 psi (138 kPa) and 8 psi (55 kPa), respectively, a backpulse duration of 0.5 s, and an average feed velocity of 2.8 m/s, using a O.&pm ceramic tubular membrane. The solid curve shows the flux projected by theory.

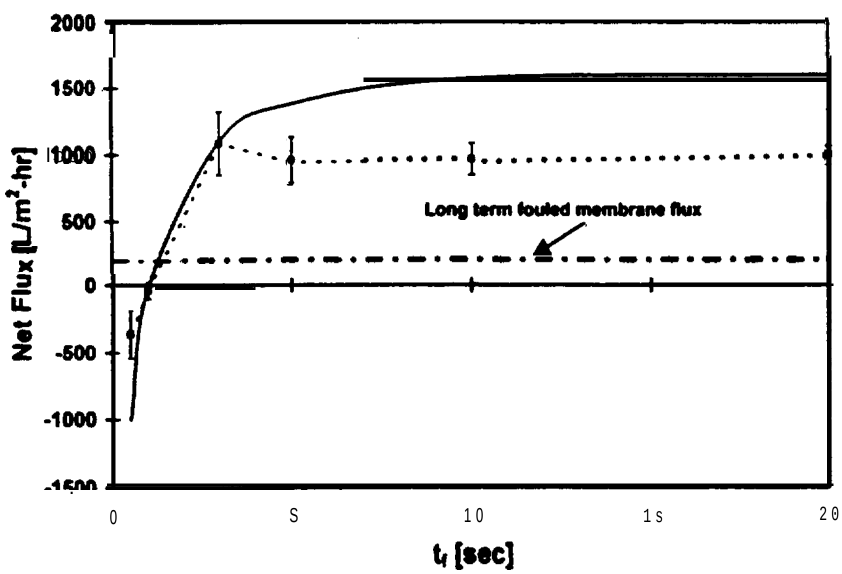

Figure 12. Net permeate flux versus the duration of forward filtration between backpulses for crossflow microfiltration of a 0.2-g/L bentonite suspension in water at equal forward and reverse transmembrane pressure of 20 psi (138 kPa), backpulse duration of 0.5 s, and an average feed velocity of 2.6 m/s, using a 0.8-μm ceramic tubular membrane. The solid curve shows the flux projected by theory.

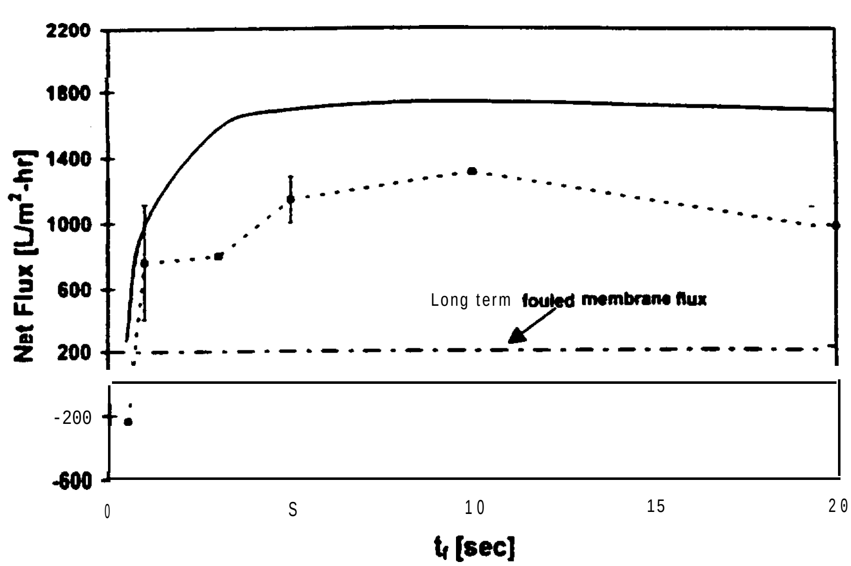

Figure 13. Net permeate flux versus the duration of forward filtration between backpulses for crossflow microfiltration of a 0.2-g/L bentonite suspension in water at equal forward and reverse transmembrane pressure of 20 psi (138 kPa), backpulse duration of 0.2 s, and an average feed velocity of 2.6 m/s, using a 0.8 μm ceramic tubular membrane. The solid curve shows the flux projected by theory

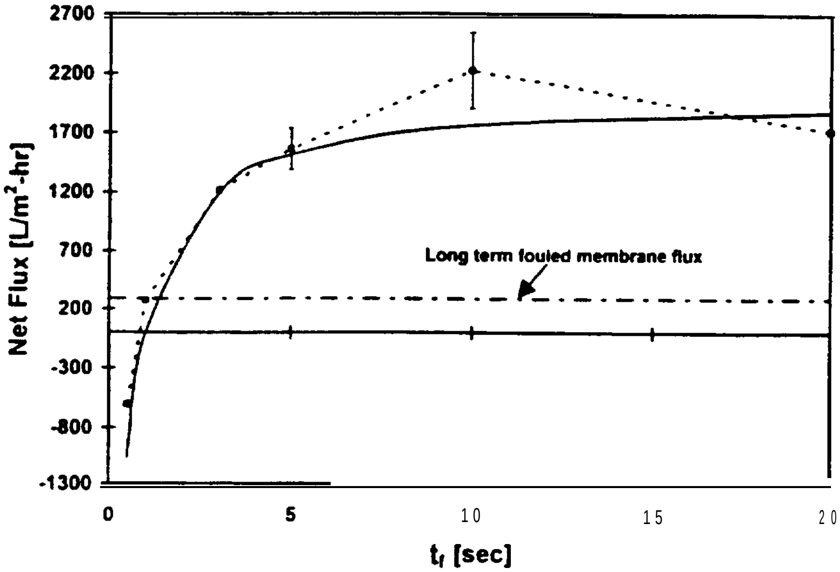

Figure 14. Net permeate flux versus the duration of forward filtration between backpulses for crossflow microfiltration of a 0.04-g/L bentonite suspension in water at equal forward and reverse transmembrane pressure of 20 psi (138 kPa), backpulse duration of 0.5 s, and an average feed velocity of 2.6 m/s, using a 0.8-μm ceramic tubular membrane. The solid curve shows the flux projected by theory.

forward filtration. In general, short backpulse durations, low feed concentrations, high shear rates, and high forward transmembrane pressures give the highest net fluxes, whereas the magnitude of the reverse transmembrane pressure has a relatively small effect (Redkar et al., 1996). Good agreement between theory and experiment was obtained for yeast suspensions and the low-concentration bentonite suspension, whereas bacterial suspensions and the more concentrated bentonite suspension had lower net fluxes than predicted. Presumably these latter solutions had incomplete cake removal during backpulsing, which would result from irreversible or adhesive fouling. No theoretical analysis was attempted for the oil experiments, because the combination of internal and external fouling observed is not described by the current theory (which is restricted to external fouling).

5. System Description

A schematic of the experimental setup is shown in Figure 15. The feed and backpulse reservoirs are pressurized using nitrogen cylinders with regulator valves. A pump circulates the feed through the retentate side of the membrane module. For the microbial suspensions, a peristaltic pump is used, whereas the setup for the oil and particulate suspensions employs a gear pump capable of generating higher flow rates. A Minitan-S flat-plate membrane module is used for the microbial suspensions, typically with cellulose acetate polymeric microfiltration membranes. This'module has nine parallel channels, each 0.4 mm high × 7 mm wide × 50 mm long. A tubular membrane module is used for the oil and particulate solutions, typically with an alpha-alumina ceranuic membrane. This module has a single tube of inside diameter 0.7 cm and length 21 cm.

Cellulose acetate membranes were used with the microbial suspensions because such membranes are widely used in biotechnology and water purification. The flat-sheet geometry was chosen because it has well-defined flow profiles and because the membrane can easily be removed and inspected or replaced. A ceramic membrane was used with the oil-water emulsions and bentonite suspensions because ceramic membranes can be operated under high temperature and withstand vigorous backpulsing. Because they are brittle, however, ceramic membranes are not currently available in flat sheets, and tubular geometries provide a well-defined alternative.

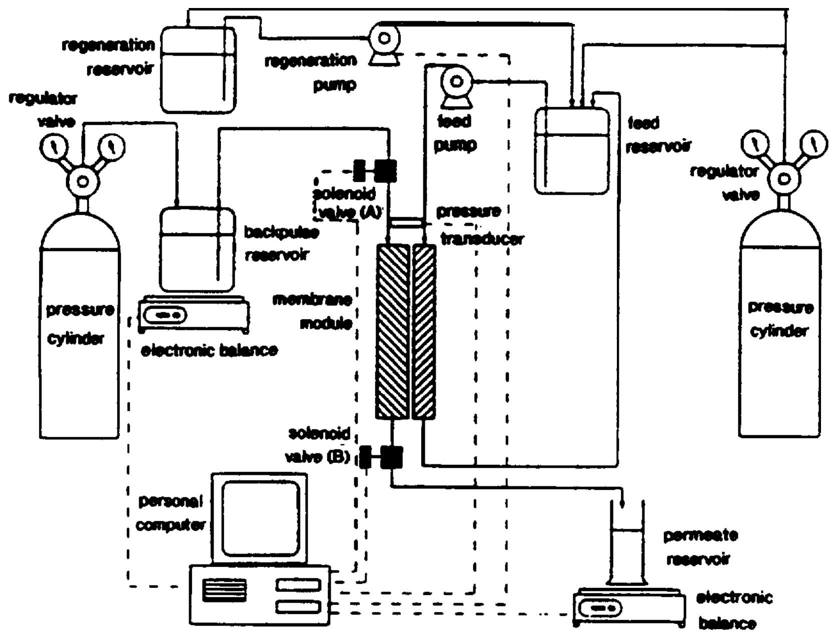

Figure 15. Schematic of the experimental apparatus for cross-flow microfiltration with back-pulsing.

Solenoid valves controlled by a microcomputer are used to switch between forward and reverse filtration. Valve B is open and valve A is closed during forward filtration, whereas valve A is open and valve B is closed during each backpulse. The switching time of the solenoid valves is about 50 ms, although the clock speed of the microcomputer limits the minimum backpulse time to about 100 ms (0.1 s).

Both the backpulse and permeate reservoirs sit on electronic balances interfaced to the microcomputer. These allow for the forward, reverse, and net permeate fluxes to be monitored. A regenerator pump controlled by the computer keeps the feed concentration constant by replacing the permeate fluid. The retentate is recycled to the feed reservoir (in practice, a fraction of the retentate would be collected as the concentrate stream).

6. Economic Analysis

This section evaluates the economic aspects of using crossflow microfiltration (CFMF) with flux enhancement by permeate backpulsing as a pretreatment step in reverse-osmosis plants. It compares this process with a conventional flocculation-multimedia filtration scheme. The study is based on experimental data obtained for bentonite suspensions using 0.8-μm ceramic membranes. The analysis shows that CFMF is not competitive with the conventional technique if run under normal conditions without backpulsing, except in small plants that have capacities less than 0.5 MGD.[2] Flux enhancement by backpulsing makes CFMF a viable option as a pretreatment technique (at least for the plant scales considered), roughly cutting in half the unit cost for treated water. Pretreatment costs range from $0.30/m³ ($1.13/1000 gal) for a 0.5-MGD plant, to $0.14/m³ ($0.53/1000 gal) for a 5-MGD facility, when net permeate fluxes of 1000 $L/m^2 \cdot h$ are maintained. On the other hand, treated

*Curiously MGD (millions of gallons per day) seems to be the unit of choice in the water treatment industry (1 MGD=157.7 m^3/hr) concerning plant capacity, while costs are often measured in terms of dollars per cubic meter (1 m^3 = 264 gal; 1000 gal = 3.78 m^3) of treated water.

water derived using a typical long-term forward filtration flux of 200 $L/m^2 \cdot h$, without backpulsing, costs at least $0.43/m³ ($1.63/1000 gal), which is above the cost for conventional treatment for all but small-capacity plants (<0.5 MGD).

6.1 Cost Model for Crossflow Microfiltration

The cost of water treatment can be apportioned into capital and operating costs. Capital costs represent the investment required to provide a given capacity for treated water production. This investment includes costs such as land, engineering, construction, and installations (civil, electrical, mechanical, etc.). Common practice for calculating the investment required for a unit volume of plant capacity (say in dollars per cubic meter of treated water) involves annualizing the initial cost incurred in erecting the facility and dividing by its intended capacity.

The operating costs include those expenses associated with plant operation and maintenance, such as energy consumption, membrane replacement, and labor. The cost of waste disposal is highly dependent on the nature of the feed and on plant location. For the water source considered herein, these costs were assumed to be negligible (though we note that conventional treatment with filter aids has higher waste disposal costs than does membrane filtration).

The present model is based on the plant schematic of Figure 16. Water from a natural source (such as a river, lake, or reservoir) is pumped through the CFMF module at a pressure of approximately 130 kPa. The concentrate is recycled through pump P2 at a rate dependent on the design cross-flow velocity. The design assumes use of tubular ceramic membrane elements contained in a large cylindrical module. The membrane module characteristics are contained in Table 4. A timer-controlled system permits switching operating mode from forward filtration to reverse filtration (for application of the permeate backpulse). As a generality it is assumed that the backpulse pressure is supplied by a pressurized reservoir, with a backpulse pump P_{bp} providing the necessary pressure difference.

Calculation of Capital Costs

Capital costs are subdivided into membrane costs and nonmembrane costs. The membrane costs include the initial cost of the membrane modules (C_{mem}). This is given by

$$C_{mem} = c_{mod} N_{mod} \tag{1}$$

where c_{mod} represents the cost of a single module and N_{mod} the number of required modules. The number of modules required is calculated from the expression (Pickering and Wiesner, 1993)

$$N_{mod} = Integer\left\{\frac{A_{reqd}}{A_{mod}} + 0.5\right\} \tag{2}$$

Here, A_{mod} is the membrane area of the membrane module in question and A_{reqd} is the required total membrane area for the design flow at a given net flux. In mathematical terms,

$$A_{reqd} = \frac{Q_{reqd}}{J_{net}} \tag{3}$$

where Q_{reqd} is the flow rate at the plant design capacity and J_{net} is the membrane net flux.

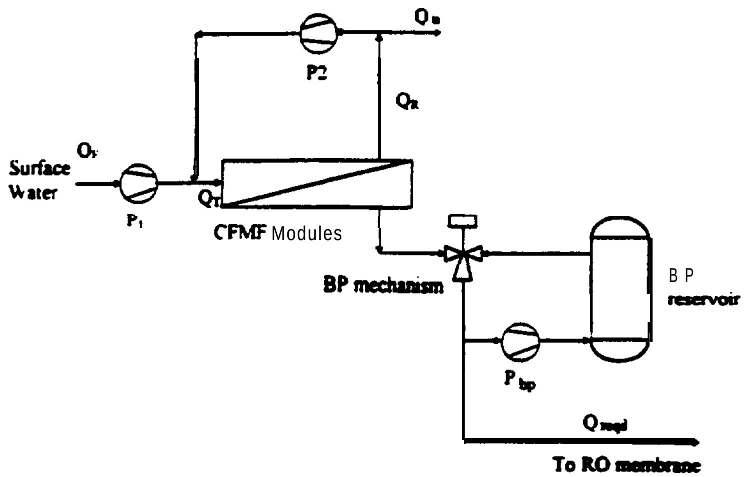

Figure 16. Schematic of crossflow microfiltration water pretreatment system.

Table 4. Design and cost parameters used in CFMF membrane economic model

Parameter	Symbol	Quantity
Diameter of membrane element	d_{elem}	4 mm
Number of elements per module	N_E	684
Length of module	L_{mod}	850 mm
Diameter of module	d_{mod}	320 mm
Cost of each module	c_{mod}	$14,600
Recommended membrane life	ML	8 years
Feed gauge pressure	P_1	130 kPa
Backpulse gauge pressure	P_{bp}	260 kPa
Backpulse duration	t_{bp}	0.2 s
Forward-flow duration	t_{ff}	5 s
Plant recovery	R	90%
Amortization period of plant	T	20 years

The nonmembrane costs ($C_{n\text{-}mem}$) are calculated from an expression compiled by Pickering and Wiesner (1993) using data from membrane and engineering and construction companies, which includes costs due to engineering and construction, and installations:

$$C_{n\text{-}mem} = \$1.50 \times 10^5 \, N_{mod}^{0.74} \tag{4}$$

The total capital cost is then amortized over the design life of the plant to yield an annualized capital cost. Thus,

$$CC = \frac{(C_{mem} + C_{n\text{-}mem})(AF)}{Q_{reqd}} \tag{5}$$

where CC represents the annualized plant capital cost, and the amortization factor AF is given by

$$AF = \frac{i(1+i)^T}{(1+i)^T - 1} \tag{6}$$

where T is the plant life and i the annual discount rate for capital investment.

Calculation of Operating Costs

The main operating costs to consider consist of energy (for pumping feed, recycling concentrate and pumping backpulse permeate), membrane replacement, maintenance, and labor.

Cost of Pumping Feed

The cost of pumping the feed from ambient pressure P_0 to P_1 is given by the simple expression

$$E_1 = \frac{(P_1 - P_0)\frac{Q_F}{\eta_1}}{Q_{reqd}} \tag{7}$$

where η_1 is the efficiency of the pump and the cost is expressed per unit volume of treated water. Q_F is the feed flow rate drawn into the treating plant and differs from the plant capacity Q_{reqd} by the quantity Q_w, as shown in Figure 16.

Cost of Pumping Recycle

The cost of recycling concentrate will depend directly on the pressure drop through the module on the concentrate side. This can be calculated by using the pressure-drop equation for flow through a tube:

$$P_{mod} = \frac{2 L_{mod}\, \rho CFV^2}{d_{elem}} f_f \tag{8}$$

where L_{mod} represents the length of the module, ρ is the density of the fluid, CFV the average cross-flow velocity within the membrane element, and d_{elem} the diameter of the membrane element. The Fanning friction factor f_f can be approximated by the expressions (Perry and Green, 1984)

$$f_f = \frac{16}{N_{Re}} \qquad N_{Re} < 4000 \tag{9}$$

$$f_f = \frac{0.079\ 1}{N_{Re}^{0.25}} \qquad N_{Re} > 4000 \tag{10}$$

Here, the Reynolds number, N_{Re} is based on the diameter of the element and the average cross-flow velocity, CFV. Once the pressure drop through the element is known, it is straightforward to calculate the energy required to pump the recycled concentrate stream as

$$E_2 = \frac{\Delta P_{mod}\frac{Q_R}{\eta_2}}{Q_{reqd}} \tag{11}$$

where Q_R is the recycle stream flow-rate and η_2 is the recycle pump efficiency. The recycle flow rate Q_R is found from the expression

$$Q_R = Q_T - Q_{reqd} - Q_w \tag{12}$$

where Q_T is the total flow rate entering the module and Q_w is the waste concentrate flow rate. Q_T is easily found by fixing the cross-flow velocity within the membrane elements:

$$Q_T = \mathrm{CFV}(A_{cs})\, N_{mod} N_E \tag{13}$$

with A, being the membrane element cross sectional area, N_{mod} the number of modules required and N_E the number of elements per module. The waste concentrate is fixed by specifying a plant recovery, R:

$$Q_w = Q_F\,(1-R) \tag{14}$$

A typical value of R = 0.90 (Pickering and Wiesner, 1993) is used herein.

Cost of Backpulsing

The energy cost of pumping the permeate for backpulse is given by

$$E_{bp} = \frac{(P_{bp} - P_0)\dfrac{Q_{bp}}{\eta_{bp}}}{Q_{reqd}} \tag{15}$$

where η_{bp} is the backpulse pump efficiency, and Q_{bp} is the **backpulse** flow rate. Q_{bp} is obtained by using the clean permeate flux, J_{bp} , *as*

$$Q_{bp} = J_{bp} A_{mod} N_{mod} \tag{16}$$

Since backpulsing only occurs for part of a cycle, the energy calculated in Equation 15 is weighted by the equation

$$E'_{bp} = E_{bp} \frac{t_{bp}}{t_{cycle}} \tag{17}$$

where t_{bp} is the backpulse duration time and t_{cycle} is the duration of the operating cycle (backpulse + forward filtration).

The total energy cost is then given by

$$C_E = C_{kWh}(E_1 + E_2 + E'_{bp}) \tag{18}$$

where C_{kWh} is the cost per kilowatt-hour in dollars.

Cost of Membrane Replacement

It is sensible to consider the membrane replacement cost as a variable (or operating) cost rather than a periodic investment of capital. It can be modeled as a constant operating cost by **assuming** that the membranes will be replaced at a fixed interval per the manufacturer's recommendation. Thus the cost of membrane replacement may be annualized over one replacement period by using the amortization factor of

$$AF_M = \frac{i_M}{(1 + i_M)^{ML} - 1} \tag{19}$$

Here, ML is the recommended membrane life (in years) and i_M is the annual discount rate for membrane replacement. The annual cost of membrane replacement is then calculated from

$$C_{MR} = \frac{C_{mod} N_{mod} AF_M}{Q_{reqd}} \tag{20}$$

Cost of Labor and Maintenance

The required labor is calculated from graphical data for man-hour requirements for fluid processing plants from Peters and Timmerhaus (199 1), and the cost of maintenance is taken as an annualized 1.5% of the nonmembrane cost (Owen et al., 1995). The values for cost and design parameters used can be found in Table 4.

The total cost of the treated water (in dollars per cubic meter of treated water) is found by adding the capital and operating costs discussed above.

6.2 Cost Model for Conventional Treatment

The costs for a conventional treatment plant have been calculated previously (Wiesner et al., 1994), based on the detailed model of Clark and Dorsey (1982) and Clark and Morand (198 1). The analysis is based on the basic schematic of the treatment plant of Figure 17. Waters of average turbidity (25-50 ntu) will require rapid-mixing/flocculation followed by gravity settling and multimedia filtration. 'Chlorine must be added for bacterial treatment, and sodium hydroxide and sulfuric acid are necessary for pH adjustment. Specific values of the range of design parameters are included in Table 5.

6.3 Analysis and Discussion

Capital and operating costs were calculated over a range of values of the permeate flux, at two different plant capacities (see Figures 18 and 19). It is observed that, with the commercially available modules, increasing the flux decreases the cost of water treatment, although the effect is small when the flux is above approximately 1000 $L/m^2 \cdot h$. As expected, the cost of treated water decreases with increasing plant capacity; however, the decrease becomes less significant for higher ranges of plant capacity.

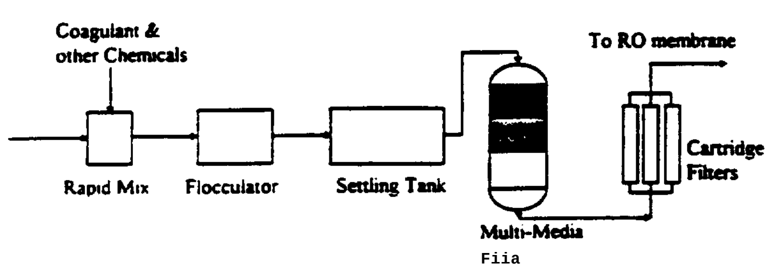

Figure 17. Schematic of conventional water pretreatment systems.

Table 5. Design and cost parameters used in conventional treatment plant model

Cost Parameters	
Amortization period	20 years
Annual interest rate	10%
Annual discount rate	8%
Labor	$0.07/hr
Electricity fuel	$0.9/gal
Natural gas	$0.001/ft^3

Cost of Chemicals	
Alum (dry form)	$250/ton
Soda ash (dry form)	$150/ton
Sulfuric acid (98% liq.)	$200/ton
Powdered act. carbon	$0.55/lb
Chlorine (liq.)	$250/ton

Design Parameters	
Rapid mix detention time	30 s
Rapid mix velocity gradient	6001s
Coagulant dosage	30 mg/L
Pump efficiency	70%
Flocculation detention time	20 min
Flocculation velocity gradient	600/s
Sediment basin geometry	Rectangular
Sediment basin overflow rate	540 gsd
Sed. basin sludge conc. (% solids)	0.7%
Filter media depth	12-18 in
Filter media diameter	0.55/1.15 mm
Uniformity coefficient	1.55/1.55
Filtration rate	5 GPM/ft^2
Backwash velocity	20 GPM/ft^2

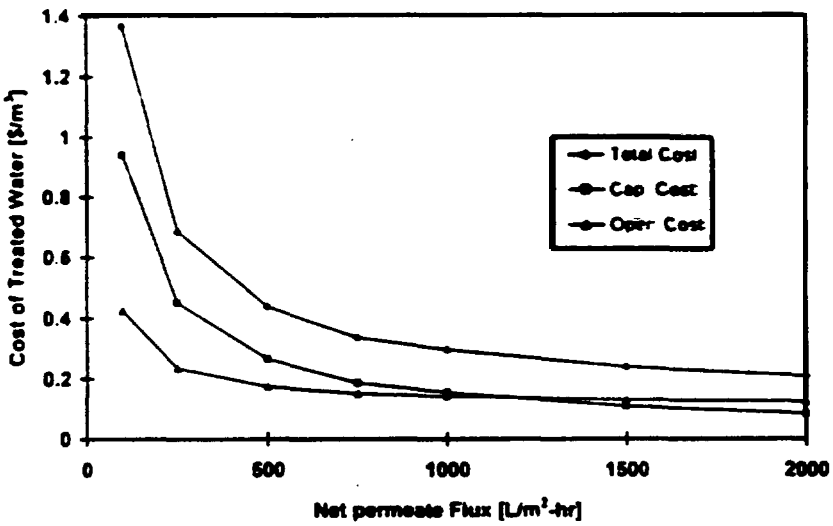

Figure 18. Water pretreatment cost versus permeate flux for a OS-MGD facility using membrane filtration with backpulsing.

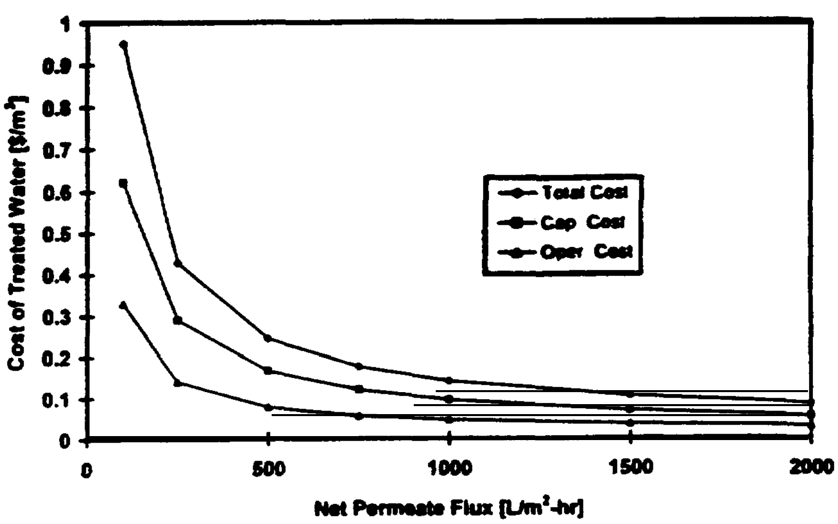

Figure 19. Water pretreatment cost versus permeate flux for a 5.0-MGD facility using membrane filtration with backpulsing.

If the membrane plant is operated at a flux of 200 $L/m^2 \cdot h$, pretreatment costs are competitive with those of small conventional plant installations (OS MGD and lower capacity) (Figure 20). If the membrane flux is raised to 1000 $L/m^2 \cdot h$ by backpulsing, membrane pretreatment becomes a viable alternative for even high-capacity systems, incurring costs that are approximately half of those from a large conventional plant (Figure 21).

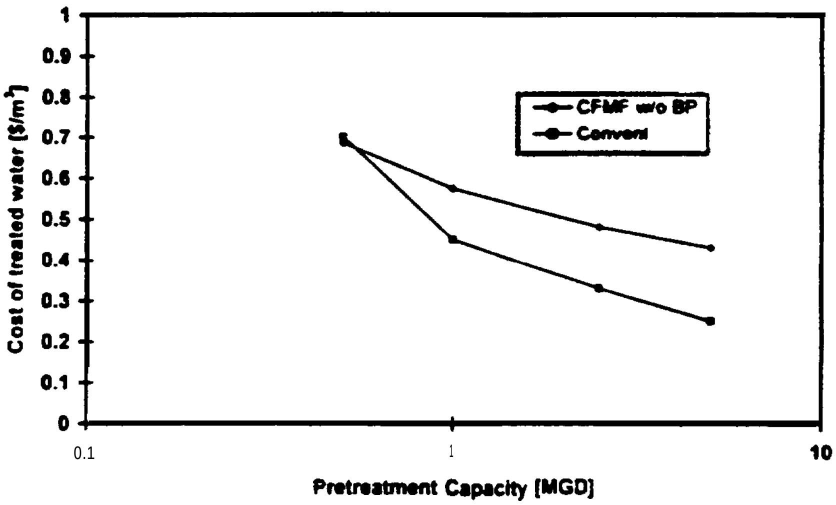

Figure 20. Comparison of water pretreatment cost versus capacity for membrane filtration without backpulsing (200 $L/m^2 \cdot h$ flux) and conventional filtration.

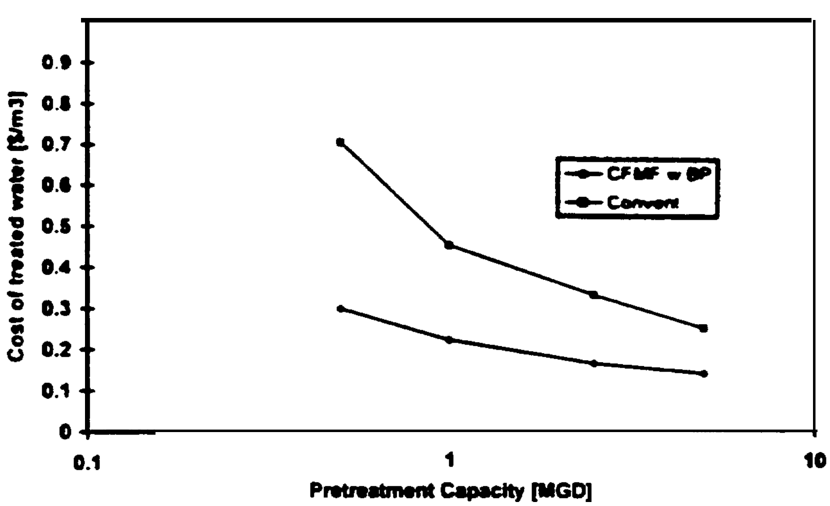

Figure 21. Comparison of water pretreatment cost versus capacity for membrane filtration with backpulsing (1000 $L/m^2 \cdot h$ flux) and conventional filtration.

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APPENDIX • DATA RECORD

A. Data for oily wastewater corresponding to **Figure** 1 and Tabte 2 for the 0.2 **µm** ceramic membrane.

Statistical Data Analysis of the Membralox alpha alumina ceramic membrane Experiment Results								
Experiment	Flux (Liter/m' hr)							
	Initial First	Final First	Flux	Flux	Flux	Fin Emuls	Final 2nd	Permeate
	Water	Water/Init	at	at	at	Flux/Init	Water	Oil
	Flux	Emul Flux				2nd Wat F	Flux	Conc.
	50 sec	1450 sec	2000 sec	4000 sec	8000 sec	8850 sec	10800 sec	(ppm)
MI - Baseline								
Run 1	656.6	560.5	480.5	86.5	40.8	27.5	35.2	2
Run 2	585.1	464.4	416.4	96.1	47.3	34.4	34.4	5.7
Run 3	640.6	537	352.3	106.6	77.7	63.7	55	3.5
Mean	627.4	520.6	416.4	96.4	55.3	41.9	41.5	3.7
Standard Deviation	37.5	50.1	64.1	10.1	19.7	19.2	11.7	1.9
90% Confidence Inter	35.6	47.6	60.9	9.5	18.7	18.3	11.1	1.8
M2 - High Pressure								
Run 1	776.7	777.3	432.4	50.5	31.3	23.1	20.7	3.2
Run 2	545.1	480.5	328.3	57.7	30.5	20.7	15.2	5.9
Run 3	728.7	713.4	609.2	53.7	27.2	18.4	18.3	4.3
Mean	883.5	857.1	458.8	54.0	29.7	20.7	18.1	5.1
Standard Deviation	1222	158.3	142.0	3.8	22	2.4	2.8	0.8
90% Confidence Inter	116.1	148.4	134.9	3.4	2.11	2.2	2.6	0.8
M3 - High CFV								
Run 1	258.2	199.4	160.11	81.71	57.7	43.2	NT	4.7
Run 2	239.4	200.2	183.41	78.91	44.8	34.4	NT	4.4
Run 3	312.3	232.2	198.6	81.71	48.5	17.8	NT	4.7
Mean	289.3	210.8	174.01	80.1	491.7	31.7	NT	4.8
Standard Deviation	38.2	18.71	21.3	2.81	7.0	13.0	NT	0.2
90% Confidence Inter	38.3	17.81	20.31	2.6	6.7	12.41	NT	0.2

A. ...(continued)

Experiment	Flux (Liter/m² hr)							
	Initial First Water Flux	Final First Water/Init Emul Flux	Flux at	Flux at	Flux at	Fin Emuls Flux/Init 2nd Wat F	Final 2nd Water Flux	Permeate Oil Conc.
	50 sec	1450 sec	2000 sec	4000 sec	6000 sec	8650 sec	10800 sec	(ppm)
M4 • Law Temp								
Run 1	268.8	262.9	253.4	151.4	112.4	98	NT	4.37
Run 2	284.4	231.1	231.9	132.3	99.6	106.7	NT	6.11
Run 3	233.4	247.2	247.2	144.3	86.9	69.3	NT	5.7
Mean	262.2	247.1	244.2	142.7	99.6	91.3	NT	5.4
Standard Deviation	26.1	15.9	11.1	9.7	12.8	19.6	NT	0.9
90% Confidence Inter	24.8	15.1	10.5	9.2	12.1	18.6	NT	0.9
Run 1	344.3	360.7	256.5	69.7	52.1	31.2	37.4	6
Run 2	3362	272.3	149.7	49.6	29.6	19.9	23.2	6.6
Run 3	264.3	270.1	169.6	49.7	29.6	23.9	22.3	4.7
Mean	319.6	301.0	192.0	56.3	37.1	25.0	27.6	5.8
Standard Deviation	31.4	51.7	56.8	11.6	13.01	5.7	8.5	1.0
90% Confidence Inter	29.8	49.1	53.91	11.0	12.3	5.4	8.0	0.9
M6 - Dead-End								
Run 1	352.3	280.5	240.5	118.5	76.9	63.2	NT	4
Run 2	288.3	224.41	126.7	76.61	56.1	33.6	NT	5.3
Run 3	344.7	208.2	166.6	124.1	100.9	74.6	NT	4.2
Mean	328.4	237.7	177.9	107.1	78.0	57.1	NT	4.5
Standard Deviation	35.0	37.9	57.7	24.8	22.4	21.2	NT	0.7
90% Confidence Inter	33.2	36.0	54.8	23.6	21.3	20.1	NT	0.7

B. Data for oily wastewater corresponding to Figure 1 and Table 2 for the 0.1 pm polymeric **(polyacrylonitrile)** membrane.

Statistical Data Analysis of the Zenon Surface Modified PAN membrane Experiment Results								
Experiment	Flux (Liter/m^2 hr)							
	Initial First Water Flux 50 sec	Final First Water/Init Emul Flux 1450 sec	Flux at 2000 sec	Flux at 4000 sec	Flux St 6000 sec	(Fin Emuls Flux/Init Water (2nd Wat F 8650 sec	(Final 2nd Water Flux 10800 sec	Permeate Oil Concentr (ppm)
Z1 - Baseline								
Run 1	314	220	123.7	57	42.3	32	30.1	2.5
Run 2	304.5	217.91	129.6	62.8	47.1	37.1	34.6	1.4
Run 3	218.11	172.71	115.81	[illegible]	47.11	34.21	[illegible]	1.5
Mean	278.91	203.51	123.01	60.2	[illegible]	34.4	[illegible]	1.8
Standard Deviation	52.8	26.7	[illegible]	3.0	2.8	2.6	2.6	0.6
90% Confidence Inter	50.2	25.4	6.6	[illegible]	2.6	2.41	2.5	0.6
Z2 - High Pressure								
Run 1	439.6	268.91	202.41	62.4	49.1	30.4	23.11	0.6
Run 2	404.7	270.9	131.5	[illegible]	[illegible]	[illegible]	24.81	0.7
Run 3	573.1	322.2	155.2	68.8	49.1	35.9	30.4	0.3
Mean	472.5	287.3	163.0	68.7	45.9	32.0	26.1	0.5
Standard Deviation	88.9	30.2	36.1	13.7	5.5	3.4	3.8	0.2
90% Confidence Inter	84.4	28.7	34.3	13.0	5.3	3.2	3.6	0.2
Z3 - High CFV								
Run 1								
Run 2								
Run 3								
Mean								
Standard Deviation								
90% Confidence Inter								

B. . ..(continued)

Experiment	Flux (Liter/m^2 hr)							
	Initial First Water Flux: 50 sec	Final First Water/Init Emul Flux 1450 sec	Flux at 2000 sec	Flux at 4000 sec	Flux at 6000 sec	Fin Emuls Flux/Init 2nd Wat F 8650 sec	Final 2nd Water Flux 10800 sec	Permeate Oil Concentr' (ppm)
Z4 - Low Temp								
Run 1	187.5	162.1	129	68.3	43	33.6	29.9	1.2
Run 2	236.5	167.9	134.7	70.3	45	27.8	21.3	3.6
Run 3	209	158.3	127.1	70.4	46.9	30.5	26.2	3.6
Mean	211.0	162.8	130.3	69.7	45.0	30.6	25.8	2.8
Standard Deviation	24.6	4.8	4.0	1.2	2.0	2.9	4.3	1.4
90% Confidence Inter	23.3	4.6	3.8	1.1	1.9	2.8	4.1	1.3
Z5 - High Oil Conc								
Run 1	439.6	404.3	367	51	16.5	9.22	7.47	0.8
Run 2	614.9	447.9	217.9	35.5	16.1	5.27	3.73	1.2
Run 3	580.9	461.7	337.9	27.9	18.1	6.08	5.5	0.6
Mean	545.1	438.0	307.6	38.1	16.9	6.9	5.6	0.9
Standard Deviation	93.0	30.0	79.0	11.8	1.1	2.1	1.9	0.3
90% Confidence Inter	88.3	28.5	75.1	11.2	1.0	2.0	1.8	0.3

C. Data for the oily wastewater corresponding to the 02 µm ceramic membrane in Figure 4.

LOW OIL CONCENTRATION (60 ppm) FOR 02 µm CERAMIC MEMBRANE

Time	Flux.	Cperm.
0.5	-1161	NT
1	75	NT
3	1280	2 ppm
5	1212	3.1ppm
10	936	2.0ppm
15	819	NT
20	498	NT

NT • not taken

For these runs the clean flux observed was of 2100 L/m2hr

D. Data for yeast suspensions corresponding to the 0.07 µm cellulose acetate membrane in Figure 5 and Table 3.

t_b(sec)	t & c)	Net Flux (cm/sec)
0.1	0.5	0.012632 ± 0.001842
0.1	1.0	0.016579 ± 0.000526
0.1	15	0.018289 ± 0.001447
0.1	2.0	0.015789 ± 0.001974
0.1	3.0	0.016250 ± 0.004145
0.1	4.0	0.013750 ± 0.002829
0.1	5.0	0.009803 ± 0.000461
0.1	6.0	0.009211 ± 0.000526
0.2	0.5	0.009737 ± 0.001842
02	1.0	0.015921 ± 0.000395
0.2	15	0.017368 ± 0.001316
0.2	2.0	0.017237 ± 0.000132
02	3.0	0.019868 ± 0.000921
02	4.0	0.018421 ± 0.001053
0.2	5.0	0.018421 ± 0.002237
0.2	6.0	0.015132 ± 0.004342
03	05	0.007150 ± 0.000789
03	1.0	0.013750 ± 0.000855
03	15	0.016711 ± 0.000263
03	3.0	0.017895 ± 0.002632
0.3	5.0	0.019342 ±0.001184
03	6.0	0.011447 ± 0.001053
03	8.0	0.009408 ± 0.000724

E. Data for bacterial suspensions corresponding to the 0.22 μm cellulose acetate membrane in Figure 7 and 8.

Normal Crossflow Experiments:

Date	Flux (cm/sec)
3/6/96	0.000331
3/6/96	0.000307
7/23/96	0.000309

Backpulsing Experiments:

(sec)	t_f(sec)	J_f(cm/sec)	J_b(cm/sec)	Net Flux (cm/sec)	Date
0.1	0.1	0520249	0.524133	-0.0019	7/22/96
0.1	0.2	0.258333	0502692	0.00465	7/17/96
0.1	0.2	0.259622	0.507969	0.00375	7/18/96
0.1	0.2	0.250886	0.491296	0.00349	8/12/96
0.1	0.2	0.252774	0.494069	0.00382	8/13/96
0.1	05	0.056566	0.261477	0.00355	7/5/96
0.1	05	0.102183	0.488892	0.00367	8/15/96
0.1	0.5	0.107375	0.512425	0.00407	8/16/96
0.1	1	0.057842	0.540656	0.00343	7/11/96
0.1	1	0.056964	0547966	0.00197	8/19/96
0.1	1	0.056692	0.539118	0.00252	8/20/96
0.1	1.1	0.044374	0.469185	0.00174	5/17/96
0.1	1.5	0.035835	0.512537	0.00156	5/15/96
0.1	2	0.027277	0.522767	0.001~	5/14/96
0.1	3	0.017824	0.501518	0.00107	5/10/96
0.1	4	0.014076	0.533722	0.00071	4/3/96
0.1	5	0.011157	0529369	0.00055	4/1/96
0.2	0.2	0.234044	0.255011	-0.0104	7/16/96
0.2	0.5	0.098023	0.256838	-0.0033	7/12/96
0.2	1.5	0.032908	0.238633	0.00096	7/3/96
0.2	1.5	0.0359 16	0.264415	0.00058	7/8/96
0.2	1.5	0.033709	0.245202	0.00089	7/9/96
0.2	2	0.026002	0.249241	0.00097	7/2/96
0.2	2	0.026336	0.254838	0.00077	7/29/96
0.2	2	0.025556	0.244604	0.00099	8/8/96
0.2	2	0.025342	0.243281	0.00092	8/9/96
0.2	3	0.017475	0.249404	0.00079	6/28/96
0.2	3	0.017426	0249115	0.00076	7/1/96
0.2	4	0.013326	0.250249	0.00077	6/27/96
0.2	5	0.011170	0.260342	0.00072	6/26/96
1	5	0.012296	0.064780	-0.0005	7/24/96
1	10	0.008600	0.082539	0.00031	3/8/96
1	10	0.006445	0.060912	0.00032	7/25/96
1	10	0.006506	0.060938	0.00037	8/2/96
1	15	0.005191	0.067094	0.00067	3/7/96
1	20	0.005131	0.082377	0.00096	3/11/96
1	25	0.003926	0.071469	0.00102	3/19/96
1	30	0.003111	0.055651	0.00121	3/19/96
1	40	0.003439	0.081318	0.00137	3/20/96
1	80	0.002152	0.066910	0.00130	3/20/96

F. Data for a bentonite suspension corresponding to the 0.8 mm ceramic membrane in Figures 12 and 13.

Experiments with high turbidity bentonite feed (C=0.2 g/L)

t_b=0.2 s

Tf (s)	Meas. flux [L/m²hr]	Std. Dev [L/m²hr]	90% C.I. [L/m²hr]	Perm. NTU[⊥]
0.5	-239	NT	NT	0.30
1	761	304	±354	0.90
3	800	NT	NT	1.56
5	1139	119	±138	1.70
10	1305	274	NT	1.71
20	953	6	±7	0.20

t_b=0.5 s

IT'f(s)	Meas. flux [L/m²hr]	Std. Dev [L/m²hr]	90% C.I. [L/m²hr]	Perm. NTU[⊥]
0.5	-364	187	±178	0.23
1	-39	65	±61	0.46
3	1087	250	±238	1.10
5	947	180	±171	1.24
10	961	123	±117	0.53
20	992	73	±70	0.26

-Average uncertartainty in the turbidity values was always less than 12%.
C.I. - Confidence interval
NT - Not enough repeats were performed for a formal uncertainty analysis

G. Data for a bentonite suspension corresponding to the 0.8 mm ceramic membrane in Figure 14.

Experiments with dilute bentonite feed (00.04 g/L)

Tf (s)	Meas. flux [L/m^{2}hr]	Std. Dev [L/m^{2}hr]	90% C.I. [L/m^{2}hr]	Perm. NTU[⊥]
0.5	-608	NT	NT	0.27
1	274	NT	NT	0.70
5	1214	NT	NT	1.50
10	1559	150	±175	0.90
20	2224	274	±318	0.40
40	1713	NT	NT	0.50

[⊥]Average uncertartainty in the turbidity values was always less than 12%.
C.I. - Confidence interval
NT - Not enough repeats were performed for a formal uncertainty analysis