Large-Scale Microfiltration:

Results of the Phase I and Phase II Tests

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Selected Abbreviations and Acronyms

A.S.	Activated sludge
BOD	Biochemical oxygen demand (5-day test)
cBOD	Carbonaceous BOD (5-day test)
DAF	Dissolved air flotation
gpd/ft^2	Flux in gallons of flow per day per square foot of membrane surface area
GWRS	Groundwater Replenishment System
HPC	Heterotrophic plate count
MBR	Membrane bioreactor
MF	Microfiltration
OCWD	Orange County Water District
RO	Reverse osmosis
TOC	Total organic carbon
TSS	Total suspended solids
scfm	Standard cubic feet per minute
TKN	Total Kjeldahl nitrogen
VSS	Volatile suspended solids
WAS	Waste activated sludge

EXECUTIVE SUMMARY

Background

A microfiltration (MF) demonstration test unit from ZENON Environmental, Inc. (Zenon) capable of producing 40 gpm of permeate (treated water) was tested at OCSD in 1998 and 1999. Tests in 1996 using a 10 gpm pilot unit had shown that MF could produce a high quality secondary effluent, but questions about acceptable operating conditions and likely operating costs could not be answered due to various design limitations in the pilot equipment. The larger-scale demonstration unit was designed to address these questions. In addition, new membranes offering superior performance had become available from Zenon after the small-scale tests were completed, so the later tests explored these new membranes' performance.

The tests were conducted in two phases. The Phase I tests were conducted from June 15, 1998 to September 22, 1998, and primarily were meant to provide data to be used by OCWD in planning the GWRS project. The MF feed was unsettled mixed liquor from the Plant 1 A.S. effluent channel, and the membrane was Zenon's ADC membrane with a mean pore size of 0.08 µm. During these tests, the operating conditions were varied to investigate the MF's performance as a secondary clarifier with various reactor concentrations and flux rates.

Table ES-1 presents the operating targets for the three test conditions that were investigated in Phase 1.

Table ES-1. Phase I Operating Targets

Flux (gpd/ft^2)	Permeate Flow (gpm)	Reactor TSS (mg/L)	Sludge Age (hours)
13	40	1200	~4
20	40	2400	~8
28	30	1200	~4

Approximately one-half of the MF permeate was fed into a reverse osmosis (RO) unit because the earlier tests had suggested that MF permeate without additional treatment might not be a suitable RO feed. OCWD personnel maintained the RO unit and regularly sampled and analyzed the MF and RO permeates.

Phase II of the study was conducted to test microfiltration of primary effluent in a membrane bioreactor (MBR). In an MBR, the membranes are immersed in an aerated tank of concentrated mixed liquor, so the MBR acts as both an aeration basin and a clarifier. Phase II used Zenon's OCP membranes, which have a 0.035-µm mean pore size.

Phase II tested four operating conditions, using two reactor TSS concentrations and two membrane fluxes. The reactor concentrations spanned a much wider range than in Phase 1, reflecting the MBR’s ability to operate with mixed liquor TSS concentrations substantially higher than in conventional A.S. processes.

Table ES-2 summarizes the Phase II test conditions.

Table ES-2. Phase II Operating Conditions

Condition	Target TSS Concentration (mg/L)	Flux (gpd/ft^2)
Low concentration, low flux	4,000	19
High concentration, low flux	12,000	19
Low concentration, high flux	4,000	29
High concentration, high flux	12,000	29

Phase I Results

The membrane flux is a measure of the amount of permeate that a specified amount of membrane will produce. The higher the flux, the less membrane surface area will be needed to achieve a given permeate flow rate. However, higher fluxes generally also require higher operating pressures (or higher vacuum levels with the Zenon membranes) to force more water through the membrane’s pores.

Figure ES-1 shows the effect of membrane flux on the vacuum pressure required to maintain the permeate flow rate in Phase I. A clear positive correlation was observed. As the average flux approximately doubled (from 13 to 28 gpd/ft^2), the average vacuum approximately tripled (from 5.6 to 15.4 "Hg).

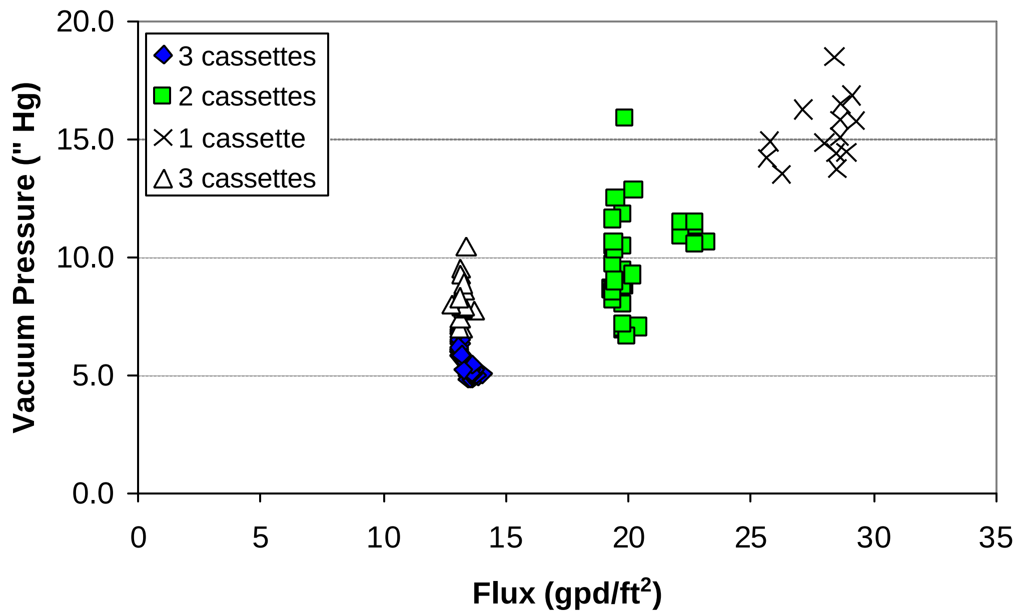
As the pressure increases, the power required for pumping the permeate at a fixed flow rate also increases. Table ES-3 compares the theoretical pumping power requirements for the four Phase 1 flux conditions. Using the low-flux/low-vacuum condition as the baseline (relative power = 1.0), the pumping power increases for the other conditions and is nearly doubled (to a relative value of 1.9) for the highest flux condition, even though the flow rate in that condition is 30 percent less than in the baseline condition.

Table ES-3. Comparison of Power Requirements for Phase I Conditions

Average Flux (gpd/ft^2)	Average Vacuum ("Hg)	Average Flow Rate (gpm)	Relative Pumping Power Needed
13	5.6	42	1.0
20	9.8	42	1.8
28	15.4	29	1.9
13	8.4	41	1.5

The operating vacuum also changed between membrane cleanings. At each flux the vacuum increased over time, indicating fouling that reduced the membrane's permeability. The permeability was recovered by weekly maintenance cleanings using a bleach solution.

Figure ES-1. Phase 1: Effect of Flux on Vacuum Pressure



A variety of chemical and physical analyses were performed on the MF feed, reactor mixed liquor, and permeate. Table ES-4 summarizes many of the test results for the feed and permeate during Phase I. The complete results are presented in Appendices B, C, and D.

Table ES-4. Phase I: Selected Chemical and Physical Test Results (Average Values)

	Feed	Permeate	Removal
Ammonia (mg/L)	17	13	24%
TKN (mg/L)	77	15	81%
Nitrite (mg/L)	0.36	2.92	-----
Nitrate (mg/L)	<1.77	2.66	-----
Alkalinity (mg/L)	272	214	21%
TSS (mg/L)	587	<0.6	>99%
cBOD (mg/L)	-----	<4.0	-----
TOC (mg/L)	46.8	7.59	80%
pH	-----	7.77	-----
Turbidity (NTU)	-----	0.41	-----

The permeate analyses indicated excellent rejection of suspended material by the membranes. The permeate TSS concentration often was less than the detection limit (0.4 mg/L), indicating 99.9 percent TSS removal from the feed. Similarly, permeate cBOD readings most often were below the detection limit (4.0 mg/L), and the feed TOC was reduced 80 percent. These results were not affected by the reactor TSS concentration.

Feed and permeate samples were taken for bacteriological analyses: total coliform, fecal coliform (feed only), HPC, and E. coli (permeate only). The results are shown in Table ES-5.

Table ES-5. Phase I: Bacteriology Results Summary

	Feed			Permeate			Log Removal
	Number of Samples	Range (10^6)	Average (10^6)	Number of Samples	Range	Average	
Total Coliform (MPN/100 mL)	39	2.2-50	13.2	14	4-130	23	5.8
Fecal Coliform (MPN/100 mL)	39	0.2–7.0	2.4	----	----	----	----
HPC (CFU/mL)	39	1.3-19	6.6	38	400-86,000	25,000	2.4
E. Coli (MPN/100 mL)	----	----	----	38	<1-10	<1	----

Substantial rejection of microbiological organisms by the membranes was evident. The total coliform results indicated an average of 5.8 log removal, and the HPC results, which look at a wider variety of organisms, showed an average of 2.4 log removal. Concentrations of E. coli in the permeate averaged less than 1 MPN/100 mL.

In addition to the routine bacteriology sampling, twice-daily virus (coliphage) sampling was conducted for five days. Of the nine samples taken, seven samples showed virus concentrations at or below the detection limit (4.5 PFU/100 mL). These results indicated 3 to 4 log reductions in virus concentrations between the feed and permeate.

Phase II Results

Since Phase II looked at a wider range of reactor TSS concentrations than Phase I, any effects of concentration on the operating vacuum pressure should be more evident in Phase II. The results indicated that the reactor concentration in the range tested was not an important factor in determining the operating vacuum.

The operating vacuum did increase at higher flux conditions; in fact, higher fluxes were associated with both higher initial vacuum levels and faster fouling. With a newly cleaned membrane, the vacuum at the high flux condition was 4 to 6 "Hg higher than at the low flux condition, and the fouling rates at the high flux conditions were 3 to 18 times faster than at the low flux conditions. This has important implications for a full-scale installation since the flux affects both the initial capital cost (the amount of membrane surface area needed) and the ongoing operating costs.

The chemical and biological test matrix used in Phase II differed somewhat from the Phase I matrix. TSS and VSS tests were performed daily, and other tests (BOD, TKN, turbidity, pH, alkalinity, and total coliform) were performed less frequently. Nitrate, nitrite, and TOC were not tested in Phase II.

The permeate TSS and VSS concentrations generally were less than 1.0 mg/L and often were below the detection limit (0.4 mg/L). Similarly, the permeate BOD concentration generally was at or below the detection limit of 4.0 mg/L. These permeate values were not affected by either the reactor TSS concentration or the membrane flux.

The reactor concentration had a considerable effect on the net amount of nitrification that occurred. At the high reactor TSS concentration, there was little difference between the feed (22 to 27 mg/L) and permeate (16 to 24 mg/L) ammonia concentrations, so minimal net nitrification had occurred. At the low reactor TSS concentration, very little ammonia (<1 to 3 mg/L) remained in the permeate, so substantial nitrification had occurred. These results were not affected by the membrane flux.

For all test conditions, the permeate turbidity did not change during the time between chemical cleanings. A relationship was seen between high reactor TSS concentrations and higher permeate turbidity, but the practical significance of this is questionable, since in all cases the permeate turbidity was less than 1 NTU.

The coliform rejection performance of the system was inconsistent. At its best, the system showed approximately 6 log coliform removal (to <10 MPN/100 mL), but some permeate samples had coliform concentrations exceeding 1000 MPN/100 mL. Various piping leaks and defective seals were discovered during and after the tests, so these may be responsible for the higher readings. It was noted that other agencies reportedly have experienced erratic bacterial removal with Zenon membranes, though, which might be due to small areas of delamination between the polymer coating and the substrate (although this explanation is speculative and has not actually been observed). Good coliform rejection should be achievable with intact membranes.

A large part of the operating costs for a Zenon MBR system is due to the power required for the aeration system. Air is used both for biological stabilization and for membrane scouring, and the scouring requirements dominate. This means that the MBR system always will use more air than a well-operated A.S. process accomplishing the same amount of treatment. In these tests, the aeration rate was kept essentially constant at about 25 cfm/module; this was the minimum value Zenon would consider for a full-scale installation. Depending on the test condition, this corresponded to specific air rates of 2500 to 4300 ft^3 per pound of cBOD removed from the feed. This was 3.3 to 5.6 times the air volume used in OCSD Plant 1's A.S. system.

Since OCSD tries to minimize nitrification, its air usage is low compared to more typical A.S. installations. In addition, since the air in an MBR could enter at a shallower depth than in an A.S. system, the operating cost difference would not necessarily correspond to the absolute air flow difference. Nevertheless, the aeration cost would be higher for an MBR than for A.S.

After the Phase II tests were completed, Zenon announced its development of an air cycling system in which the membranes are aerated intermittently rather than continuously. Zenon claims that this would reduce the air required for MF more than 50 percent when treating primary effluent (from 5 cfm/100 ft^2 of membrane as tested in Phase II to 2.3 cfm/100 ft^2 of membrane). The air cycling operation currently is being tested by OCWD. If it performs as promised, then Zenon's MF air usage will be much closer to the air usage in a typical A.S. system.

The combined effects of a shallower depth for the membranes and a reduced air flow with cycling were estimated. Depending on the operating conditions (reactor concentration and membrane flux) chosen, the MF blower power usage could be more or less than the existing A.S. power usage. The maximum MF cost savings occur with low-concentration/high-flux operation. With electricity valued at 6¢/kWh, MF could reduce OCSD’s electricity costs $172,000/year.

If MF were installed in place of conventional A.S., it also would affect the downstream solids handling processes (DAF and anaerobic digestion). Zenon has reported a waste sludge rate for a full-scale installation that is one-third less than OCSD's A.S. WAS rate. If this reduction were applied to OCSD's operations (based on FY 98-99 data), annual cost savings of about $363,000 in the total DAF and digestion operations would be realized.

Much of the interest in MF has been related to its role in the GWRS project, for which it is expected to be a key part of the complete treatment train. As GWRS continues into its second and third phases, the unique characteristics of MBRs may be increasingly important. By combining secondary treatment of high TSS mixed liquor and clarification into a single process, MBRs provide effluent stabilization while requiring much less land area than conventional A.S. processes. Also, microfiltration produces an effluent that can be fed directly to an RO process, which eliminates the intermediate filtration that is needed for A.S. effluent. Finally, as mentioned previously, the results of this project indicated that energy costs can be less with an MF MBR (at appropriate operating conditions) than with A.S.

There is a major regulatory barrier that must be addressed before MF MBRs could be used in place of conventional secondary processes in water reclamation. In the regulators’ concept of “barriers” for reducing the amounts of microbiological organisms in the water, each processing step is seen as a barrier to accidental process failure and contamination of the product water. By combining secondary stabilization and clarification into one step, there is one less barrier in the process train. Even though the coliform and virus removal with MF exceeds that of A.S. with clarification, regulators have not yet accepted membrane systems as equivalent to conventional secondary treatment systems.

MF also has other possible applications for OCSD. Ocean discharge limits might be met by blending the high quality effluent produced by MF with filtered primary effluent at a lower total cost than is possible now. The insensitivity of the MF effluent quality to changes in operating conditions might make MF a good candidate for use in high flow situations; the MF could act as a buffer, absorbing changes in influent flows and solids loadings, undergoing membrane flux and reactor solids concentration adjustments as necessary, but producing a uniform quality permeate throughout. This could be especially valuable if the permitted emergency discharge points (the 78” outfall and the Santa Ana River) ever had to be used.

OCSD's current ocean discharge permit limits the microorganism levels that are detected three miles offshore (rather than at the outfall exit). During emergencies, these discharge limits would not be enforced, but using MF would substantially reduce the microorganism levels in the effluent. Similarly, if there ever were a need to use an alternative outfall routinely, the microorganism reductions with MF could be critical for meeting the ocean discharge limits.

Historical OCSD data show the advantages of MF over A.S. in reducing indicator microorganism concentrations. Whereas A.S. averaged log reductions of 1.6 in total coliform concentration and 2.6 in coliphage (virus) concentration, MF in the current project demonstrated up to 6 log reduction in total coliform concentration and up to 4 log reduction in coliphage concentration.

Conclusions

The following overall conclusions can be drawn from the results of this test program:

- The Zenon ADC membranes in Phase I operated acceptably when treating unclarified secondary effluent with reactor concentrations of 1200 to 2400 mg/L and fluxes between 13 and 28 gpd/ft^2. The permeate often had TSS and cBOD levels at or below their detection limits (0.4 mg/L and 4.0 mg/L, respectively).

- The ADC membranes successfully rejected large fractions of the microbiological organisms in the feed. The total coliform tests indicated an average of 5.8 log removal, and the HPC tests showed an average of 2.4 log removal. The virus tests averaged 3 to 4 log removal.

- The OCP membranes in Phase II operated acceptably when treating primary effluent with target reactor concentrations of 4000 and 12,000 mg/L and fluxes of 19 and 29 gpd/ft^2. The permeate TSS and BOD levels often were at or below their detection limits regardless of the reactor concentration or membrane flux.

- The operating vacuum in Phase II was increased by higher flux conditions. Higher fluxes were associated with higher initial vacuum levels and with faster flux increases during operation (that is, faster fouling).

- The permeate turbidity in Phase II stayed below 1 NTU regardless of the test conditions.

- The coliform rejection by the system in Phase II was inconsistent. In some instances, the permeate coliform concentrations were below 10 MPN/100 mL (approximately 6 log removal), but other samples had coliform concentrations two orders of magnitude higher. Several known mechanical problems probably allowed some permeate samples to become contaminated, but there also might have been problems with the membranes themselves. Intact membranes and peripheral equipment should provide substantial bacterial removal.

- In a full-scale installation with the membranes suspended in aeration basins and air cycling being used to reduce the total aeration requirement, the MF blower power usage could be more or less than the existing A.S. power usage, depending on the MF operating conditions. The maximum MF aeration cost savings would occur with low-concentration/high-flux operation. With electricity valued at 6¢/kWh, MF could reduce OCSD's electricity costs $172,000/year.

- Replacing an A.S. system with an MBR would affect the DAF and anaerobic digestion processes also. It is estimated that the annual operating costs for these areas would be reduced $363,000 if MF were used.

- During Phase I and much of Phase II, OCWD fed an RO with the MF permeate. The RO reportedly performed well, producing a satisfactory quality permeate. The fouling rate was somewhat faster than has been experienced with other feeds, but the cycle time between cleanings was within acceptable limits.

1.0 PHASE I: INTRODUCTION

A large-scale microfiltration demonstration test unit from ZENON Environmental, Inc. (Zenon) was tested at OCSD. The Phase I tests were conducted from June 15, 1998 to September 22, 1998. The microfiltration unit's feed was unsettled mixed liquor from the effluent channel of the Plant 1 activated sludge process. During these tests, the operating conditions were varied to analyze the microfiltration unit's performance as a secondary clarifier with various reactor concentrations and flux rates.

The unit was operated with one, two, or three cassettes of Zenon's ADC membranes (0.08 µm mean pore size) in a 5000 gallon reactor tank. Each cassette had 1500 ft^2 of filter area.

Table 1 presents the operating targets for the three test conditions that were investigated. The wasting rate was adjusted to maintain the target reactor concentrations, and cassettes were taken on- or off-line to vary the flux rates.

Table 1. Operating Targets for Phase I Tests

Number of Cassettes	Flux (gpd/ft^2)	Permeate Flow (gpm)	Reactor TSS (mg/L)	Sludge Age (hours)
3	13	40	1200	~4
2	20	40	2400	~8
1	28	30	1200	~4

The first five weeks of testing used three cassettes. This was followed by six weeks of testing with two cassettes and one week of testing with one cassette. The final five weeks of testing returned to using three cassettes.

The unit was set to perform a backpulse for 30 seconds every 10 minutes, during which permeate was pumped from the collection tank in reverse through the membranes to clean the membrane pores. A weekly chemical cleaning (called a "maintenance cleaning" in some sources) was also performed by adding bleach to the permeate collection tank and manually backwashing the unit.

Approximately one-half of the MF permeate was fed into a reverse osmosis (RO) unit operated by the Orange County Water District (OCWD). Previous tests had suggested that MF permeate without additional treatment might not be a suitable RO feed. OCWD personnel maintained the RO unit and regularly sampled and analyzed the MF permeate and RO permeate. The analytical results are presented in Appendix C.

2.0 PHASE I: OPERATING ANALYSIS

Vacuum pressure and permeate flow readings were recorded hourly, and variations due to flux, time, airflow, chemical cleaning and reactor concentration were analyzed. Samples were collected daily from the feed, reactor and permeate for physical, chemical, and microbiological analyses. The permeate also was sampled and analyzed by OCWD. Virus removal by the membrane was tested with special sampling for five days in September.

2.1 Vacuum vs. Flux

Table 2 shows the average and range of the vacuum pressure, permeate flow rate, and flux for each cassette setting. The averages (arithmetic means) were calculated from data in Appendix A. Operation of the MF unit showed an overall increase in vacuum with an increase in flux as seen in Figure 1.

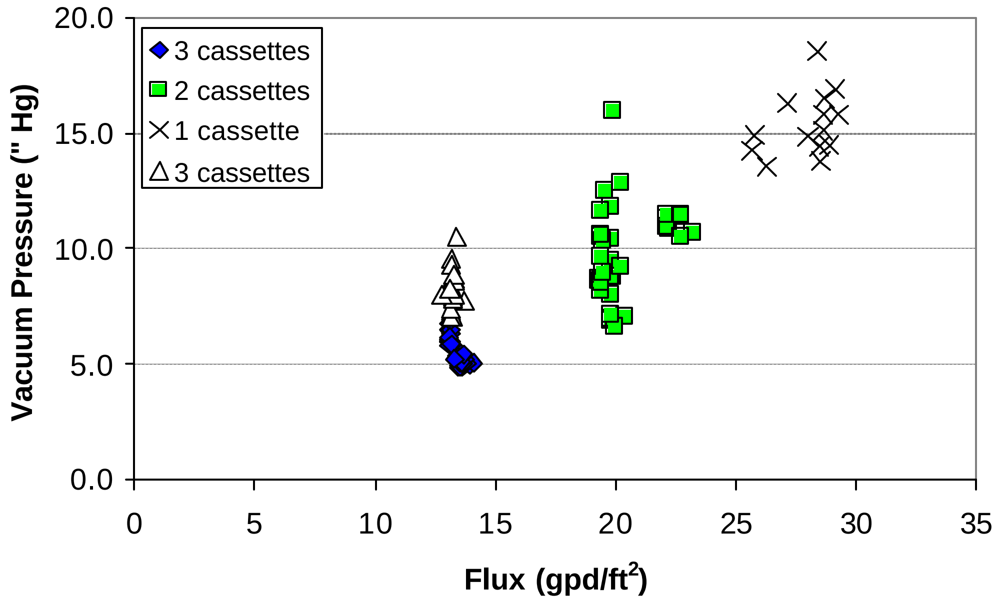
For the initial testing of three cassettes, the vacuum averaged 5.6 "Hg at an average flux of 13.4 gpd/ft^2. With two cassettes, the vacuum averaged 9.8 "Hg at an average flux of 20.4 gpd/ft^2. Some unusually high vacuum pressures (15 to 20 "Hg) were experienced with the two-cassette setting on two days; these may be related to abnormally high reactor TSS concentrations on those days. With only one cassette in operation, the vacuum averaged 15.4 "Hg at an average flux of 28.0 gpd/ft^2. Returning to the three-cassette setting produced an average vacuum of 8.4 "Hg at an average flux of 13.3 gpd/ft^2.

Table 2 also shows the relative theoretical power requirements for pumping the permeate at the average flow rates and vacuum pressures of the four test conditions.

Table 2. Actual Operating Conditions and Relative Pumping Power Requirements for Phase I Tests

Dates of Operation (in 1999)	No. of Cassettes	Vacuum ("Hg)		Flow Rate (gpm)		Flux (gpd/ft^2)		Relative Pumping Power Required
		Avg.	Range	Avg.	Range	Avg.	Range	
6/15 -7/21	3	5.6	4.3–8.2	41.8	38.6-49.1	13.4	12.4-15.7	1.0
7/21 - 8/26	2	9.8	6.4–20.1	42.5	39.8-49.0	20.4	19.1-23.5	1.8
8/26 - 9/8	1	15.4	13.4–20.0	29.2	25.7-33.0	28.0	25.2-31.7	1.9
9/8 - 9/22	3	8.4	6.6 –12.0	41.4	39.7-49.1	13.3	12.7-15.7	1.5

Figure 1. Effect of Flux on Vacuum Pressure

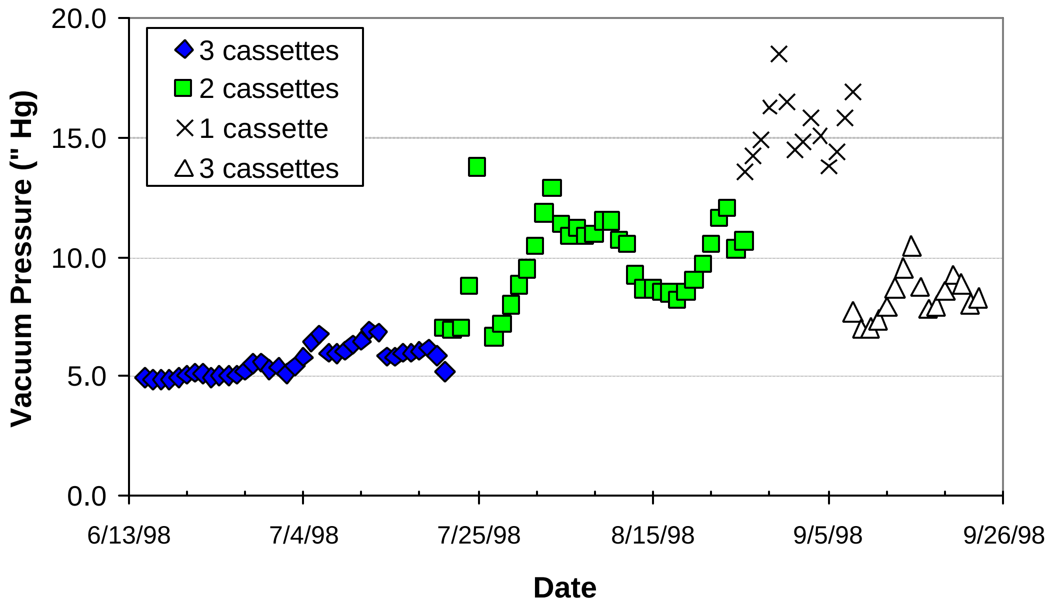


2.2 Vacuum vs. Time

Figure 2 shows that the vacuum pressure varied not only with the flux rate, but also during the period of operation of each cassette setting. Chemical cleanings were done weekly. The first two weeks of operation with three cassettes showed vacuum readings ranging between 4 and 6 "Hg, with no obvious pattern to the peaks and low points. Beginning in the third week, the vacuum pressure gradually increased until a chemical cleaning was performed that returned the vacuum pressure to approximately the starting (clean membrane) value. This pattern of increasing vacuum, then recovery with a chemical cleaning, was seen throughout the rest of the test.

The operation with two cassettes shows an unusual pattern after August 4: a small vacuum decrease and a stable period after a chemical cleaning, then a further decrease for one week before increasing again. Examination of the daily permeate flow data (Appendix A) shows that the permeate flow from August 4 through August 12 was set unusually high, causing the membrane to operate at a higher-than-target flux rate. When this was corrected, the pressure decreased and the vacuum profile became typical.

Figure 2. Average Daily Vacuum[1]



2.3 Airflow

The total airflow was decreased each time a cassette was removed from service so as to maintain a constant rate of approximately 75 scfm through each operating cassette. The only exception was during the last five days of operation (9/17 to 9/22) of the second three-cassette testing period, when the airflow was lowered to approximately 60 scfm per cassette at OCWD's request.

[1] The average daily vacuum is the arithmetic mean of the hourly vacuum readings.

3.0 PHASE I: LABORATORY RESULTS

Chemical and microbiological results for Phase I are shown in Appendix B. The results from OCWD's permeate sampling are shown in Appendix C. Virology results can be found in Appendix D.

3.1 Chemical Analysis Results

Table 3 summarizes the chemical and physical tests that were performed daily on the samples. The reactor mixed liquor was analyzed for TSS and VSS only. The feed was tested for TSS, VSS, ammonia, nitrite, nitrate, TKN, TOC and alkalinity, and the permeate was tested for TSS, VSS, cBOD, BOD, ammonia, turbidity, and pH. In addition, OCWD sampled the permeate for TKN, ammonia, nitrite, nitrate, TOC, and alkalinity. Average (arithmetic mean) results for the chemical tests are shown in Table 4.

Table 3. Feed, Reactor, and Permeate Analyses

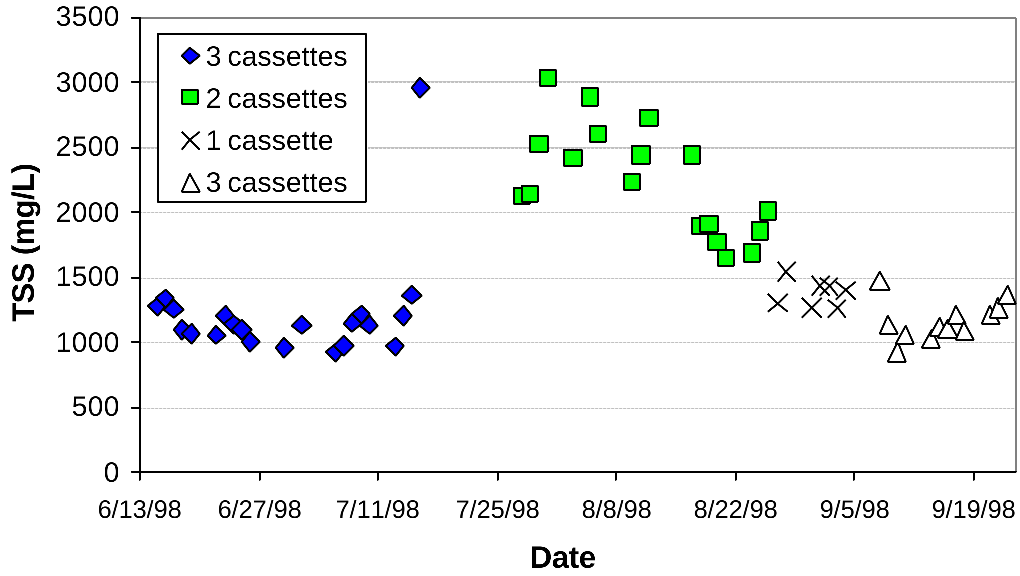
	TSS & VSS	cBOD & BOD	NH_3	TKN	Nitrite & Nitrate	Alkalinity	TOC	Turbidity	pH
Feed	X		X	X	X	X	X		
Reactor	X								
Permeate	X	X	X	X	X	X	X	X	X

Table 4. Average Chemical and Physical Results

	Feed	Permeate	Removal
Ammonia (mg/L)	17	13	24%
TKN (mg/L)	77	15	81%
Nitrite (mg/L)	0.36	2.92	-----
Nitrate (mg/L)	<1.77	2.66	-----
Alkalinity (mg/L)	272	214	21%
cBOD (mg/L)	-----	<4.0	-----
TOC (mg/L)	46.8	7.59	80%
pH	-----	7.77	-----
Turbidity (NTU)	-----	0.41	-----

The variations in reactor TSS concentration can be seen in Figure 3. The target value was 1200 mg/L with three cassettes operating and 2400 mg/L with two cassettes operating. Near the end of the two-cassette operation, the concentration decreased to about 1800 mg/L. During the subsequent one-cassette operation, the reactor TSS was kept around 1400 mg/L, and for the final setting with three cassettes, the TSS was returned to 1200 mg/L.

Figure 3. Reactor TSS Concentration



TSS removal appeared to be unaffected by the high concentration of solids in the reactor. The permeate TSS ranged from <0.4 mg/L (the detection limit) to 8 mg/L over the first two weeks of testing. Afterwards, the value remained consistently below 1.0 mg/L. As seen in Figure 4, the microfiltration unit removed 99.9 percent of the TSS.

Microfiltration reduced the total amount of nitrogen-containing compounds appearing in the permeate, and there was evidence of nitrification in the reactor. Figure 5 shows the ammonia concentrations in the feed and permeate; the ammonia in the permeate generally was about 24 percent less than in the feed. Figure 6 shows the corresponding TKN concentrations. The average TKN removal was about 81 percent from the feed to the permeate.

Figure 4. Comparison of TSS Concentrations

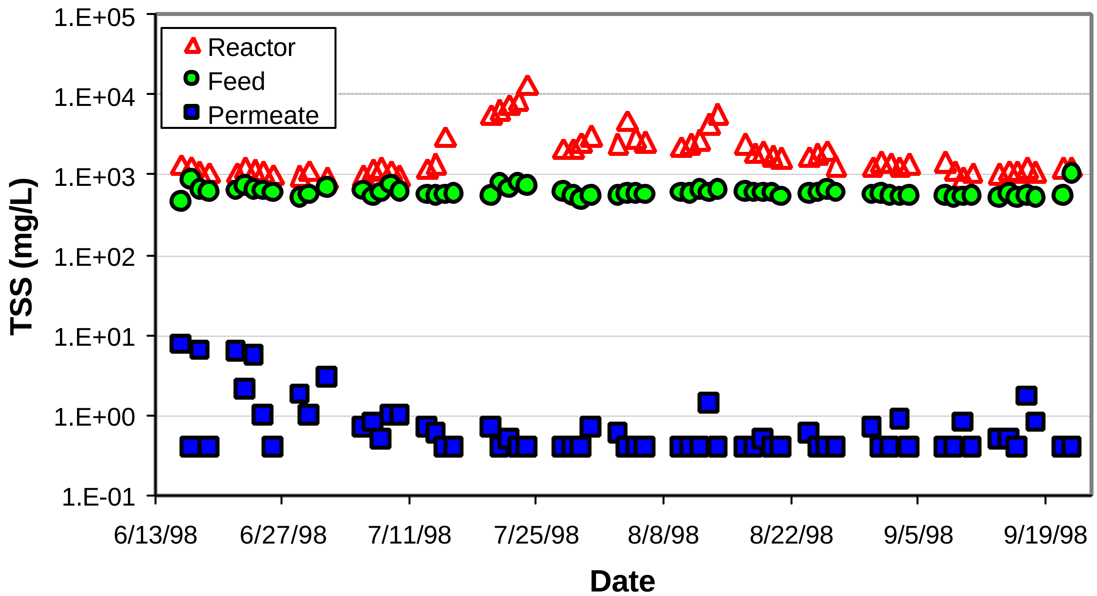


Figure 5. Ammonia Concentrations

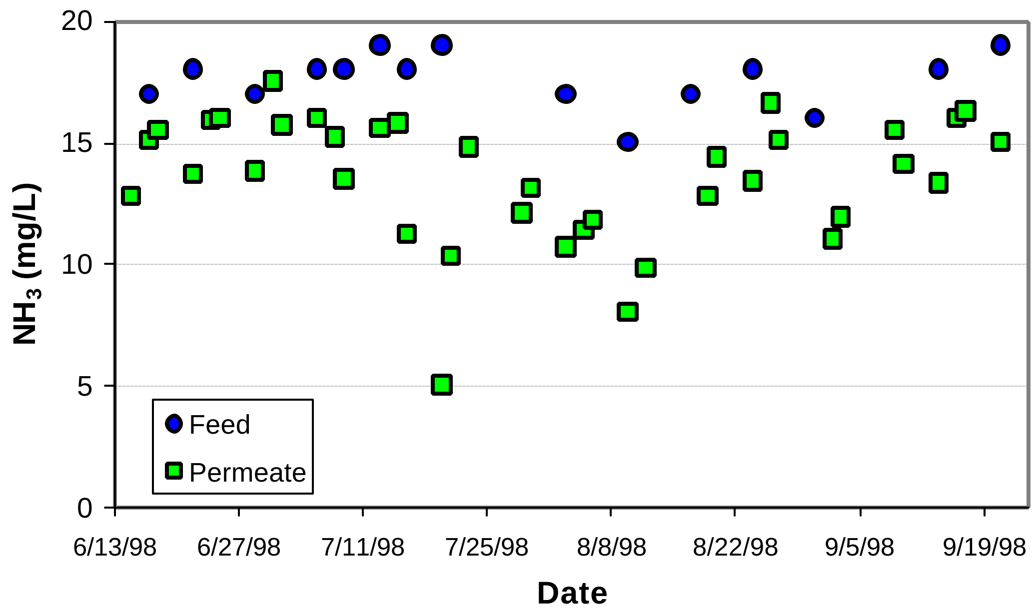


Figure 6. TKN Concentrations

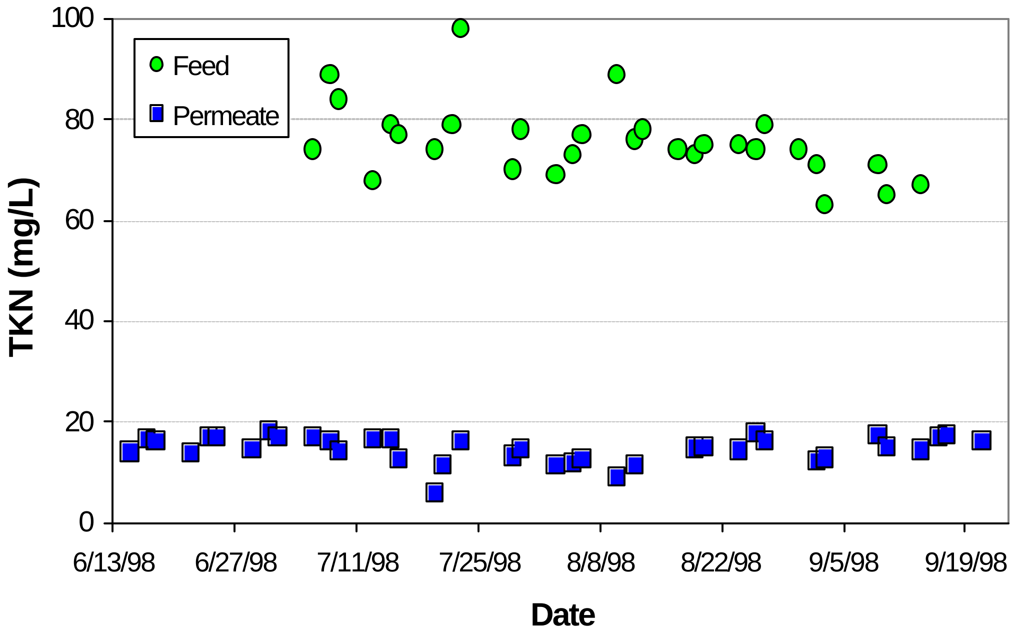


Figure 7 shows the nitrite concentration in the feed ranged from 0.01 to 1.83 mg/L, and the concentration in the permeate ranged from 0.80 to 7.20 mg/L. The nitrate concentration in the feed was usually below the detection limit of 1.77 mg/L and the concentration in the permeate ranged from 0.5 to 4.35 mg/L. (The nitrate detection limits were 0.2 mg/L for OCSD's laboratory and 1.77 mg/L for an outside laboratory that processed some samples.)

Figure 8 presents the TOC concentrations in the feed and permeate. The TOC in the feed ranged from 16.3 to 113 mg/L, and the TOC in the permeate ranged from 6.45 to 8.95 mg/L. The permeate TOC values ranged from 50 percent less to greater than 90 percent less than the corresponding feed values.

Figure 9 shows the alkalinity concentrations in the feed and permeate. The alkalinity in the permeate was generally about 24 percent (60 mg/L) less than in the feed.

The permeate turbidity ranged from 0.1 to 0.9 NTU. The permeate pH was slightly basic (between 7.50 and 8.05).

Figure 7. Nitrite and Nitrate Concentrations

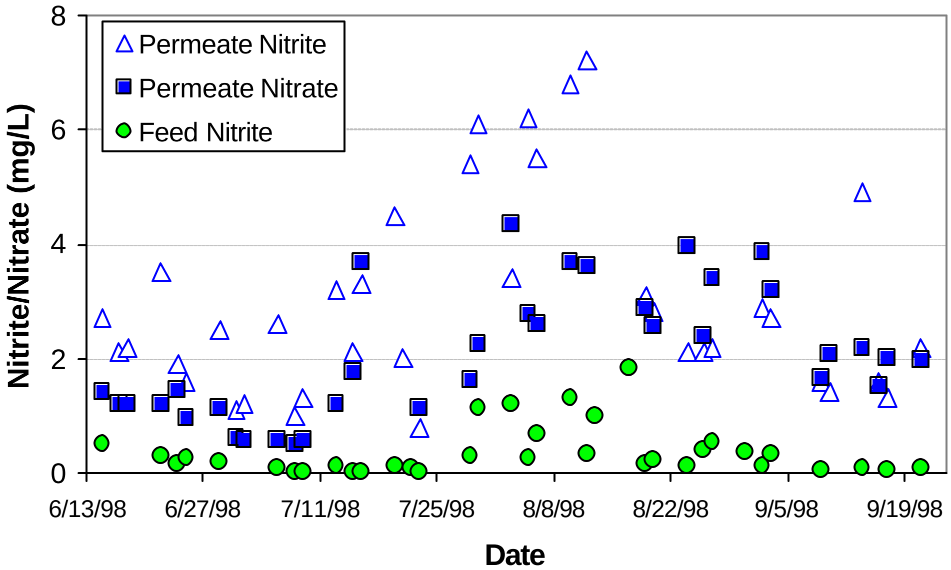


Figure 8. TOC Concentrations

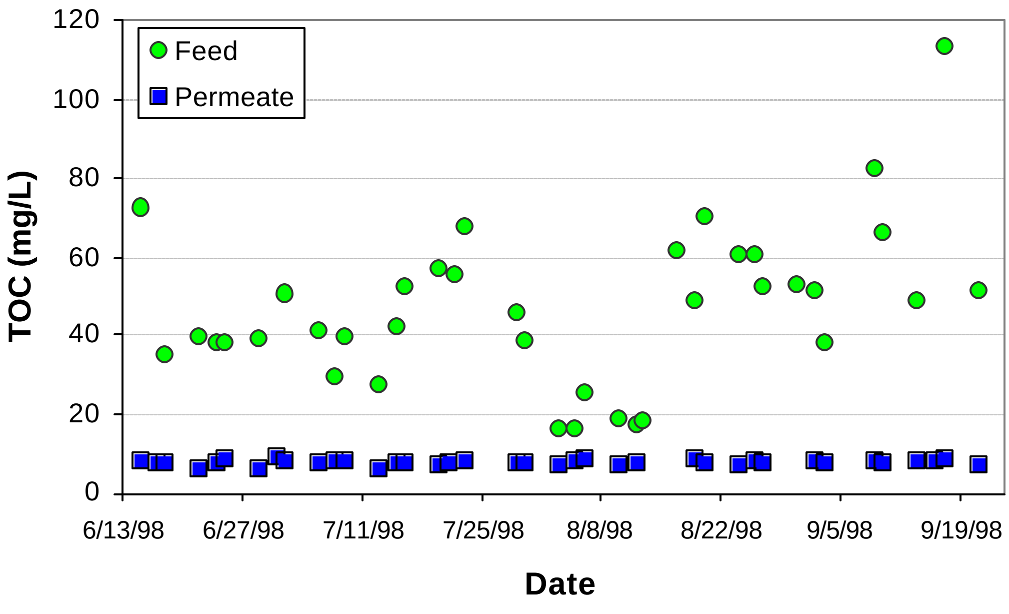
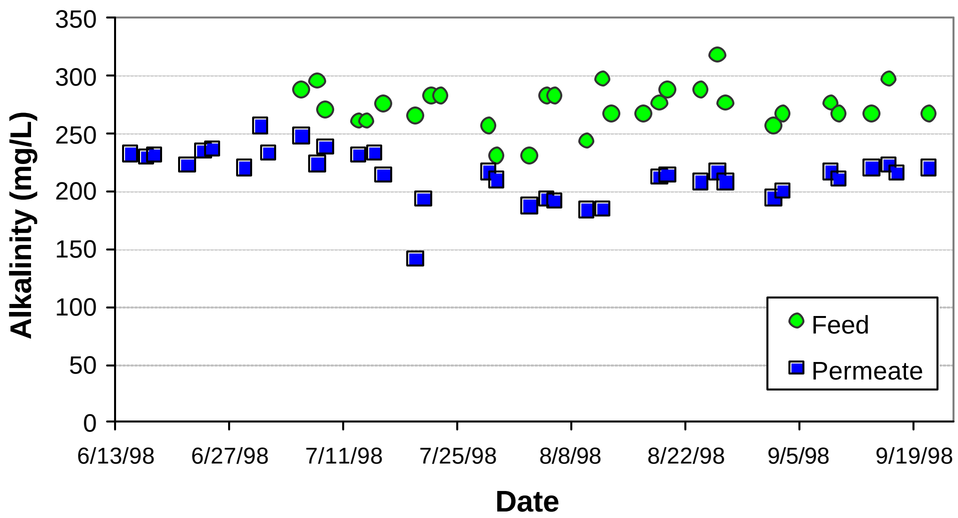


Figure 9. Alkalinity Concentrations



3.2 Bacteriology Results

Table 5 shows the averages for the total coliform, fecal coliform, HPC, and E. coli analyses. OCWD did not request fecal coliform analyses of the permeate. The HPC and E. coli tests for the permeate were performed by OCWD. (HPC detects a wide range of microorganisms but does not differentiate among them.)

The total coliform data, shown in Figure 10, show nearly 6 log removal, and the HPC data, shown in Figure 11, show greater than 2 log removal.[2] The permeate E. coli was consistently below the detection limit of 1 MPN/100 mL, with one exception of 10 MPN/100 mL.

[2] An upward trend in permeate HPC results during June and July is apparent in Figure 11. The reason for this is unknown. It is not accompanied by a similar increase in the permeate coliform.

Table 5. Bacteriology Results Summary

	Feed			Permeate			Log Removal
	Number of Samples	Range (10^6)	Average (10^6)	Number of Samples	Range	Average	
Total Coliform (MPN/100 mL)	39	2.2-50	13.2	14	4-130	23	5.8
Fecal Coliform (MPN/100 mL)	39	0.2–7.0	2.4	----	----	----	----
HPC (CFU/mL)	39	1.3-19	6.6	38	400-86,000	25,000	2.4
E. Coli (MPN/100 mL)	----	----	----	38	<1-10	<1	----

Figure 10. Total Coliform Analysis Results

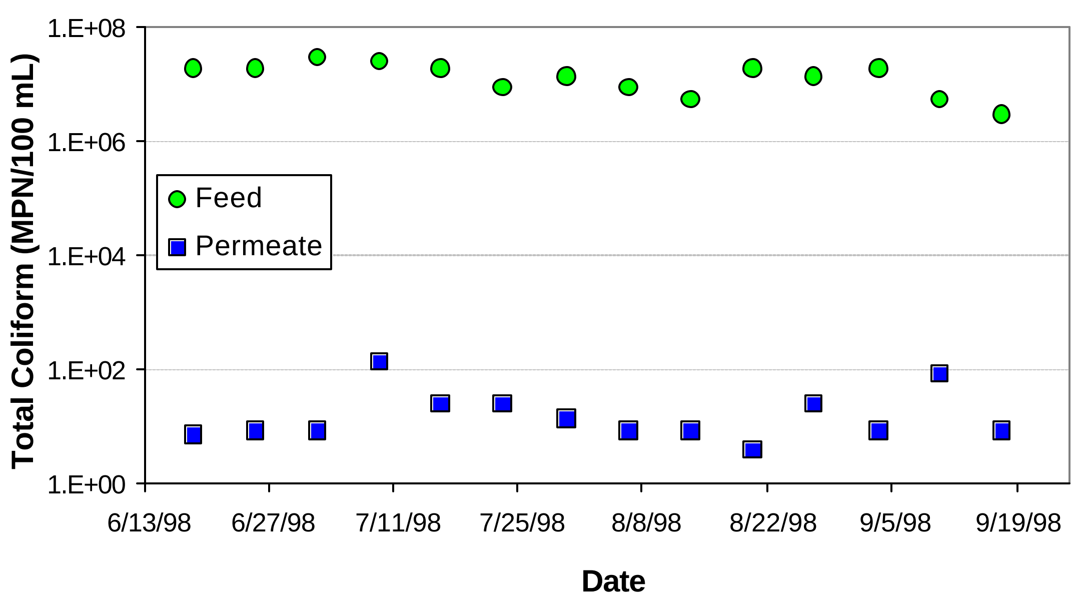
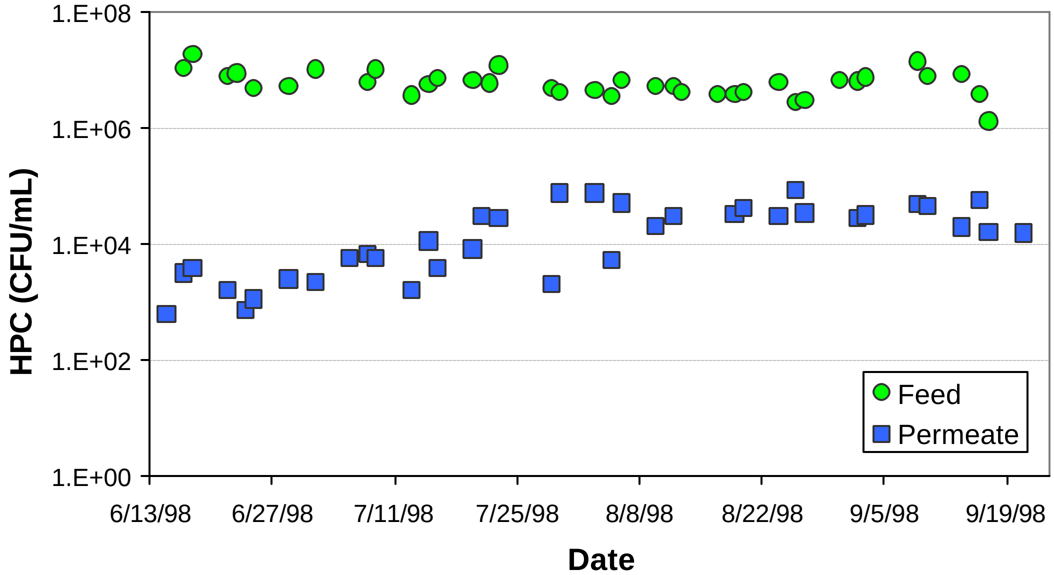


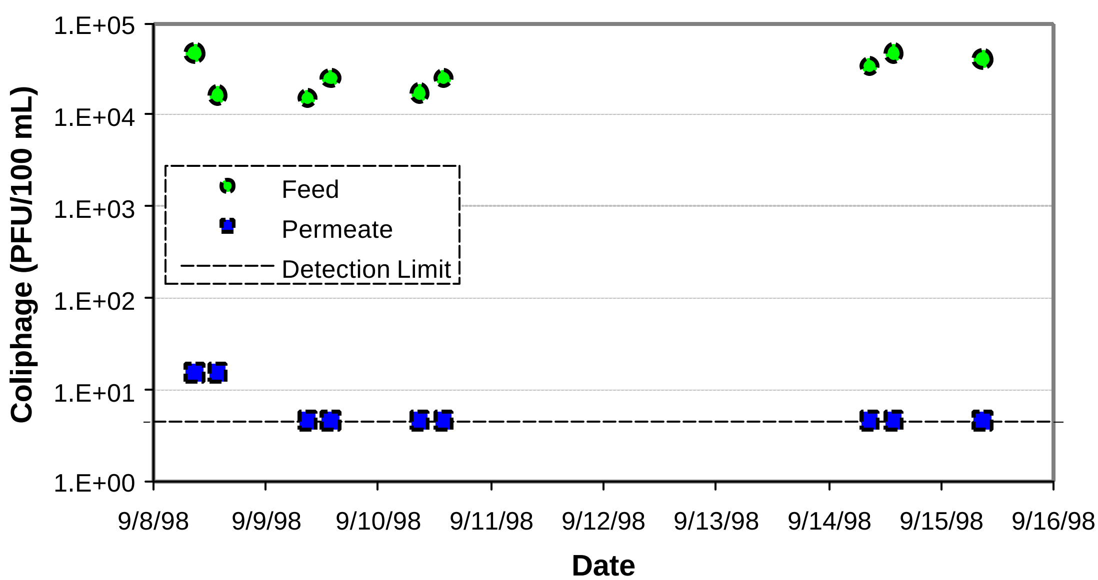
Figure 11. HPC Analysis Results



3.3 Virology Results

Twice-daily sampling for virus (coliphage) concentrations in the feed and permeate was performed for five days near the end of the Phase I tests. The average coliphage concentration in the feed was 29,000 PFU/100 mL and the concentration in the permeate was usually less than the detection limit (4.5 PFU/100 mL), indicating 3 to 4 log virus removal by the membrane. The permeate virus concentration was at or below the detection limit 75 percent of the time. Figure 12 shows the virology test results. (Permeate values less than the detection limit are shown as 4.5 PFU/100 mL.)

Figure 12. Virus Analysis Results



4.0 PHASE II: OVERVIEW

Phase II of the study was conducted to test microfiltration of primary effluent in a membrane bioreactor (MBR). In an MBR, the membranes are immersed in an aerated tank of concentrated mixed liquor, so the MBR acts as an aeration basin and a clarifier. Phase II used Zenon's OCP-500 membranes, which have a 0.035-µm mean pore size and 500 ft^2 of membrane in each cassette.

Four operating conditions were studied:

- low concentration, low flux;
- high concentration, low flux;
- low concentration, high flux;
- high concentration, high flux.

The target reactor TSS concentrations were 4000 mg/L (low concentration) and 12,000 mg/L (high concentration). The low flux conditions used six modules in one cassette, which provided a membrane surface area of 3000 ft^2, and the high flux conditions used four modules, which provided a membrane surface area of 2000 ft^2. Since the permeate flow target was 40 gpm in all cases, the flux was approximately 19 gpd/ft^2 during the low flux conditions and 29 gpd/ft^2 during the high flux conditions.

The target air flow to the membranes was 25 cfm per active module in all four operating conditions; this was the minimum value that Zenon would consider for a full-scale installation. The actual total airflow during the test was approximately 160 cfm (27 cfm/module) for the low flux condition and 100 cfm (25 cfm/module) for the high flux condition.

Table 6 summarizes the Phase II test conditions.

Table 6. Phase II Conditions

Condition	Target TSS Concentration (mg/L)	Membrane Surface Area (ft^2)	Total Aeration (cfm)	Specific Air Flow (cfm/ft^2)	Flux (gpd/ft^2)
Low concentration, low flux	4,000	3,000	160	0.053	19
High concentration, low flux	12,000	3,000	160	0.053	19
Low concentration, high flux	4,000	2,000	100	0.050	29
High concentration, high flux	12,000	2,000	100	0.050	29

The unit was operated with a 30-second backpulse in every 600 seconds (10 minutes) of operation. The first 570 seconds of each cycle were used for filtration, and the last 30 seconds for backpulsing with unchlorinated permeate.

At Zenon's request, a chemical (maintenance) cleaning of the membranes was performed three times per week (usually Monday, Wednesday, and Friday) unless operational conditions did not allow this. For these cleanings, two gallons of household bleach were mixed with 800 gallons of permeate in the CIP (clean-in-place) tank. Approximately 200 gallons of the mixture were backwashed through the membranes at ten minute intervals until the tank was emptied.

The chemical and biological tests performed during Phase II were:[3]

- Feed: TSS, VSS, NH_3, cBOD, BOD, TKN, alkalinity, and total coliform count.
- Reactor: TSS and VSS.
- Permeate: TSS, VSS, NH_3, cBOD, BOD, turbidity, pH, and total coliform count.

TSS and VSS tests were performed daily, and the other tests were performed less frequently. Table 7 lists the testing schedule that was followed.

Table 7. Phase II Testing Schedule

	Monday			Tuesday			Wednesday			Thursday			Friday		
TSS	F	R	P	F	R	P	F	R	P	F	R	P	F	R	P
VSS	F	R	P	F	R	P	F	R	P	F	R	P	F	R	P
NH_3	F		P												
cBOD	F		P							F		P	F		P
BOD	F		P							F		P	F		P
Turbidity			P						P			P			
pH			P												
TKN	F														
Alkalinity	F						F			F					
Total coliform	F		P				F		P						

F = Feed, R = Reactor, P = Permeate.

[3] More extensive microbiological analyses were requested by OCWD in Phase I, reflecting the importance of disinfection in tertiary treatment. Since Phase II was oriented more toward OCSD's interests, only total coliform analyses were done in Phase II.

In addition to the automatic data recording by the on-site computer, manual readings of various operational parameters were taken daily, usually once if a chemical cleaning was not performed and twice (before and after the cleaning) if a chemical cleaning was performed. These readings included the permeate flow (instantaneous rate and totalizer value), operating vacuum, backpulse flow (instantaneous rate and totalizer value), backpulse vacuum, air flow, and wasting period setting.

The permeate flow rate and the vacuum readings were taken approximately 30 seconds before a backpulse and 90 seconds after a backpulse. This provided an immediate observation of the backpulse's effect in reducing the vacuum.

5.0 PHASE II: RESULTS

5.1 Low Concentration, Low Flux Condition

The data for this condition were collected during 24 days' operation (3/2-3/4, 3/8, 3/19, 3/23-3/24, 3/29-4/4, 4/21-4/30). The discontinuity in the dates is due to various mechanical problems (e.g., malfunctioning valves) that interfered with the unit's operation. During the given dates, the reactor's TSS concentration ranged from 2625 to 4780 mg/L and averaged 3288 mg/L. This yields a calculated sludge age of approximately 5 days.

5.1.1 Vacuum

During this testing period the data collection computer failed, causing some of the vacuum data to be lost. Weekday vacuum readings recorded manually therefore were the only vacuum available data from 4/21 through 4/30.

The vacuum data were analyzed in two different forms. First, average hourly vacuum levels throughout the test period were calculated from the computer's vacuum readings taken at one-minute intervals. Second, the vacuum reading as a function of the elapsed time from a chemical cleaning was determined.

Figure 13 shows the hourly average vacuum during the test period. No general patterns of change are apparent in this data; the vacuum readings were nearly constant at this operating condition. This is clear in Figure 14, which shows the vacuum as a function of the elapsed time from a chemical cleaning. The vacuum ranged from 2.7 to 4.7 "Hg with no significant increase as the time after chemical cleaning increased; the vacuum increased only approximately 0.06 "Hg per day. This suggests that no substantial membrane fouling occurred between chemical cleanings, so at least in the short term, allowing a longer interval between cleanings would not harm performance for this low-stress operating condition.

While chemical cleanings normally were performed every 2-3 days, the maximum elapsed time shown in Figure 14 is approximately 8.3 days. This was the result of numerous mechanical problems during the period 3/15-3/24 that affected the operating schedule and, consequently, the cleaning schedule.

Figure 13. Hourly Vacuum Profiles
(Low Concentration, Low Flux Condition)

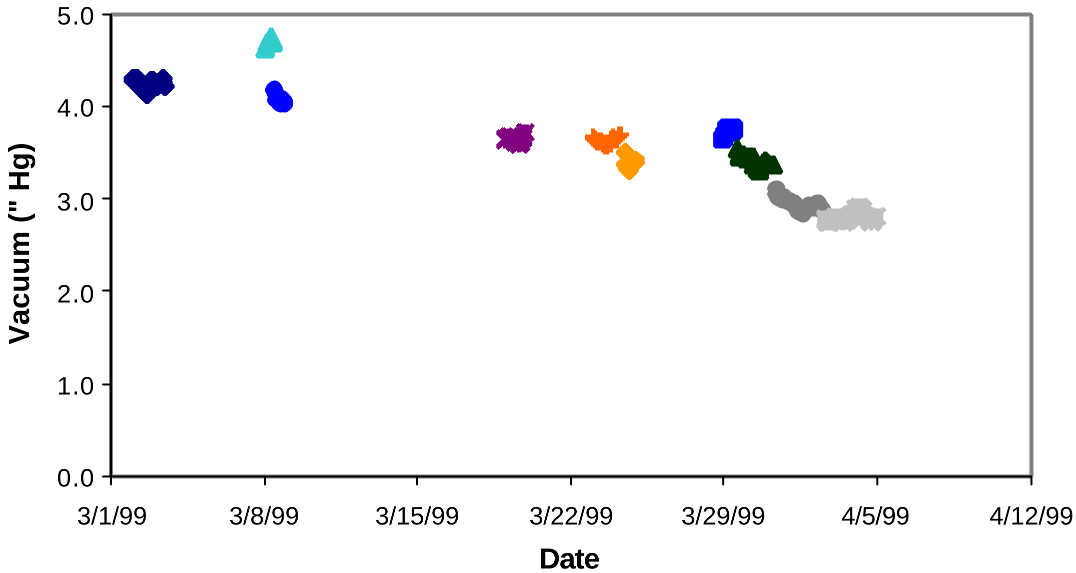
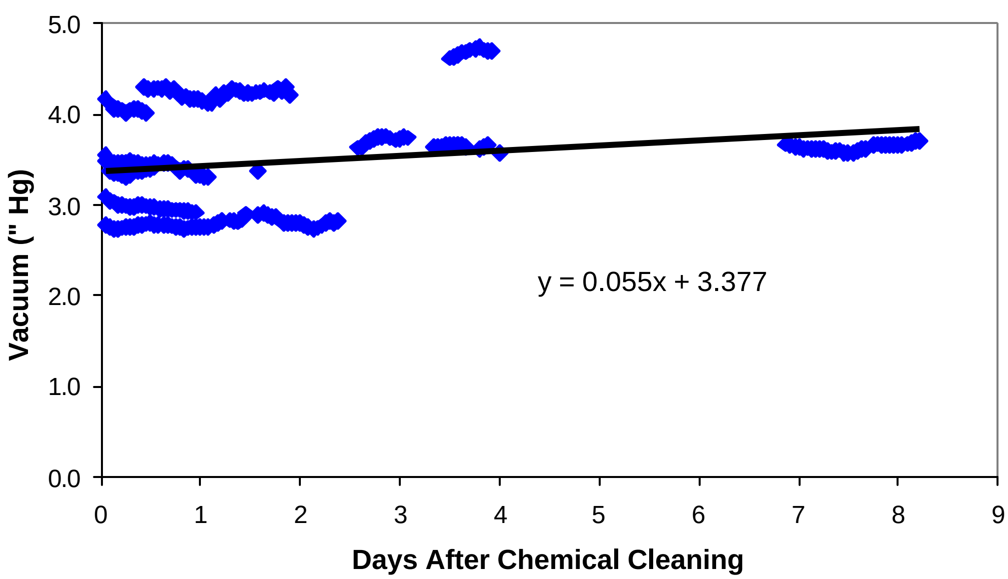


Figure 14. Vacuum vs. Days After Chemical Cleaning
(Low Concentration, Low Flux Condition)



5.1.2 Permeate Quality

Table 8 summarizes the permeate quality results for the low concentration, low flux condition. Approximately 99 percent of the TSS and VSS and 96 percent of the BOD and cBOD in the feed were removed by the MF. These removal efficiencies exceed the typical removal efficiencies achieved by OCSD's current secondary treatment. Nitrification reduced the NH_3 in the feed by 94 percent, although nitrification was not a goal for this test.

Table 8. Permeate Quality
(Low Concentration, Low Flux Condition)

Test	Number of Samples	Feed		Permeate		Average Removal
		Average	Range	Average	Range	
TSS (mg/L)	18	56	36-83	0.5	<0.4-0.8	99%
VSS (mg/L)	17	52	30-76	0.4	<0.4-0.6	99%
NH_3 (mg/L)	3	34	33-35	2	1-3	94%
$cBOD_5$ (mg/L)	10	113	89-142	4	<4	96%
BOD_5 (mg/L)	10	123	93-160	5	<4-14	96%

The permeate TSS, VSS, cBOD, and BOD concentrations were often reported as nondetectable at their respective detection limits, 0.4 mg/L for TSS and VSS and 4 mg/L for cBOD and BOD. The values listed in the table were calculated assuming the values were at these limits.

The coliform removal results obtained at this operating condition are presented in Table 9. The log removal ranged from 5.1 to 6.2 with an average removal of 5.9. The average was calculated excluding the high permeate coliform value of 3/24, which is not believed to be representative of the technology's capabilities during normal operation.

A compromised seal in the permeate piping was discovered at the end of the Phase II testing that could have allowed the permeate to become contaminated. High permeate coliform readings obtained at various times during Phase II could be the result of such contamination since other (low) readings indicate that the membranes are able to exclude most of the coliforms from the permeate. However, erratic bacterial removal with Zenon membranes has been noted in tests by OCWD and others that is not attributable to failures of specific system parts. It has been speculated that delamination of the polymer coating from its substrate may occur, or that the polymer may crack, either of which (on a very small scale) would allow bacteria to pass into the permeate without going through the polymer's pores and without necessarily leading to substantial contamination of the permeate by larger solid particles.

Table 9. Coliform Removal Results (Low Concentration, Low Flux Condition)

Days After Chemical Cleaning	Date	Total Coliform (MPN/100 mL)		Log of Coliform Value		Log Removal
		Feed	Permeate	Feed	Permeate	
2.8	3/8/99	8,000,000	50	6.90	1.70	5.2
8.2	3/24/99	30,000,000	230	7.48	2.36	5.1*
1.8	3/31/99	24,000,000	30	7.38	1.48	5.9
2.9	4/26/99	13,000,000	9	7.11	0.95	6.2
1.7	4/28/99	24,000,000	17	7.38	1.23	6.1
					Average	**5.9**

* Excluded from average value calculation; see text for explanation

The permeate turbidity ranged from 0.18 to 0.40 NTU and averaged 0.28 NTU. The pH ranged from 7.10 to 7.70 and averaged 7.44. These tests were not performed on the feed stream, so quantitative comparisons between the feed and permeate cannot be made. (By inspection, though, the permeate obviously was less turbid than the feed.)

5.2 High Concentration, Low Flux Condition

The data for this condition were collected during 16 days' operation (5/6, 6/4, 6/26-7/9). As in the low concentration, low flux tests, various computer and mechanical failures unrelated to the membranes were responsible for the scattered dates during which usable data were obtained. In this operating mode, a higher concentration of solids was maintained in the reactor while continuing the 40 gpm permeate flow through all six cassette modules. During the given dates, the reactor's TSS concentration ranged from 8,480 to 16,371 mg/L and averaged 12,703 mg/L. The calculated sludge age was approximately 17 days.

5.2.1 Vacuum

Figure 15 shows the hourly vacuum during the test period. The vacuum profiles generally are J-shaped, indicating that the vacuum continued to decrease for some time after a chemical cleaning was performed, after which the vacuum increased as expected. This behavior may actually reflect an instrument malfunction; this is discussed further in Section 5.3.1.

Figure 15. Hourly Vacuum Profiles
(High Concentration, Low Flux Condition)

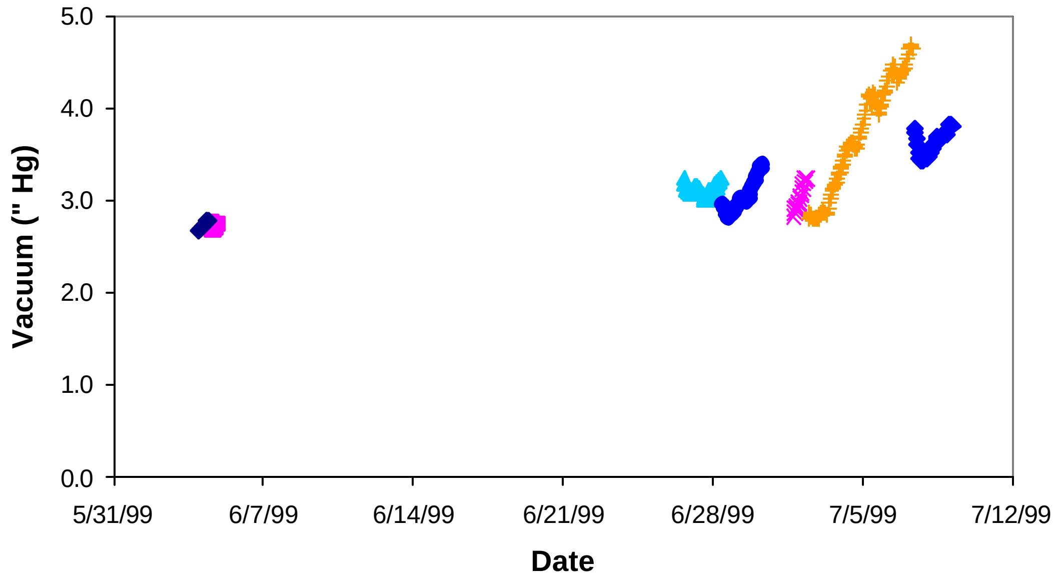
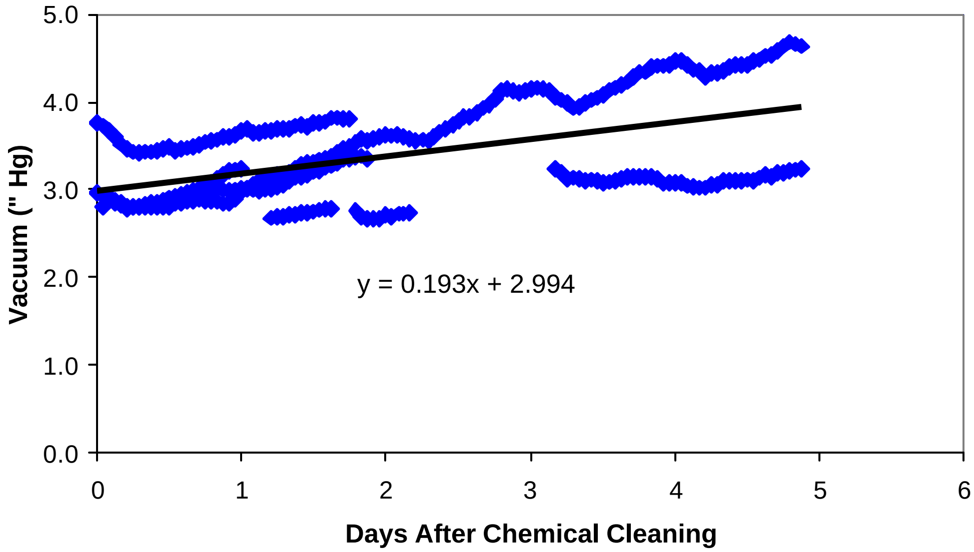


Figure 16 shows the vacuum as a function of the elapsed time from a chemical cleaning. The vacuum ranged from 2.7 to 4.7 "Hg and increased at a rate of approximately 0.19 "Hg per day after chemical cleaning.

In Figure 16, there appears to be a cyclic variation in the vacuum with a period of about 1½ days starting approximately one day after a chemical cleaning. Whether this pattern actually exists or is just an artifact of the analysis is unknown. However, the amplitude probably is too small to be of practical significance.

Figure 16. Vacuum vs. Days After Chemical Cleaning (High Concentration, Low Flux Condition)



5.2.2 Permeate Quality

Table 10 summarizes the feed and permeate quality results for the high concentration, low flux condition. The removals of TSS, VSS, cBOD, and BOD were similar to those observed in the corresponding low concentration condition. Approximately 99 percent of the TSS and VSS and 97 percent of the cBOD and BOD were removed from the feed. The removal efficiencies exceed the typical removal efficiencies achieved by OCSD's current secondary treatment.

The net ammonia removal observed at this condition was approximately 22 percent. There are at least two possible factors that could contribute to this low nitrification level. Since the air rate was not increased when going from the low concentration to the high concentration condition, the high concentration of biologically active organisms in the reactor could have reduced the dissolved oxygen (DO) level in the reactor below 1 mg/L. In this condition, oxygen becomes a limiting nutrient for nitrifying bacteria and nitrification slows or ceases. Another possibility is that the relatively long sludge age allowed ammonia nitrogen that had already been assimilated into the cells to be returned to the water through lysis and autooxidation of dead bacterial cells; this would reduce the apparent amount of nitrification taking place.[4]

[4] Tchobanoglous, G. and Burton, F., 1991. *Wastewater Engineering: Treatment, Disposal, and Reuse (3rd ed.).* New York: McGraw-Hill Inc., pp. 430-431.

Table 10. Permeate Quality (High Concentration, Low Flux Condition)

Test	Number of Samples	Feed		Permeate		Average Removal
		Average	Range	Average	Range	
TSS (mg/L)	11	67	43-111	0.6	<0.4-1.0	99%
VSS (mg/L)	10	62	42-99	0.5	<0.4-0.9	99%
NH_3 (mg/L)	1	27	27	21	21	22%
$cBOD_5$ (mg/L)	6	120	94-148	4	<4	97%
BOD_5 (mg/L)	6	130	100-158	4	<4	97%

The permeate TSS, VSS, cBOD, and BOD concentrations were often reported as non-detectable at their respective detection limits, 0.4 mg/L for TSS and VSS and 4 mg/L for cBOD and BOD. The values listed in the table were calculated assuming the values were at these limits.

The coliform removal results obtained at this operating condition are presented in Table 11. The log removal ranged from 5.1 to 5.5 with an average value of 5.5. The average was calculated excluding the high permeate coliform value of 7/7, which is believed to reflect contamination caused by compromised seals in the permeate piping that were discovered after the testing was finished. Therefore, this value is not considered to be representative of the Zenon membrane's capabilities. However, as mentioned in Section 5.1.2, erratic bacterial rejection results have been noted in other Zenon tests, so this may be characteristic of laminated membranes.

Table 11. Coliform Removal Results (High Concentration, Low Flux Condition)

Days After Chemical Cleaning	Date	Total Coliform (MPN/100 mL)		Log of Coliform Value		Log Removal
		Feed	Permeate	Feed	Permeate	
4.9	6/28/99	24,000,000	80	7.38	1.90	5.5
1.9	6/30/99	13,000,000	50	7.11	1.70	5.4
6.0	7/7/99	30,000,000	240	7.48	2.38	5.1 *
					Average	**5.5**

* Excluded from average value calculation; see text for explanation

The permeate turbidity ranged from 0.30 to 0.74 NTU and averaged 0.50 NTU. The only permeate pH reading, taken on 6/28, was 8.10.

5.3 Low Concentration, High Flux Condition

The data for this condition were collected during 13 consecutive days of operation (11/7-11/19). In this operating mode the membrane was stressed by increasing the flux rate about 50 percent (from 19 to 29 gpd/ft^2). During this period the reactor TSS concentration ranged from 1988 to 3883 mg/L and averaged 3194 mg/L. The calculated sludge age was approximately 5 days.

5.3.1 Vacuum

Figure 17 shows the hourly vacuum during the test period. There is a common pattern in the vacuum readings between chemical cleanings (which are indicated by the breaks in the data).[5] Immediately after a chemical cleaning, the vacuum increased for 11 to 20 hours, then decreased for 6 to 13 hours, then increased thereafter until the next chemical cleaning. The initial increases may reflect erratic operation of the sensor due to the bleach used in the chemical cleanings. (See Section 5.4.1 for additional details about this phenomenon.) However, the pattern seen in this condition does not exactly match the pattern reported by Zenon. Therefore, the possibility exists that this is a genuine pattern of the microfiltration process, although there is no apparent reason why this would occur.

Figure 18 shows the vacuum as a function of the elapsed time from a chemical cleaning. The vacuum ranged from 8.2 to 12.6 "Hg and increased at a rate of approximately 1.01 "Hg per day after chemical cleaning. There is some indication of the cyclic variation that was observed in the high concentration, low flux data (cf. Figure 16), but the effect is less distinct and may simply be normal scatter in the data.

[5] The decrease on 11/14 does not indicate a chemical cleaning. This was a weekend day during which an operational upset shut down the unit for 8 hours. By 11/15, the unit was operating normally.

Figure 17. Hourly Vacuum Profiles
(Low Concentration, High Flux Condition)

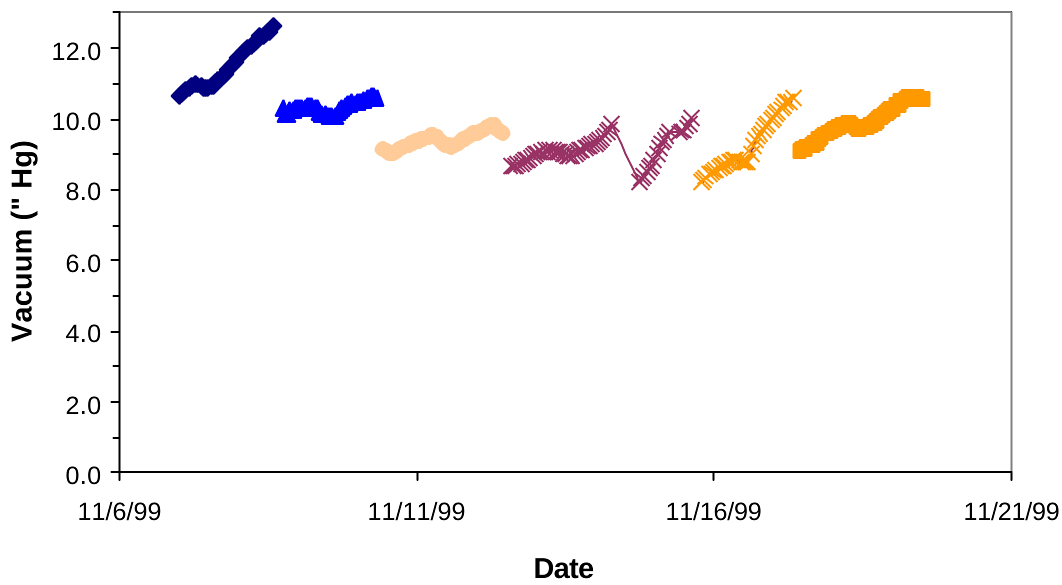
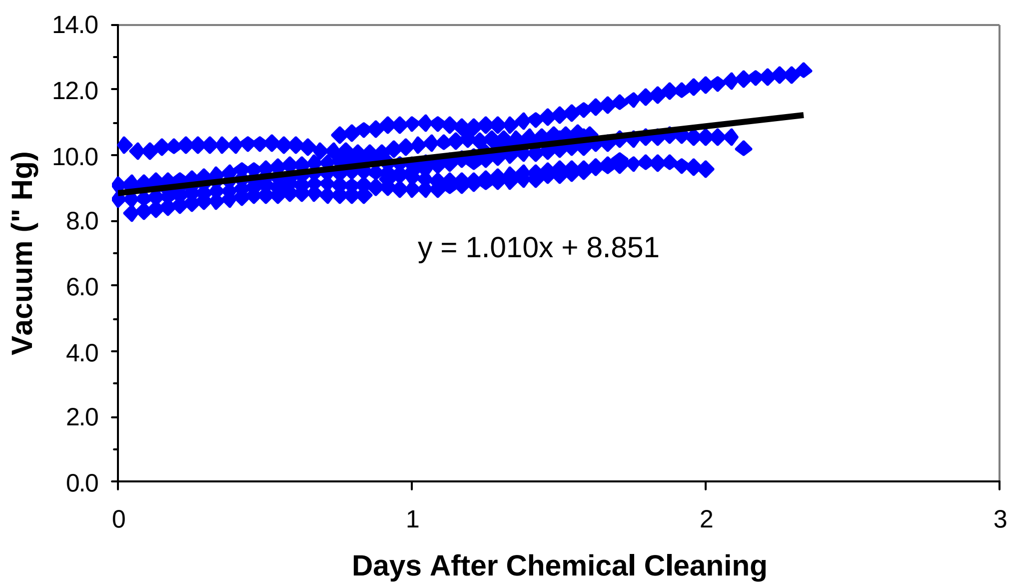


Figure 18. Vacuum vs. Days After Chemical Cleaning
(Low Concentration, High Flux Condition)



5.3.2 Permeate Quality

Table 12 summarizes the feed and permeate quality results for the low concentration, low flux condition. The removals of TSS, VSS, cBOD, and BOD were comparable to those observed in the other conditions. Approximately 99 percent of the TSS and VSS and 97 percent of the cBOD and BOD were removed from the feed. Nitrification removed 96 percent of the NH_3, which was comparable to the result obtained in the other low concentration condition.

Table 12. Permeate Quality
(Low Concentration, High Flux Condition)

Test	Number of Samples	Feed		Permeate		Average Removal
		Average	Range	Average	Range	
TSS (mg/L)	8	58	49-73	0.5	<0.4-1.2	99%
VSS (mg/L)	4	52	37-60	0.5	<0.4-0.9	99%
NH_3 (mg/L)	2	26	24-27	1	1	96%
$cBOD_5$ (mg/L)	5	122	94-145	4	<4	97%
BOD_5 (mg/L)	5	129	97-148	4	<4	97%

The permeate TSS, VSS, cBOD, and BOD concentrations were often reported as non-detectable at their respective detection limits, 0.4 mg/L for TSS and VSS and 4 mg/L for cBOD and BOD. The values listed in the table were calculated assuming the values were at these limits.

The coliform removal results obtained at this operating condition are presented in Table 13. The log removal ranged from 5.3 to 6.4 with an average value of 6.0. The low permeate coliform values are consistent with expectations for this technology. The faulty seals in the permeate piping do not appear to have resulted in any of the samples being contaminated.

Table 13. Coliform Removal Results
(Low Concentration, High Flux Condition)

Days After Chemical Cleaning	Date	Total Coliform (MPN/100 mL)		Log of Coliform Value		Log Removal
		Feed	Permeate	Feed	Permeate	
2.7	11/8/99	13,000,000	8	7.11	0.90	6.2
1.7	11/10/99	30,000,000	13	7.48	1.11	6.4
2.7	11/15/99	11,000,000	50	7.04	1.70	5.3
1.6	11/17/99	24,000,000	13	7.38	1.11	6.3
					Average	**6.0**

The permeate turbidity ranged from 0.08 to 1.55 NTU and averaged 0.40 NTU. The pH ranged from 7.36 to 7.83 and averaged 7.60.

5.4 High Concentration, High Flux Condition

The data for this condition were collected during 24 consecutive days of operation (9/16-10/8), during which the membrane was operated at a high flux rate (~29 gpd/ft^2) and a high reactor solids concentration. The reactor TSS concentration varied from 9,039 to 14,949 mg/L and averaged 11,509 mg/L, which results in a calculated sludge age of about 17 days.

5.4.1 Vacuum

Figure 19 shows the hourly average vacuum during the test period. After each chemical cleaning (corresponding to a break in the data line), the vacuum was reduced as expected, but it continued to decrease during the next 7 to 39 operating hours (usually for at least 30 hours), after which it increased until the next chemical cleaning. This pattern was counterintuitive but undeniable, and therefore quite puzzling.

After the Phase II testing was completed, Zenon discovered that the pressure transducer was incompatible with acid, and very likely incompatible with bleach as well. Exposing the transducer to these chemicals affects its sensitivity and produces erroneous readings. Zenon claims the effect lasts only about 12 hours, but this is based on tests with acid. The anomalous pattern seen in Figure 19 therefore is likely to be a result of instrumentation error and is not an accurate depiction of the vacuum profile following a chemical cleaning.

Figure 19. Hourly Vacuum Profiles
(High Concentration, High Flux Condition)

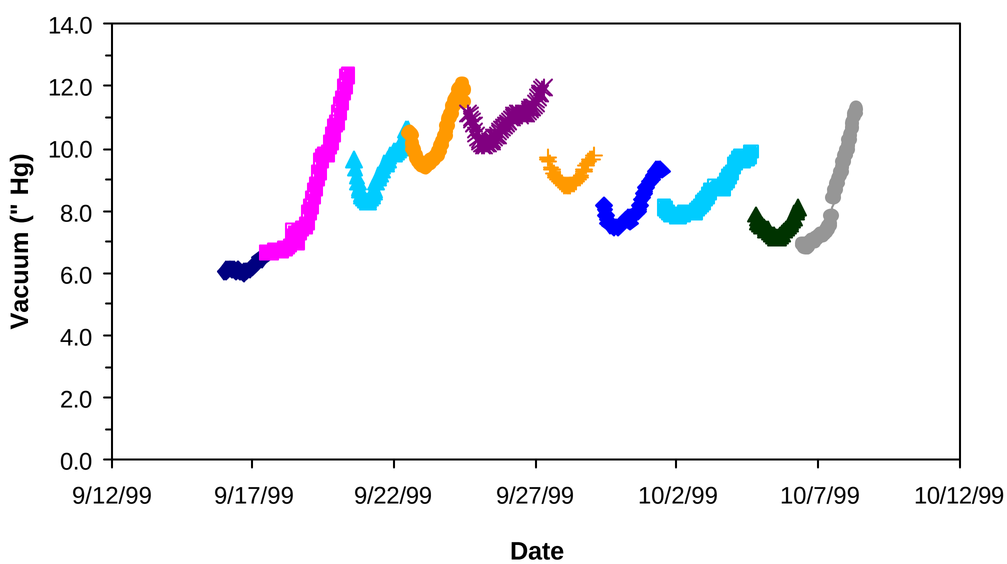
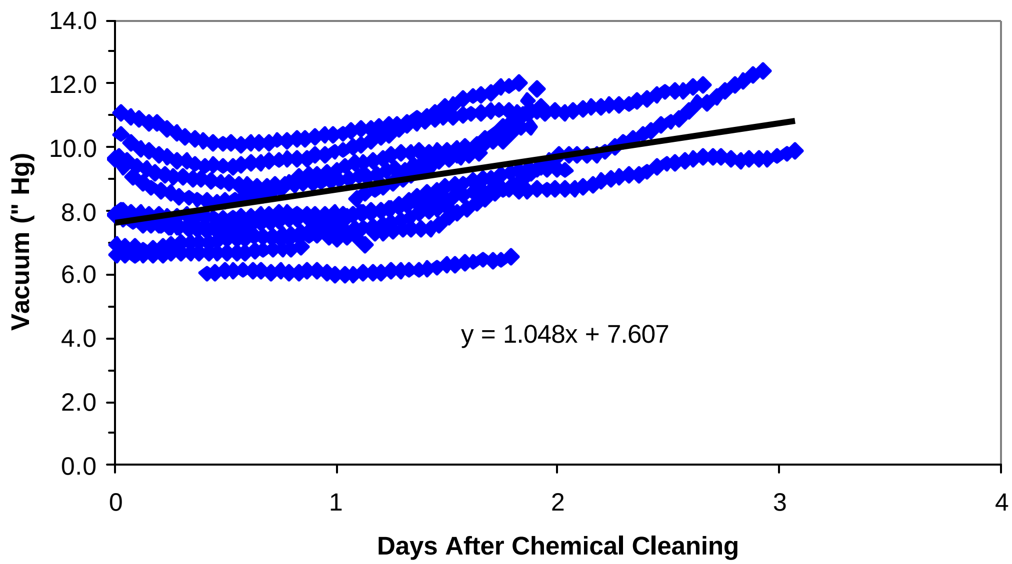


Figure 20 shows the vacuum as a function of the elapsed time from a chemical cleaning. The vacuum for this test period ranged from approximately 6 to 12 "Hg and increased an average of approximately 1.05 "Hg per day. The data show substantial scatter, even immediately after a chemical cleaning. As in the low concentration, high flux test, there may be some small cyclic variation visible in the data, but it is not pronounced and may reflect normal data scatter.

Figure 20. Vacuum vs. Days After Chemical Cleaning (High Concentration, High Flux Condition)



5.4.2 <u>Permeate Quality</u>

Table 14 summarizes the feed and permeate quality results for the high concentration, high flux condition. Despite the severe operating condition, the membrane's performance was comparable to that in the other test conditions. Approximately 99 percent of the TSS and VSS and 96 percent of the cBOD and BOD were removed from the feed. The net ammonia removal was approximately 12 percent, a low value that is consistent with the removal observed in the other high concentration condition (cf. Section 5.2.2)

Table 14. Permeate Quality
(High Concentration, High Flux Condition)

Test	Number of Samples	Feed		Permeate		Average Removal
		Average	Range	Average	Range	
TSS (mg/L)	15	56	52-85	0.5	<0.4-1.0	99%
VSS (mg/L)	10	53	29-75	0.5	<0.4-0.9	99%
NH_3 (mg/L)	3	22	22-23	20	16-24	12% *
$cBOD_5$ (mg/L)	10	105	27-133	4	<4	96%
BOD_5 (mg/L)	10	121	110-133	4	<4-6	96%

The permeate TSS, VSS, cBOD, and BOD concentrations were often reported as nondetectable (ND) at their respective detection limits (0.4 mg/L for TSS and VSS and 4 mg/L for cBOD and BOD). The table values were calculated assuming that ND results were at these limits.

* The NH_3 removal calculated for 9/20/99 was negative; the value was replaced by zero in the calculation of the average removal.

The coliform removal results obtained at this operating condition are shown in Table 15. The log removal ranged from 4.2 to 6.3 with an average value of 5.1. The high permeate coliform readings are believed to be the result of permeate contamination, so they might not reflect the intrinsic capability of the filtration system.

Two observable contamination sources and one possible micro-scale source were identified. One observable source was the leaking seals discussed in Sections 5.1.2 and 5.2.2. The other observable source was leaks in the permeate header hose above the reactor tank that could allow the permeate to be contaminated by reactor foam. One such leak was discovered and repaired in early September before this operating condition was tested, but there may have been additional leaks that were not discovered and repaired. Very small-scale delamination or cracking of the membrane's polymer coating has been proposed by other researchers to explain erratic bacterial removal results from earlier Zenon membrane tests; if this occurs, erratic bacterial removal could be characteristic of laminated membranes in general.

The permeate turbidity ranged from 0.34 to 0.90 NTU and averaged 0.63 NTU. The pH ranged from 7.75 to 8.07 and averaged 7.93.

Table 15. Coliform Removal Results (High Concentration, High Flux Condition)

Days After Chemical Cleaning	Date	Total Coliform (MPN/100 mL)		Log of Coliform Value		Log removal
		Feed	Permeate	Feed	Permeate	
2.9	9/20/99	30,000,000	300	7.48	2.48	5.0
1.8	9/22/99	30,000,000	170	7.48	2.23	5.2
2.7	9/27/99	90,000,000	500	7.95	2.70	5.3
1.7	9/29/99	24,000,000	1,600	7.38	3.20	4.2
2.8	10/4/99	24,000,000	900	7.38	2.95	4.4
1.6	10/6/99	17,000,000	8	7.23	0.90	6.3
					Average	**5.1**

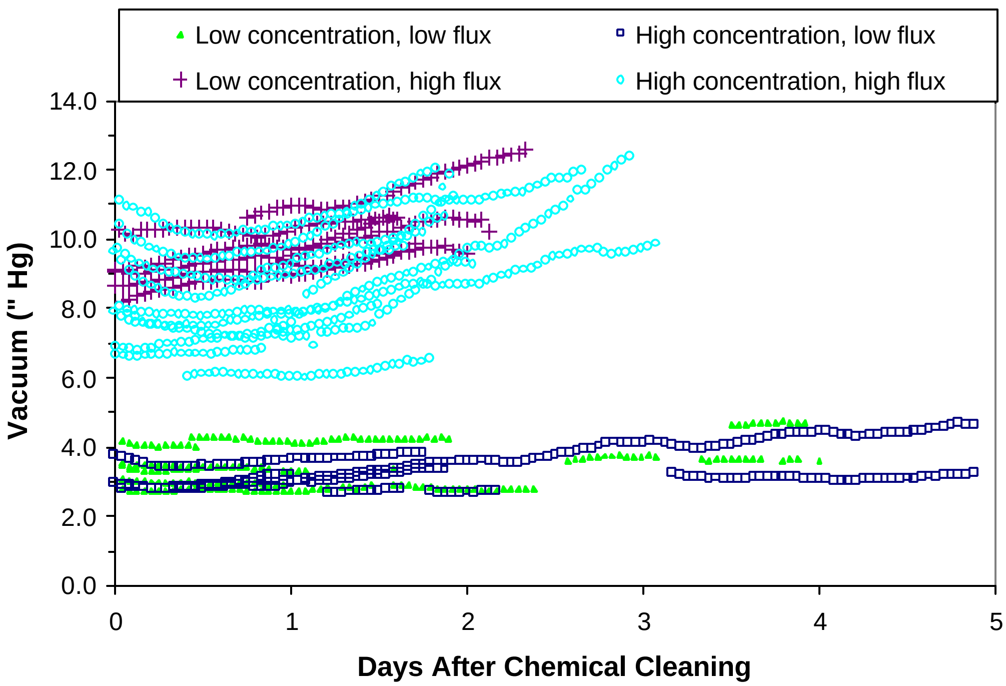
6.0 PHASE II: DISCUSSION

6.1 Vacuum

Figure 21 shows the vacuum profiles for the four Phase II conditions as a function of the elapsed time from a chemical cleaning. The linear regression equations for each condition, which have been presented previously, are summarized below:

- low concentration, low flux: y = 0.055x + 3.377
- high concentration, low flux: y = 0.193x + 2.994
- low concentration, high flux: y = 1.010x + 8.851
- high concentration, high flux: y = 1.048x + 7.607

Figure 21. Vacuum Profiles (All Phase II Conditions)



The operating vacuum increased at higher flux conditions. The difference in vacuum between the two flux conditions averaged 4 to 6 "Hg even immediately after a chemical cleaning. One would expect that forcing more flow through the pores in each time interval would require a greater driving force, so this result was not surprising.

The higher flux conditions also led to more rapid membrane fouling as indicated by the slopes of the regression lines. Increasing the flux by 50 percent increased the fouling rate by factors of 5 (high concentration) and 18 (low concentration). However, even in the high flux conditions the vacuum did not reach the high limit (18 “Hg) within three days after a chemical cleaning was performed, and the low flux conditions could have continued for much longer intervals between chemical cleanings without approaching a critical vacuum level. [Zenon recommends the more frequent chemical cleanings, though, to preserve the long-term membrane performance.]

Higher reactor concentrations did not increase the average clean membrane vacuum levels (as indicated by the regression equation y-intercepts). In fact, higher concentrations appeared to reduce the vacuum levels slightly. Zenon has stated that higher solids concentrations provide a more effective scrubbing action for the membrane surface, so perhaps the observed differences are real. However, more effective scrubbing would be expected to affect the fouling rate more than the clean membrane vacuum, and the fouling rate in the higher concentration tests at a fixed flux was faster, not slower, than in the lower concentration tests. Given the amount of scatter in the data and the relatively small differences involved, the reactor concentration in the range tested may be unimportant in determining the operating vacuum.

6.2 Permeate TSS and VSS Concentrations

Figure 22 shows the permeate TSS concentration as a function of elapsed time after chemical cleaning for all Phase II conditions, and Figure 23 shows the comparable VSS data. The TSS concentrations ranged from 0.4 mg/L to 1.2 mg/L, and the VSS concentrations ranged from 0.4 mg/L to 0.9 mg/L. For both tests, the lower detection limit was 0.4 mg/L. Neither the solids concentration in the reactor nor the flux affected the TSS or VSS permeate concentrations. The elapsed time after chemical cleanings also did not affect the permeate TSS and VSS.

Figure 22. Permeate TSS (All Phase II Conditions)

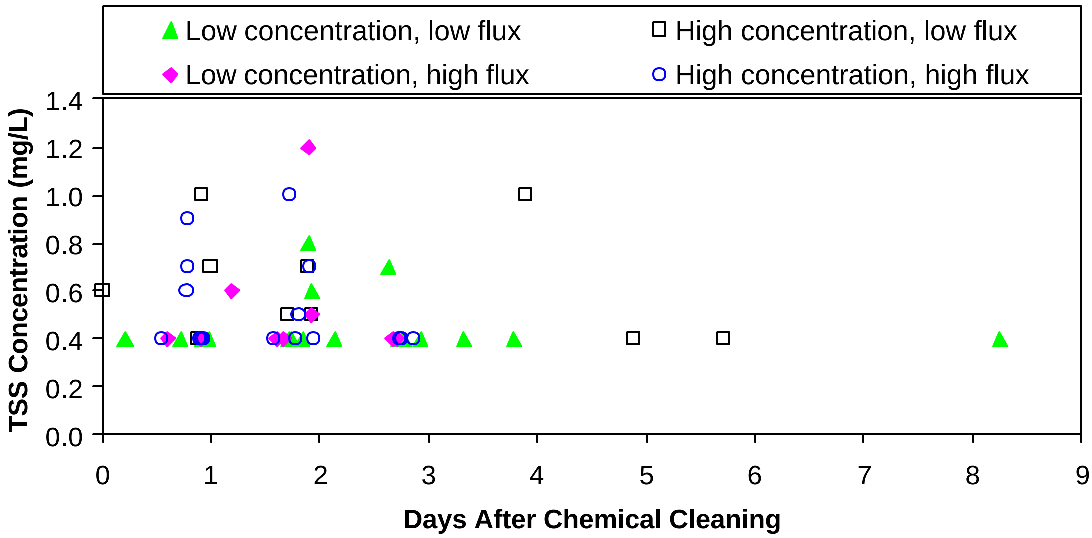
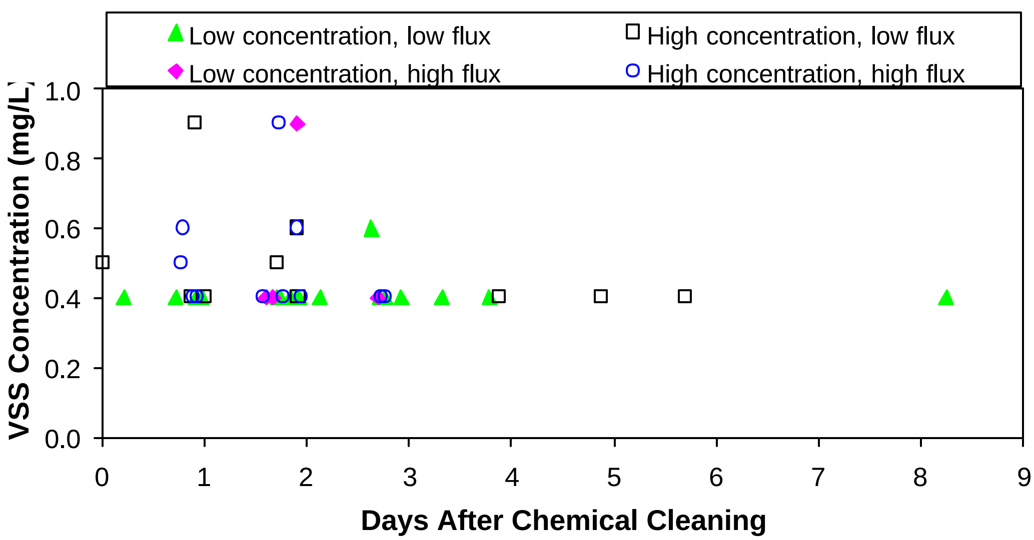


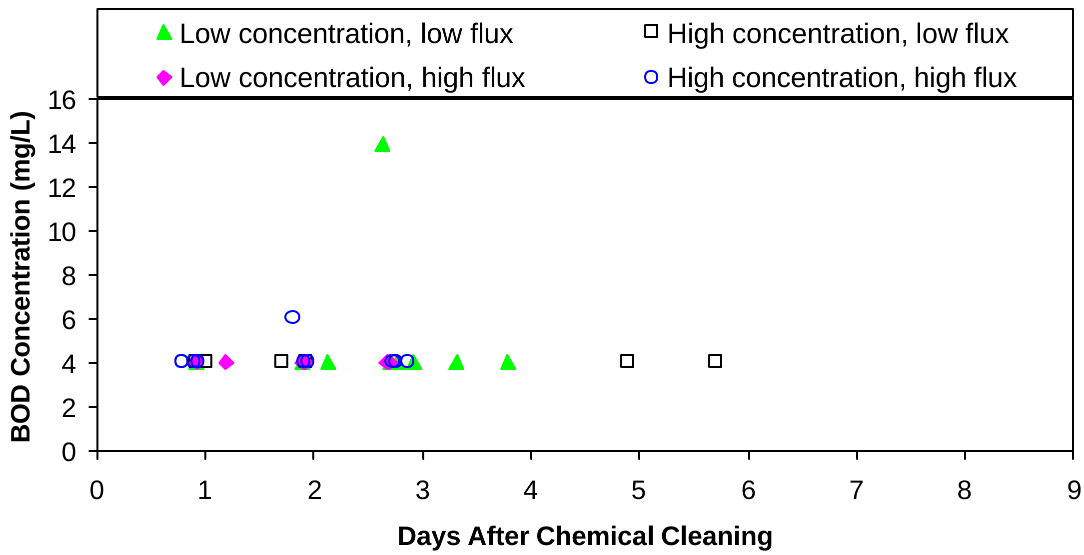
Figure 23. Permeate VSS (All Phase II Conditions)



6.3 Permeate BOD Concentrations

Figure 24 shows the permeate BOD concentration as a function of the elapsed time after a chemical cleaning for the Phase II conditions. The plot reveals that the permeate BOD remained at or below the detection limit, 4 mg/L, during all the conditions tested, except for 2 (out of 33) occasions. It seems likely that the samples on those occasions were contaminated or the test results are in error because the results are so clearly atypical.

Figure 24. Permeate BOD (All Phase II Conditions)

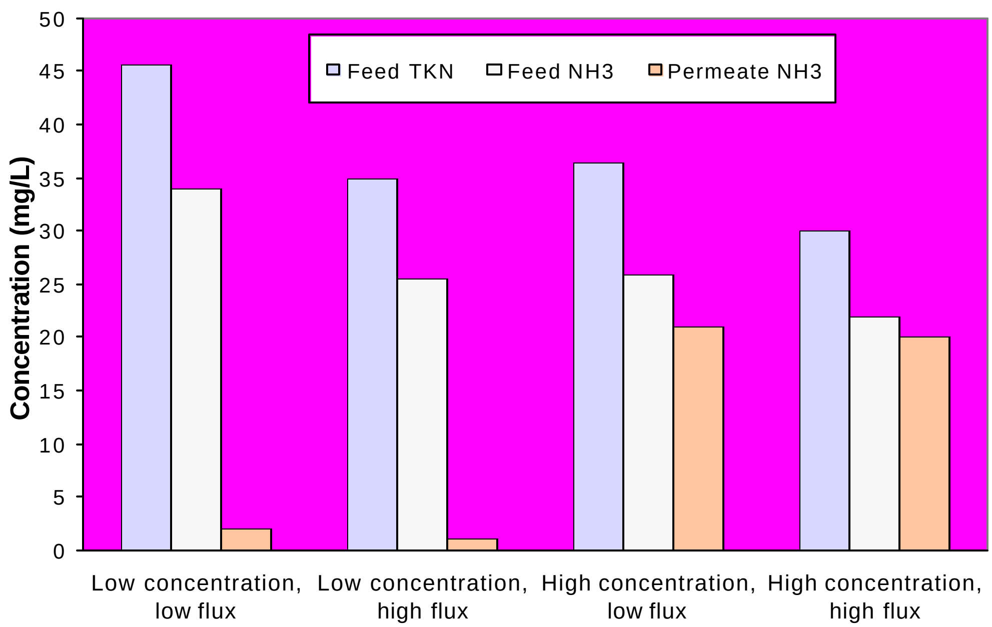


6.4 TKN and NH_3 Concentrations

Figure 25 presents the average feed TKN and both feed and permeate NH_3 concentrations for the four test conditions. The difference in the net amount of nitrification occurring in the low concentration and high concentration conditions is readily apparent. At the low reactor solids concentration, very little NH_3 remained in the permeate, so substantial nitrification had occurred. At the high reactor solids concentration, there was little difference between the NH_3 concentrations in the feed and permeate, so minimal net nitrification had occurred. These results were not affected by changing the membrane flux.

At least two possible factors could account for this difference in net nitrification levels. In the high reactor solids concentration tests, the reactor DO levels likely were lower than in the low concentration tests. When DO levels are below 1 mg/L, nitrification slows or ceases because oxygen becomes a limiting nutrient for the nitrifying bacteria. Another possibility is that the longer sludge age in the high solids concentration tests allowed nitrogen in the bacteria to be returned to the water through lysis and autooxidation of dead cells; this would reduce the apparent amount of nitrification that occurred. Either or both of these factors could be responsible for the difference in nitrification that was observed in these tests.

Figure 25. Comparison of Nitrogen Compound Concentrations (All Phase II Conditions)

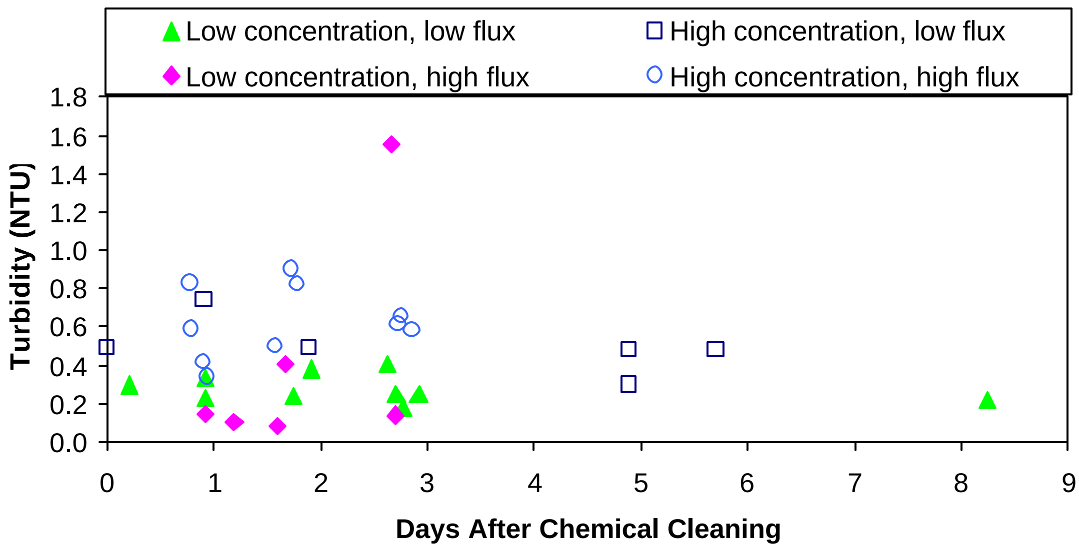


6.5 Permeate Turbidity and pH

Figure 26 shows the permeate turbidity as a function of the elapsed time after a chemical cleaning for all Phase II conditions. The permeate turbidity was not affected by the amount of time elapsed after chemical cleanings; this was true for all conditions.

A possible correlation between reactor concentration and permeate turbidity is seen in Figure 26. The high concentration conditions appear to have produced permeates with turbidities higher than the low concentration counterparts. During the low flux conditions, the permeate turbidity in the high concentration condition averaged 0.50 NTU, while in the low concentration condition the permeate turbidity averaged 0.28 NTU. During the high flux conditions, the permeate turbidity in the high concentration averaged 0.63 NTU, whereas in the low concentration condition the permeate turbidity averaged 0.16 NTU. Statistically there is a significant difference in turbidity between the high concentration and the low concentration conditions (Student's t-test, p <0.01); however, in practical terms no difference exists between the two as in virtually all cases the turbidity is less than 1 NTU.

Figure 26. Permeate Turbidity (All Phase II Conditions)

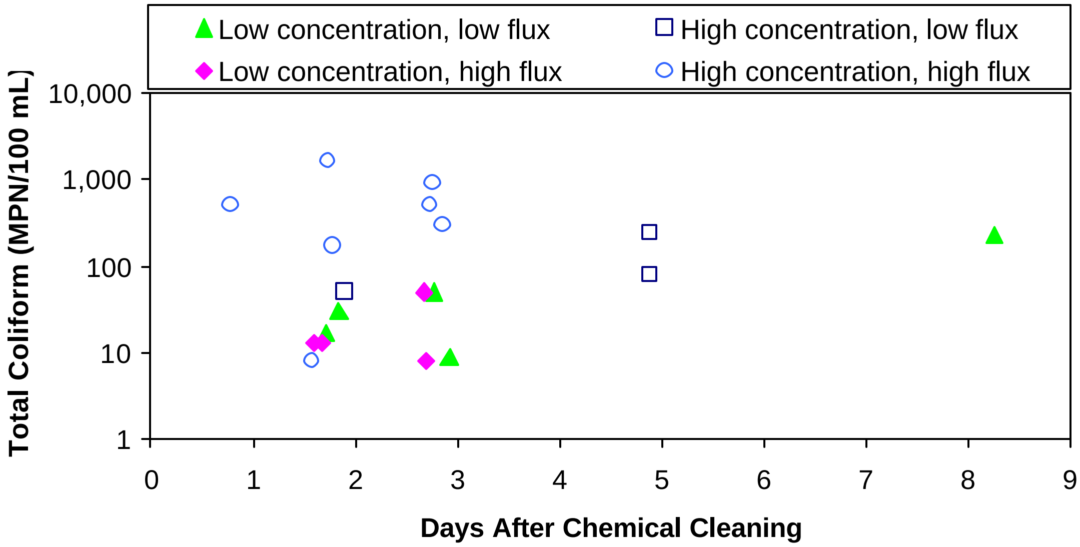


The permeate pH tended to be higher when the reactor solids concentration was higher. During the high concentration tests, the permeate pH averaged 7.97 (range: 7.75-8.10), while during the low concentration tests, the pH averaged 7.49 (range: 7.10-7.83). These averages are not statistically different (Student's t-test, p<0.05), although this result may reflect the small number of samples. Even if a pH difference exists, though, it does not seem to be of practical significance for OCSD's operations.

6.6 Permeate Total Coliform

Figure 27 shows the permeate total coliform readings as a function of the elapsed time after a chemical cleaning for the four Phase II conditions tested. The higher readings are attributed to leaks and defective seals that were found during and after the test period, although there have been reports of erratic bacterial removal by Zenon membranes in other agencies' tests, perhaps due to small areas of delamination between the polymer coating and the substrate. The lower readings are believed to be more representative of the true coliform removal capability of intact membranes. The lower readings correspond to about 6 log removal of total coliform from the feed.

Figure 27. Permeate Total Coliform vs. Days After Chemical Cleaning (All Phase II Conditions)



6.7 Air Requirements

Table 16 compares the aeration requirements of microfiltration and activated sludge treatment. The Table shows the specific air rate (the volume of air supplied per pound of cBOD removed) for the MF system in each of the four test conditions and the average for OCSD Plant 1's A.S. system during FY 1998-99.

The MF system used 3.3 to 5.6 times the air volume of the A.S. system. There are multiple reasons for this difference. The MF system uses air to agitate and clean the membranes during operation, which reduces the fouling rate, but it also increases the air rate beyond the amount that would be needed solely for the biological activity in the reactor. In addition, since OCSD tries to minimize nitrification during secondary treatment, the A.S. air usage is low compared to more typical plants that are required to nitrify and denitrify their effluent. The first factor increases the absolute aeration rate for the MF process, and the second factor increases the aeration ratio between MF and OCSD's A.S. process.

The aeration cost for a full-scale MF system would not necessarily be 3.3 to 5.6 times the aeration cost for A.S, however. If MF membranes were installed in existing aeration basins, they would not have to rest on the basin bottom; in fact, the membranes would only need to be at a depth that ensured they were fully submerged. Thus, while the A.S. air is introduced at a depth of about 16 feet, the MF air could enter at a depth of perhaps 8 feet, which would reduce the operating cost substantially. (Changing the depth from 16 feet to 8 feet reduces the adiabatic blower horsepower requirement by 46 percent.)

Table 16. Aeration Requirements For MF (As Tested) and A.S.

Condition	Zenon MF			OCSD Plant 1 A.S.	Ratio of Aeration Rates
	Air flow (ft^3/day)	Average cBOD removal (lb/day)	Air / cBOD removed (ft^3/lb)	Air / cBOD removed (ft^3/lb)	
Low concentration, low flux	230,400	53	4,350	775	5.6
High concentration, low flux		55	4,205		5.4
Low concentration, high flux	144,000	56	2,585		3.3
High concentration, high flux		48	3,003		3.8

After the Phase II tests were completed, Zenon announced its development of an air cycling system that is claimed to reduce the amount of air used in its MF systems. With air cycling, the blower runs intermittently rather than continuously. Zenon claims that the air usage will be reduced to 15 cfm per 650 ft^2 of membrane (compared to the 25 cfm per 500 ft^2 of membrane used in these tests), which would be a reduction of more than 50 percent. The air cycling system currently is undergoing testing by OCWD; if this system performs as promised, the MF air usage will be much closer to the air usage in a typical A.S. treatment system.

The effects of a shallower depth for the membranes and a reduced air flow with cycling can be combined to estimate their net effect on OCSD's A.S. energy costs, as shown in Table 17. The lower limit and upper limit Zenon air flows correspond to the low-concentration/high-flux and low-concentration/low-flux test conditions, respectively (as shown in Table 16). An internal OCSD power study in 1998, later updated to reflect the use of new air diffusers in the Plant 1 A.S. basins, reported the average A.S. power consumption to be 218, 256 kWh/month at Plant 1 and 1,137,036 kWh/month at Plant 2. Assuming an average electricity cost of 6 ¢/kWh, the A.S. electricity cost change could range from a reduction of $172,000/year to an increase of $378,000/year.

Table 17. Potential A.S. Energy Cost Savings with Improved Zenon MF

	Air Usage (ft^3 / lb cBOD removed)		Relative Blower Power Requirement [1]	Electricity Cost Savings ($/year) [2]
	Without Cycling	With Cycling		
Maximum Savings (low concentration, high flux condition)	2585	1195	0.824	172,000
Minimum Savings (low concentration, low flux condition)	4350	2010	1.39	(378,000) [3]

[1] At 8 ft depth with air cycling, compared to A.S. air flow of 775 ft^3/lb cBOD at 16 ft depth
[2] At 6 ¢/kWh
[3] Cost increase

6.8 Digester and DAF Effects

Zenon has reported a waste sludge rate of 0.6 lb TSS/lb feed BOD for a full-scale installation operating at a reactor concentration of 12,000 mg/L.[6] Based on FY 98-99 data, OCSD's A.S. process generates approximately 0.9 lb TSS/lb BOD, which implies a 34 percent reduction in secondary sludge to the digesters if MF were used instead.

Table 18 presents the effects of this reduced waste sludge rate on the operating costs of the digesters and DAFs. Using flow and cost information from FY 98-99, reductions in waste sludge rates of one-third would result in total operating cost savings at both plants of $363,000 annually.

[6] Mourato, D., *et. al*., 1999. "Upgrade of a Sequential Batch Reactor into a ZenoGem® Membrane Bioreactor" in *Proceedings of the Water Environment Federation 72nd Annual Conference and Exposition (WEFTEC '99)*, New Orleans, October 9-13, 1999.

Table 18. Digester and DAF Operating Cost Savings with MF

	Plant 1		Plant 2	
Primary sludge to digesters, ft^3/day	51,949		52,340	
Secondary sludge (TWAS) to digesters, ft^3/day	12,768		15,165	
TWAS fraction of digester feed	20%		23%	
Digester feed decrease with MF	7%		8%	
Digestion area operating cost, \$/ft^3 feed	0.09		0.09	
Annual digester cost savings with MF		$149,000		$177,000
DAF average feed flow, MGD	1.68		1.43	
DAF operating cost, $/MGD	$428		$443	
DAF feed decrease with MF	7%		8%	
Annual DAF cost savings with MF		$18,000		$19,000
Total annual savings with MF		**$167,000**		**$196,000**

Data sources: OCSD 1999 Annual Report (flows) and internal data warehouse (costs)

7.0 SUMMARY, DISCUSSION, AND CONCLUSIONS

Microfiltration of wastewater was demonstrated in a series of tests using the immersed membrane technology offered by Zenon. These tests included two wastewater feed types, two different membranes, and a variety of operating conditions. The goal of the tests was to explore the responses of the system to differing operating conditions and flux rates; these responses included changes in the quality of the permeate produced by the system and operational factors such as the fouling rate of the membranes. During much of the test period, the permeate was used as feed for an RO membrane system provided by OCWD so the operation of an RO system following the microfilter could be observed.

During the Phase I tests, Zenon's ADC membrane, with a mean pore size of 0.08 μm, was used. The feed source was unclarified stabilized mixed liquor exiting OCSD's A.S. aeration basins. The membrane was used essentially as an efficient clarifier: the TSS concentration around the membranes was targeted at only 1200 mg/L or 2400 mg/L, so relatively little solids concentration was occurring in the membrane reactor. This resulted in target sludge ages of only 4 or 8 hours.

The flux was varied from target values of 13.4 gpd/ft^2 to 28.0 gpd/ft^2 by varying either the permeate flow rate or the number of membrane cassettes in service. In all cases, the membranes were backpulsed for 30 seconds every 10 minutes and chemically cleaned with a bleach solution weekly.

For the Phase II tests, Zenon's OCP membrane, with a mean pore size of 0.035 μm, was used to test filtration of primary effluent in a membrane bioreactor (MBR). In an MBR, the membrane reactor becomes a complete secondary treatment system, acting as both an aeration basin and a clarifier. An advantage of an MBR is that much higher mixed liquor TSS concentrations can be used than in a conventional A.S. system. The target reactor TSS concentrations were 4000 mg/L and 12,000 mg/L in Phase II, which provided calculated sludge ages of 6 and 17 days, respectively.

Two target flux rates, 19 gpd/ft^2 and 29 gpd/ft^2, were used. As in Phase I, the backpulse schedule was set at 30 seconds every 10 minutes, but the chemical cleanings in Phase II were performed three times per week at Zenon's request.

The vacuum pressure and permeate flow readings were taken automatically and recorded by the system. Daily samples of the feed, reactor mixed liquor, and permeate were collected for various physical, chemical, and microbiological analyses. OCWD also sampled the MF permeate (as RO feed) and the RO permeate.

Phase I

In the Phase I tests, a clear positive correlation between the membrane flux and the operating vacuum pressure was seen. As the average flux approximately doubled (from 13 to 28 gpd/ft^2), the average vacuum approximately tripled (from 5 to 15 "Hg).

As expected, at each flux the vacuum increased over time, indicating fouling that reduced the membrane's permeability. The permeability was recovered by the weekly chemical cleanings.

The permeate analyses indicated excellent rejection of suspended material by the membranes. The permeate TSS concentration often was less than the detection limit (0.4 mg/L), indicating 99.9 percent TSS removal from the feed. Similarly, permeate cBOD readings most often were below the detection limit (4.0 mg/L), and the feed TOC was reduced 80 percent (which is of interest for tertiary treatment applications). These results were not affected by the reactor TSS concentration.

Although nitrification was not a goal of these tests, MF reduced the total amount of nitrogen-containing compounds between the feed and the permeate. Approximately 24 percent of the feed ammonia and 81 percent of the feed TKN were removed in the MF process. Increases in nitrite and nitrate levels from the feed to the permeate (to average values of 3 mg/L each) were observed.

Substantial rejection of microbiological organisms by the membranes was evident. The total coliform results indicated an average of 5.8 log removal, and the HPC results, which look at a wider variety of organisms, showed an average of 2.4 log removal. Concentrations of E. coli in the permeate averaged less than 1 MPN/100 mL.

Twice-daily virus (coliphage) sampling was conducted for five days. Of the nine samples taken, seven samples showed virus concentrations at or below the detection limit (4.5 PFU/100 mL). The results indicated 3 to 4 log reductions in virus concentrations between the feed and permeate.

Phase II

Since Phase II looked at a wider range of reactor TSS concentrations, any effects of concentration on the operating vacuum pressure should be more evident in Phase II than in Phase I. In fact, the results indicated that the reactor concentration in the range tested was not an important factor in determining the operating vacuum. While it might be expected that higher concentrations would lead to higher vacuum levels, this was not observed. Higher concentrations appeared to reduce the vacuum levels slightly; however, given the amount of scatter in the data and the relatively small differences observed, this statement should be viewed cautiously.

The operating vacuum was increased by higher flux conditions. Higher fluxes were associated with both higher initial vacuum levels and faster fouling. With a newly cleaned membrane, the vacuum at the high flux condition was 4 to 6 "Hg higher than at the low flux condition, and the fouling rates at the high flux conditions were 3 to 18 times faster than at the low flux conditions. This has important implications for a full-scale installation since the flux affects both the initial capital cost (the amount of membrane surface area needed) and the ongoing operating costs.

The permeate TSS and VSS concentrations generally were less than 1.0 mg/L and often were below the detection limit (0.4 mg/L). Similarly, the permeate BOD concentration generally was at or below the detection limit of 4.0 mg/L. These permeate values were not affected by either the reactor TSS concentration or the membrane flux.

The reactor concentration had a considerable effect on the net amount of nitrification that occurred. At the low reactor TSS concentration, very little ammonia remained in the permeate, so substantial nitrification had occurred. At the high reactor TSS concentration, there was little difference between the feed and permeate ammonia concentrations, so minimal net nitrification had occurred. These results were not affected by the membrane flux.

Two possible explanations for the nitrification results were noted. Since a high reactor TSS concentration probably corresponded to a low reactor DO level, the oxygen could have been a limiting nutrient for the nitrifying bacteria, thus limiting the amount of nitrification that occurred. It also is possible that the longer sludge age in the high concentration operation allowed nitrogen in the bacteria to be returned to the water through lysis and autooxidation of dead cells; this would reduce the apparent amount of nitrification in the overall process.

For all test conditions, the permeate turbidity did not change during the time between chemical cleanings. A relationship was seen between high reactor TSS concentrations and higher permeate turbidity, but the practical significance of this is questionable. In all cases, the permeate turbidity was less than 1 NTU.

The coliform rejection performance of the system was inconsistent. At its best, the system showed approximately 6 log coliform removal (to <10 MPN/100 mL), but some permeate samples had coliform concentrations exceeding 1000 MPN/100 mL. Various piping leaks and defective seals were discovered during and after the tests, so these may be responsible for the higher readings. It was noted that other agencies reportedly have experienced erratic bacterial removal with Zenon membranes, though, which might be due to small areas of delamination between the polymer coating and the substrate (although this explanation is speculative and has not actually been observed). Good coliform rejection should be achievable with intact membranes.

A large part of the operating costs for a Zenon MBR system is due to the power required for the aeration system. Air is used both for biological stabilization and for membrane scouring, and the scouring requirements dominate. This means that the MBR system always will use more air than a well-operated A.S. process accomplishing the same amount of treatment. In these tests, the aeration rate was kept essentially constant at about 25 cfm/module; this was the minimum value Zenon would consider for a full-scale installation. Depending on the test condition, this corresponded to specific air rates of 2500 to 4300 ft^3 per pound of cBOD removed from the feed, which was 3.3 to 5.6 times the air volume used in OCSD Plant 1's A.S. system.

Since OCSD tries to minimize nitrification, its air usage is low compared to more typical A.S. installations. In addition, since the air in an MBR could enter at a shallower depth than in an A.S. system, the operating cost difference would not necessarily correspond to the absolute air flow difference. Nevertheless, the aeration cost would be higher for an MBR than for A.S.

After the Phase II tests were completed, Zenon announced its development of an air cycling system in which the membranes are aerated intermittently rather than continuously. Zenon claims that this would reduce the air required for MF more than 50 percent when treating primary effluent (from 5 cfm/100 ft^2 of membrane as tested in Phase II to 2.3 cfm/100 ft^2 of membrane). The air cycling operation currently is being tested by OCWD. If it performs as promised, then Zenon's MF air usage will be much closer to the air usage in a typical A.S. system.

The combined effects of a shallower depth for the membranes and a reduced air flow with cycling were estimated. Depending on the operating conditions (reactor concentration and membrane flux) chosen, the MF blower power usage could be more or less than the existing A.S. power usage. The maximum MF cost savings occur with low-concentration/high-flux operation. With electricity valued at 6¢/kWh, MF could reduce OCSD’s electricity costs by $172,000/year.

If MF were installed in place of conventional A.S., it would affect the downstream solids handling processes (DAF and anaerobic digestion) also. Zenon has reported a waste sludge rate for a full-scale installation that is one-third less than OCSD's A.S. WAS rate. If this reduction were applied to OCSD's operations, annual cost savings of about $363,000 in the total DAF and digestion operations would be realized.

It was not in the scope of this study to estimate full-scale capital or operating costs for an MF system. For the GWRS program, though, Zenon has provided bid, not estimated, costs for a very large installation, so that information is available and is much more reliable than any estimates that could have been developed in the course of this study.

Much of the interest in MF has been related to its role in the GWRS project, for which it is expected to be a key part of the complete treatment train. As GWRS continues into its second and third phases, the unique characteristics of MBRs may be increasingly important. By combining secondary treatment of high TSS mixed liquor and clarification into a single process, MBRs provide effluent stabilization while requiring much less land area than conventional A.S. processes. Also, microfiltration produces an effluent that can be fed directly to an RO process, which eliminates the intermediate filtration that is needed for A.S. effluent. Finally, as mentioned previously, the results of this project indicated that energy costs can be less with an MF MBR (at appropriate operating conditions) than with A.S.

There is a major regulatory barrier that must be addressed before MF MBRs could be used in place of conventional secondary processes in water reclamation. In the regulators' concept of "barriers" for reducing the amounts of microbiological organisms in the water, each processing step is seen as a barrier to accidental process failure and contamination of the product water. By combining secondary stabilization and clarification into one step, there is one less barrier in the process train. Even though the coliform and virus removal with MF exceeds that of A.S. with clarification (see below), regulators have not yet accepted membrane systems as equivalent to conventional secondary treatment systems.

MF also has other possible applications for OCSD. Ocean discharge limits might be met by blending the high quality effluent produced by MF with filtered primary effluent at a lower total cost than is possible now. The insensitivity of the MF effluent quality to changes in operating conditions might make MF a good candidate for use in high flow situations; the MF could act as a buffer, absorbing changes in influent flows and solids loadings, undergoing membrane flux and reactor solids concentration adjustments as necessary, but producing a uniform quality permeate throughout. This could be especially valuable if the permitted emergency discharge points (the 78" outfall and the Santa Ana River) ever had to be used.

OCSD's current ocean discharge permit limits the microorganism levels that are detected three miles offshore (rather than at the outfall exit). During emergencies, these discharge limits would not be enforced, but using MF would substantially reduce the microorganism levels in the effluent. Similarly, if there ever were a need to use an alternative outfall routinely, the microorganism reductions with MF could be critical for meeting the ocean discharge limits.

Historical OCSD data show the advantages of MF over A.S. in reducing indicator microorganism concentrations.[7] As Table 19 shows, analyses of A.S. influent and

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[7] Prepublication data from Yahya, M., *et. al.*, 2000. "Fate of Indicator Microorganisms during Primary and Secondary Wastewater Treatment Processes" in *Proceedings of the Water Environment 73rd Annual Conference and Exposition (WEFTEC 2000)*, Anaheim, October 14-18, 2000.

clarified effluent samples indicated average log reductions of 1.6 in total coliform concentration and 2.6 in coliphage (virus) concentration. In contrast, MF in the current project demonstrated up to 6 log reduction in total coliform concentration and up to 4 log reduction in coliphage concentration.

Table 19. Comparison of A.S. and MF Bacteria and Virus Reductions

	Activated Sludge	Microfiltration
Total Coliform		
Average Effluent, MPN/100 mL	14,000,000	<100
Log Reduction Across Process	1.6	up to 6
Coliphage		
Average Effluent, PFU/100 mL	4700	<4.5
Log Reduction Across Process	2.6	up to 4

Note: MF coliform values are from Phase II and coliphage values are from Phase I (since Phase II did not include virus testing).

Conclusions

The following overall conclusions can be drawn from the results of this test program:

- The Zenon ADC membranes in Phase I operated acceptably when treating unclarified secondary effluent with reactor concentrations of 1200 to 2400 mg/L and fluxes between 13 and 28 gpd/ft^2. The permeate often had TSS and cBOD levels at or below their detection limits (0.4 mg/L and 4.0 mg/L, respectively).

- The ADC membranes successfully rejected large fractions of the microbiological organisms in the feed. The total coliform tests indicated an average of 5.8 log removal, and the HPC tests showed an average of 2.4 log removal. The virus tests averaged 3 to 4 log removal.

- The OCP membranes in Phase II operated acceptably when treating primary effluent with target reactor concentrations of 4000 and 12,000 mg/L and fluxes of 19 and 29 gpd/ft^2. The permeate TSS and BOD levels often were at or below their detection limits regardless of the reactor concentration or membrane flux.

- The operating vacuum in Phase II was increased by higher flux conditions. Higher fluxes were associated with higher initial vacuum levels and with faster flux increases during operation (that is, faster fouling).

- The permeate turbidity in Phase II stayed below 1 NTU regardless of the test conditions.

- The coliform rejection by the system in Phase II was inconsistent. In some instances, the permeate coliform concentrations were below 10 MPN/100 mL (approximately 6 log removal), but other samples had coliform concentrations two orders of magnitude higher. Several known mechanical problems probably allowed some permeate samples to become contaminated, but there also might have been problems with the membranes themselves. Intact membranes and peripheral equipment should provide substantial bacterial removal.

- In a full-scale installation with the membranes suspended in aeration basins and air cycling being used to reduce the total aeration requirement, the MF blower power usage could be more or less than the existing A.S. power usage, depending on the MF operating conditions. The maximum MF aeration cost savings would occur with low-concentration/high-flux operation. With electricity valued at 6¢/kWh, MF could reduce OCSD's electricity costs $172,000/year.

- Replacing an A.S. system with an MBR would affect the DAF and anaerobic digestion processes also. It is estimated that the annual operating costs for these areas would be reduced $363,000 if MF were used.

- During Phase I and much of Phase II, OCWD fed an RO with the MF permeate. The RO reportedly performed well, producing a satisfactory quality permeate. The fouling rate was somewhat faster than has been experienced with other feeds, but the cycle time between cleanings was within acceptable limits.