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FILTRATION OF PARTICULATES AND EMULSIONS WITH A PLEATED, THIN CHANNEL, CROSS-FLOW MODULE

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ABSTRACT

The pleated thin channel crossflow module represents a novel flat sheet membrane configuration combining the modular advantages of hollow fiber design with the hydrodynamics of plate and frame units containing turbulence promoting elements. It consists of a circular array of parallel pleated thin channels, each of whose inner wall is a membrane supported by a porous backing resting on a central product drain tube. The outer flexible wall is separated from the membrane by a turbulence promoting spacer and serves to create the flow channel, without fulfilling a direct filtrative purpose.

The design allows operation at elevated particulate levels and is highly suited to microfiltration of stray matter and bacteria from aqueous solutions to be injected into well heads for enhanced oil recovery. Data are presented for 0.45 and 0.2 µm pore size membrane modules containing 3.3 ft^2 of membrane area and operating on dispersions of standard particulates of known size distribution (80 µm to 0.2 µm). Effects of filtration mode (i.e., constant pressure or constant permeate rate), and crossflow velocity were examined at particulate concentrations from 250 to 800 ppm.

For separation involving paraffinic oil/water emulsions, modules are either fabricated with ultrafiltration (UF) membrane, or a dynamically formed membrane is created on the microporous support. The latter approach was used to purify waste textile scouring washwater at 60°C, thereby recovering both water and energy. Waste scouring water, containing $\sim$1200 ppm of emulsified total organic carbon (TOC), 500 ppm of freon extractable hydrocarbon, and a turbidity of 66 nephalometric turbidity units (NTU) was ultrafiltered at 50 to 60°C with a dynamically formed silica membrane at an average pressure of 17 psi and 1.6 gpm crossflow in a unit containing 1 ft^2 of 0.2 μm Acropor support. The product flow rate was 35 to 50 gpd and contained 400 ppm TOC with 125 ppm freon extractable hydrocarbon and a turbidity of 0.5 NTU. Yarn samples scoured with recovered scouring water showed no difference in their dyeability.

INTRODUCTION

In applying membrane separation technology to energy related problems, success or failure may depend as much (or more) on the method of packaging the membrane as on the separation properties of the barrier itself. For flat sheet membrane, the conventional designs have been either plate and frame devices or the spiral wound module. However, both these configurations suffer from fundamental disadvantages, especially for operation at low pressures and with high levels of particulates in the feed.

Plate and frame units, in addition to lacking the advantage of a modular, sealed construction, usually possess long path lengths for the feedstream. Since thin channel systems require elevated feed flow velocities to overcome concentration polarization, this necessarily leads to high transchannel pressure drops between the feed entrance and exit. When turbulence promoting spacers are placed in the channel, the units require lower feed flow pumping rates, but their tolerance to particulates in the feed stream becomes quite limited.

Spiral wound elements do possess the convenience of a sealed modular unit and generally possess shorter feed flow channels. However, the larger area devices especially suffer from the pressure drop incurred in the permeate compartment, as the spiral permeate path resistance to the drain tube becomes quite large. Furthermore, spiral wound units are also quite sensitive to clogging by particulates and are not recommended for use with solutions containing more than 25 ppm of suspended solids or a turbidity of >1.0 NTU (1).

A novel configuration, which attempts to improve on the disadvantages just cited, is the pleated thin channel crossflow module shown in Fig. 1. Through the pleating process, a multiplicity of parallel thin flow channels is formed and arranged in a circular array about the porous support tube E, which provides

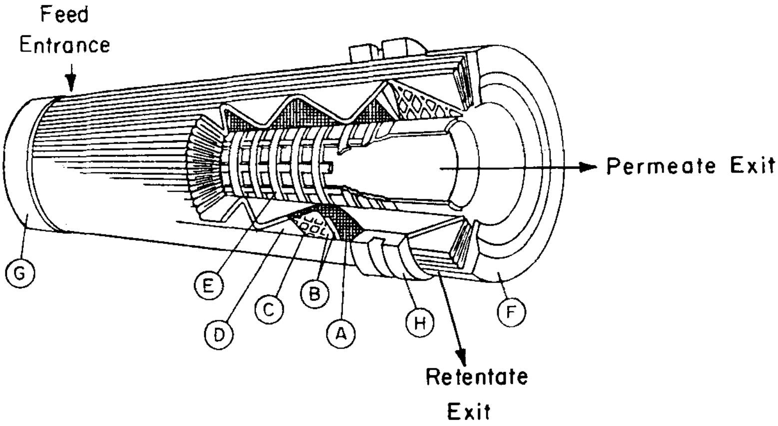

FIGURE 1. The pleated thin channel crossflow module. A, porous pleated support screen meant to provide mechanical support under the applied pressure; B, pleated microporous membrane filtration element; C, pleated spacer which creates the thin flow channel as well as promoting turbulent flow; D, impermeable film which creates the flow channel. A space between the ends of film D and the module glue seals is left so as to allow the entrance and exit of the fluid through the flow channel; E, porous internal support tube to provide an exit for permeate; F, closed end cap which completely seals one end of the module; G, closed end cap which completely seals the opposite end of the module; H, outer seal ring which creates the seal between the impermeable film in the module and the interior of the housing. The ends of the module are potted and sealed in end caps F and G with a water resistant and/or solvent resistant glue. The module is 10 inches long and has a membrane area of 3.3 ft^2.

the permeate compartment drain and exit. The inner wall of the flow channel is formed by the membrane B which is supported by a porous support screen A. The outer channel wall D is a flexible film which serves no direct filtrative purpose. Its presence serves to create the flow channel between itself, the turbulence promoting spacer C,and the membrane wall A. Unlike elements A, B, C, and E, the outer channel wall D does not extend into the potted end caps, F and G. Instead, a space of ∿7 mm exists between the end cap F and the film D thereby forming the entrance to the flow channel. The outer channel wall film D ends just beneath the seal ring H, so that a space for retentate exists between H and end cap G. The space in the depth of the pleat between adjacent pleated flow channels is filled with glue beneath the outer seal ring H. Since this ring also creates an O-ring seal to the stainless steel housing (Fig. 2), fluid entering the housing cannot bypass the cartridge and must enter the flow channels.

A most obvious advantage of the pleated design is the attainment of a small pressure drop between the membrane and the permeate drain tube, at the same time that the crossflow channel pressure drop is maintained equal to that in the spiral module design. However, the flexible outer channel wall also provides additional advantages regarding (a) the use of the module on solutions heavily loaded with particulates and (b) the cleaning of the module by backflushing from the permeate to the feed compartment.

In the normal mode of operation, the space between channel wall D and the housing becomes filled with fluid which is pressurized to that of the channel entrance pressure. Since some pressure drop exists along the length of the flow channel, the pressure in the space above D is actually greater than that in the flow channel itself. This pressure differential acts to seat the flexible channel wall D firmly in the pleats and maintains the channel dimensions as defined by the spacer. However, at times when debris

FIGURE 2. Crossflow module and housing.

clogs the spacer or during the backflush operation, the outer channel wall is effectively free to move and thereby increase the channel height. This design allows the unit to handle particulate loadings as high as 2000 ppm, without irreparable clogging of the flow channel.

In the work which follows, we shall examine the application of the pleated crossflow module to filtration problems involving particulates and emulsions. Such problems are an integral part of processes for either the increased production of oil supplies, e.g., by means of enhanced oil recovery, or the conservation of both water and energy by the recycling of large volumes of moderately warm industrial wastestreams.

EXPERIMENTAL

Filtration of Particulates

Dispersions of air cleaner (AC) fine test dust (GM Corp., ACS Division, Flint, Michigan) in prefiltered tap water were used as the standard particulates feed, which was made up to a total volume of 520 L in the filtration apparatus depicted in Fig. 3. Experiments were carried out in two filtration modes, (a) constant permeate and retentate flow, and (b) constant filtration pressure and retentate flow [the conventional mode for ultrafiltration (UF) and reverse osmosis (RO) processes].

Prior to the addition of AC dust in the tests at constant permeate and retentate flow, the transchannel pressure drop, ΔP_c =

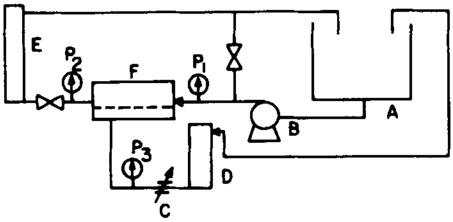

FIGURE 3. Apparatus for filtration of particulates at constant flow rate or pressure. A, 500 L SS tank; B, 80 Lpm centrifugal pump; C, automatic flow rate controller (W. A. Kates Co., IL); D, 0.40 Lpm flowmeter; E, 0 to 80 Lpm flowmeter; F, crossflow module and housing; P_1, P_2, P_3, pressure gages.

$P_1 - P_2$, was measured as a function of the retentate flow rate Q_R for a constant permeate rate, Q_P, of ∿20 Lpm.

After stopping the permeate flow and the addition of the requisite amount of AC dust, the total feed rate Q_F was set to an initial value equal to $Q_R + Q_P$. The valve controlling Q_P was then opened to the desired rate and measurements of $P_{1,2,3}$ were taken as a function of time. The average transmembrane pressure drop, P_m, was calculated according to

$$\Delta P_m = \frac{(P_1 + P_2)}{2} - P_3. \tag{1}$$

The tests at constant pressure and retentate flow were initiated by setting the feed rate Q_F and the average transmembrane pressure at approximately their desired values with the permeate valve closed. When the permeate valve was opened, the measurement of Q_P as a function of time was begun and adjustments of valves controlling pressure and Q_R were made throughout the run to maintain them as constant as possible. Since Q_P can represent a considerable fraction of Q_F in the initial period, some variation in conditions during the first ten minutes was unavoidable.

Particle Size Distribution and Concentration Measurements for AC Fine Test Dust

The particle size distribution for one lot of AC fine test dust has been carried out by sedimentation analysis (2), and is reproduced in Fig. 4. A calibration curve for particulates concentration was developed by subjecting standard samples to 20 min of sonication in a Bransonic Ultrasonic cleaning bath and measuring their turbidity with a Hach 2100 A turbidimeter. Plots of log turbidity vs log ppm standard in the range of 25 to 100 ppm yielded straight lines whose slopes were highly dependent on the dispersion conditions utilized.

SiO_2 Dynamically Formed Membranes

Experiments involving dynamically formed silica membranes were carried out on 6-in. laboratory size crossflow capsules

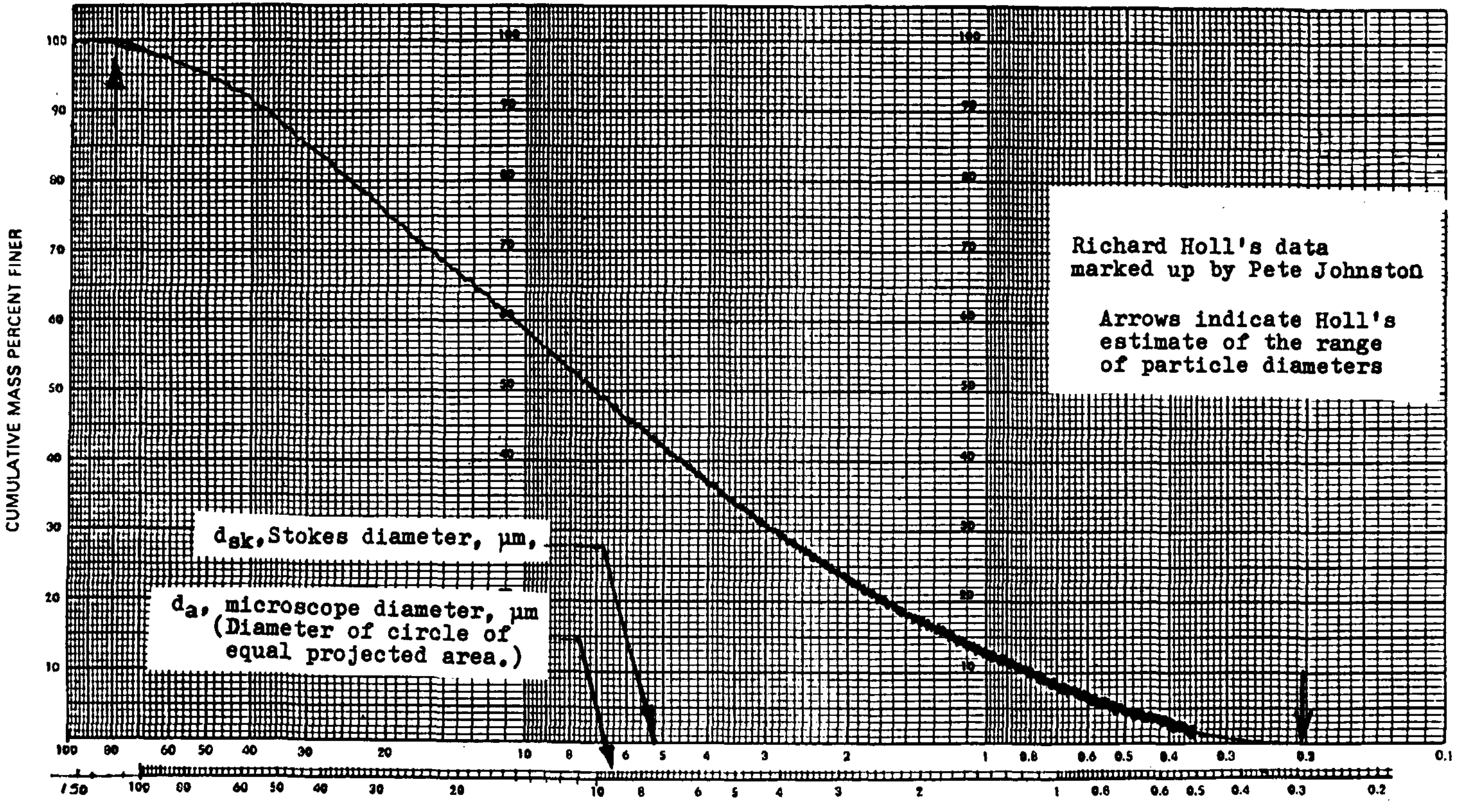

FIGURE 4. Particle size distribution of AC fine test dust.

containing ∿10^3 cm^2 of 0.2 μm Acropor membrane. These modules were fabricated with an integral plastic housing, but were otherwise functionally identical to the crossflow module and housing of Figs. 1 and 2.

Since the cleanliness of the system is very important to dynamic membrane formation, an all-polypropylene tubing and valve apparatus was constructed as shown in Fig. 5. In order to use a backwash mode first suggested for hollow fibers (3), the manifolding in the system is such that the feed and product can be switched back and forth with the permeate valve closed. Only the pump body is stainless steel, and the flowmeter is glass.

A 500-ppm solution of Cab-O-Sil EH-5 silica (Cabot Corp.) of 8-liter volume was obtained by diluting 2 liters of a 2 g/L solution which had been sonicated for 40 min with a Branson ultrasonic bath. The turbidity of the diluted feed was 24 NTU. Deionized (DI) water which had been filtered through a 0.2 μm Acroflow cartridge was used to prepare the solution and for the final washout of the system. (The quality of the water used, especially from the

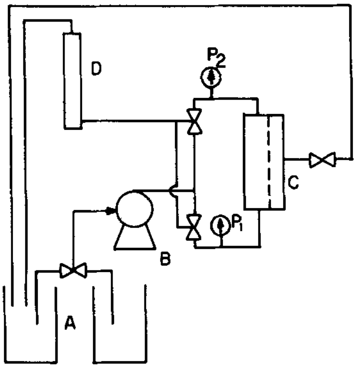

FIGURE 5. Apparatus for dynamic membrane formation and ultrafiltration of yarn scouring wastewater. A, 10 L tanks; B, Flotec R-3 pump; C, crossflow capsule; D, flowmeter; P_1, P_2, pressure gages. The system was fabricated from polypropylene fittings, valves and tubing.

point of view of its microorganism and particulates content, is critical to the reproducible achievement of good dynamically formed membranes).

With the feed entrance and exit in the normal mode, the formation was carried out at ∿25°C, an average pressure of 1 atmosphere and a retentate flowrate of 6.3 L/min for a period of 30 min. During the formation period, permeate flow rate and feed and permeate turbidity were monitored. The silica feed was then flushed from the system with DI water and replaced by the yarn scouring feed.

A Beckman 915A Total Organic Carbon (TOC) Analyzer was used to determine the TOC of feed and product samples. Extractables were measured by extracting a 1-liter acidified sample with three ∿30 ml samples of freon FT, distilling off the freon and weighing the residue. Suspended solids (feed) were measured by weight increase after hot filtration through a 0.45 μm membrane, while total solids were determined by drying at 100°C to constant weight.

RESULTS AND DISCUSSION

Enhanced Oil Recovery and Filtration of Particulates

Enhanced oil recovery by various water flooding techniques requires the subsurface injection of aqueous solutions into oil-bearing geological formations. In their filtration properties, these porous formations resemble expensive depth filters, which can be irreparably damaged if clogged by colloidal particles and bacteria (4). Current treatment practice usually involves conventional water treatment techniques, simply because the types of membrane filters commercially available at present have too limited a lifetime and are therefore very costly.

On the other hand, the crossflow module operates at quite low pressures, and would be expected to possess an extended lifetime since it can be backflushed. Thus it may be suitable for simplifying the treatment of solutions meant for subsurface injection. In earlier work (5), crossflow modules containing 5 μm and 1.2 μm pore size membranes were used to separate scleroglucan organisms from fermentation broth to be diluted and subsurface injected as a viscosity control agent. In turning our attention toward the treatment of large volume injection streams, we chose AC fine test dust as the standard particulate contaminant for our test purposes. As Fig. 4 indicates, the distribution of particle sizes is quite broad and ranges from 80 µm down to 0.2 µm.

Since the enhanced recovery process requires a constant fluid pumping rate, the first filtration mode examined was the constant permeate condition, illustrated in Fig. 6. This type of arrangement, while common for conventional cartridge filters, is a little unusual for a tangential flow system. Typically, ultrafiltration and reverse osmosis operate in the constant pressure mode (also illustrated in Fig. 6). However, the constant permeate mode is probably more suitable for the microfiltration situation, in which the membrane resistance is far less than that of the boundary layer which forms from retained particulates. If cake filtration is dominant (i.e., particles are retained as a built-up cake on

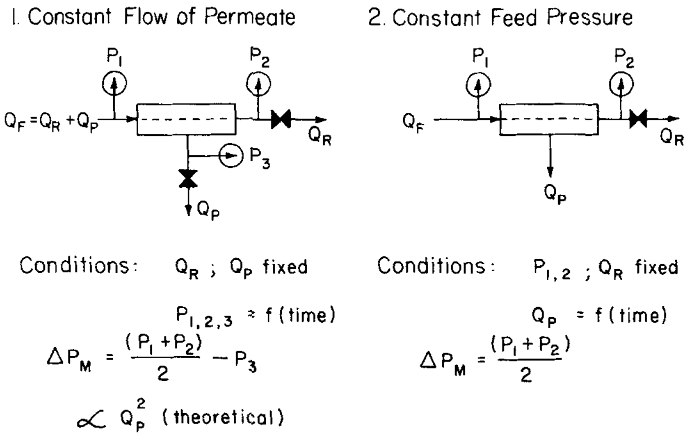

FIGURE 6. Two modes of filtration module operation. Mode 1, constant flow of permeate; mode 2, constant feed pressure.

the surface, as opposed to internal clogging of the pores), then the average transmembrane pressure drop ΔP_m should increase with time according to the equation (6)

$$\Delta P_M(t) = P_M(o)[K_c\, Q_p^2 t + 1], \qquad (2)$$

where K_c is the cake filtration constant, which is directly proportional to the particulates concentration, and $\Delta P_M(o)$ is the pressure drop at time t = 0.

For the present purposes, Eq. (2) is at best only applicable to the early stages of filtration, since it was derived for classical "dead-ended" filtration, without tangential flow which tends to limit cake growth.

As shown in Fig. 7, the filtration behavior with 0.45 µm membrane crossflow modules does appear to follow Eq. (2) in the early stages. However, the data for a permeation rate of 5 gpm are clearly approaching an asymptotic value. When the initial rate of pressure increase is plotted on a log-log basis versus the permeate flow rate (Fig. 8), one obtains a straight line with a slope of 1.3,

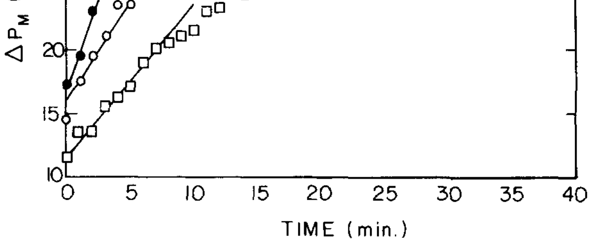

FIGURE 7. Filtration mode 1: transmembrane pressure drop vs time. Permeate flow rates: ● - 10 gpm; ○ - 7 gpm; □ - 5 gpm. Membrane pore size 0.45 µm; Q_R = 4 gpm; 250 gpm AC fine test dust.

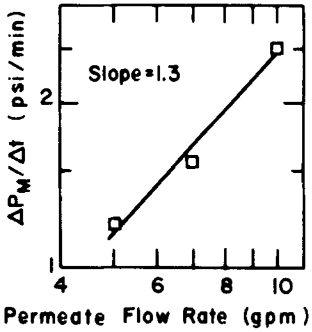

FIGURE 8. Initial rate of change of transmembrane pressure vs permeate flow rate.

and not the slope of 2.0 anticipated from Eq. (2). The dependence on Q_P^2 in Eq. (2) is the result of an assumption in conventional cake filtration that the solids content of the growing cake boundary layer per unit time is simply given by the bulk solids concentration multiplied by the permeate flow rate per unit area. However, in the presence of turbulent crossflow, one can no longer make such an assumption, since the stirring would be expected to either remove some of the solids deposited or prevent their deposition. The reduced slope of the data presented in Fig. 8 is at least qualitatively in agreement with such expectations.

The initial rate of increase of the transmembrane pressure is shown plotted against the retentate flowrate Q_R (at a constant permeate flow rate) in Fig. 9. As anticipated, increasing the retentate flow rate improves the mixing and decreases the rate of cake growth, a fact which is reflected in the reduced rate of increase of the transmembrane pressure. Under conditions of a low suspended solids concentration and a recovery ratio of $Q_P/Q_F = 0.67$, one observes that $(\Delta P_m/t)$ is proportional to $Q_R^{0.25}$. Such a relation is very interesting, because it has been found (7) for conditions of constant pressure in a laminar flow channel that $Q_P(t)$ is proportional to $Q_R^{0.3-0.4}$. On the other hand, the two data points at the higher solids concentration, but lower recovery ratio of

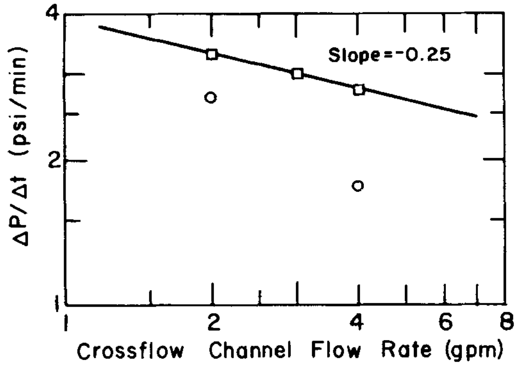

FIGURE 9. Initial rate of change of transmembrane pressure vs crossflow channel flow rate. Conditions: □ , Q_P = 10 gpm; 250 ppm AC fine test dust; ○, Q_P = 5 gpm, 1000 ppm AC fine test dust.

Q_P/Q_F = 0.3, suggest the much stronger dependence of 0.8 to 0.9 which one would anticipate for turbulent flow conditions (8).

To examine the behavior of the crossflow modules in the constant pressure mode of operation, 0.2 μm and 0.45 μm membrane modules were tested at initial feed particulate concentrations of 750 to 800 ppm at a pressure of ∿3 atm and a retentate flowrate of 5 gpm. As shown by the data presented in Fig. 10, there is almost no difference between the 0.2 μm and 0.45 μm pore size membrane modules. In this experiment the amount of particulates retained on the membrane surface was measured by the weight increase of the cartridges, which was 158 g and 136 g for the 0.2 μm and 0.45 μm modules, respectively. Dividing by the membrane area and assuming a cake porosity of at least 50%, one calculates that the cake thickness was the order of 500 μm, which is very large compared to the 650 μm thickness of the spacer. Since the crossflow channels were not clogged, nor the pressure drops excessively large, this demonstrates the utility of the flexible channel wall which apparently adjusted its height as the cake thickness increased.

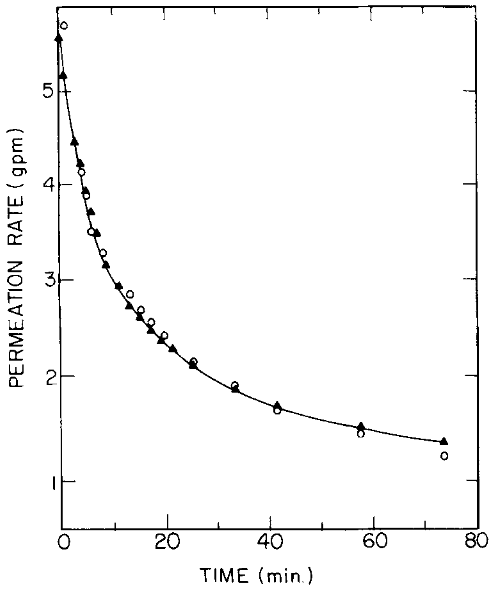

FIGURE 10. Filtration mode 2: permeate flow rate as a function of time for ○ - 0.45 μm and ▲ 0.2 μm pore size membranes at 44 ± 2 psi pressure and Q_R = 5 gpm. AC fine test dust concentration: 800 ppm initial, 500 ppm after 70 min.

With regard to the quality of the permeate from all the experiments, the turbidity was consistently better than 0.3 NTU and often less than 0.1 NTU. Thus, in principle, the application of pleated crossflow microfiltration units to filtration for enhanced oil recovery seems possible. However, only field trials with systems designed specifically for the filtration will determine whether or not they will in fact be utilized.

Ultrafiltration of Yarn Scouring Water Effluent

As often happens, the principles of water conservation and pollution control are also consistent with energy conservation. Yarn scouring water effluent, generated in the textile treatment process, is an oil/water emulsion containing additional non-emulsified free oil, at a temperature of 50 to 60°C. Due to its relatively low temperature, heat exchangers provide little gain, and at one particular plant the stream is simply discharged to a pond. However, if the water were ultrafiltered and recycled, the economics with respect to the cost of the energy saved becomes quite impressive, due to the large volumes of liquid involved. Furthermore, since the paraffin spinning lubricant (removed from the yarn by the scouring process) congeals into a waxy material at temperatures below 45°C, a large fraction of the concentrated emulsion retentate could be separated to yield a reasonably good fuel.

Earlier reports (9, 10) of scouring water ultrafiltration involved the use of conventional asymmetric UF membranes. However, since the major problem encountered involved flux decline due to membrane fouling, we chose to examine the dynamically formed membrane approach. Dynamically formed membranes are created by the in situ filtration of colloidal or polymeric species which form selective secondary layers on various types of supports, depending on the "class" of dynamic membrane (11). (A more detailed description may be found in a recent review (11) and its associated references.)

It has recently been observed (12) that colloidal silica particles 70 to 100 Å in diameter form, upon filtration through 0.2 µm microporous supports, a "Class II" dynamic membrane. This was found to be an excellent ultrafilter for concentrating oil/water emulsions used in metal cutting and grinding operations. Since this process was successfully carried out at room temperature in pleated crossflow modules (12), it seemed likely that the same system might also handle the higher temperature yarn scouring effluent emulsion.

In Table 1 the progress of the SiO_2 dynamic membrane formation is recorded for a pleated crossflow module containing 0.1 m^2 of 0.2 µm pore size membrane. The amount of silica removed from the feed by the deposition process is represented by the small drop in

TABLE 1

Dynamic Formation of an SiO_2 Ultrafiltration Membrane in a Pleasted Crossflow Module. (Module Area - 0.1 m^2; pressure - 1 atm; retentate flowrate = 6.3 L/min; SiO_2 concentration - 500 mg/L).

Time (min)	Flow rate (cc/min)	Turbidity (NTU)	
		Feed	Product
2	600.0	24.0	0.37
3.5	571.4	-	-
5	545.4	23.0	0.4
10	500.0	22.5	0.4
15	461.5	22.5	0.23
20	461.5	22.5	0.25
25	444.4	22.5	0.25
30	421.0	22.5	0.15

the feed turbidity from 24.5 NTU to 22.5 NTU which, by calculation, represents 4 mg/cm^2 of SiO_2. Utilizing the known density of 2.39 g/cm^3 for silica and the fact that such layers usually have a porosity of 0.7 to 0.8, we estimate that the dynamic membrane thickness is 40 to 50 μm, which would conform roughly to previous observations with hydrous Zr(IV)-oxide (13). It is also interesting to note that, although the membrane flux continues to decline after the 10-min mark, the feed turbidity remains constant, indicating that no silica is being removed. In addition, the drop in permeate flowrate is matched by increased silica retention, as shown by the decrease in the product turbidity (Table 1). Thus we may conclude that, after the first 10 min, the flux decline represents cake consolidation and a decrease in the effective pore size of the dynamic membrane.

The data presented in Fig. 11 and Table 2 summarize the results of ultrafiltering the yarn scouring water with the module containing the silica dynamic membrane. Although the TOC content of the permeate remains constant throughout the experiment, the permeate flowrate shows an interesting behavior as a function of time and temperature. In the first ten hr period at 50°C, severe

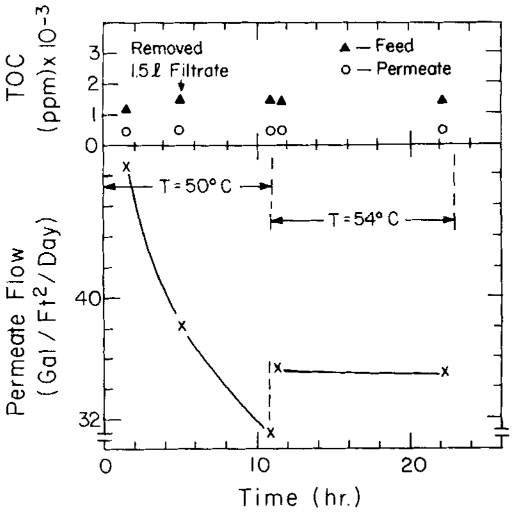

FIGURE 11. Ultrafiltration rate as a function of time for an SiO_2 dynamic membrane tested with yarn scouring water. Pressure - 15 psi; Q_R = 1.6 gpm; module contains 1 ft^2 area, 0.2 μm membrane.

TABLE 2

Composition of Feed and Permeat in Yarn Scouring Effluent Recovery with a Dynamically Formed SiO_2 Membrane

	Effluent Feed	Permeate
Total solids (ppm)	2640	1650
Suspended solids (ppm)	640	0
Total organic carbon (ppm)	1200	400
Total extractables (ppm)	500	125
pH	8.9	8.9
Conductivity, millimho at 60°C	3.72	3.70
Turbidity (NTU)	66-72	0.5

flux decline is seen to occur and the steep slope suggests that further decline would be evident had the conditions remained the same. In the second 10-hr period, the temperature was raised to 54°C, a change which not only raised the flowrate by 15% but also stabilized its performance with time. In the third part of the experiment, the temperature was raised to 60°C and the concentration of the feed was raised periodically by removing several liters of permeate and allowing the system to come to a steady state flux over a 2-hr period. These data are shown in Fig. 12, and it may be seen that at lowest the feed concentration (corresponding to that in Fig. 10), raising the temperature from 54°C to 60°C has caused the permeate flowrate to increase from 35 gfd to 49 gfd. Nonetheless, the TOC of the permeate remained constant at its earlier value of 450 to 460 ppm, despite a more than three-fold concentration increase in the feed.

In earlier studies (10,14) on oil emulsions, the temperature dependence of the ultrafiltration flux has demonstrated Arrhenius

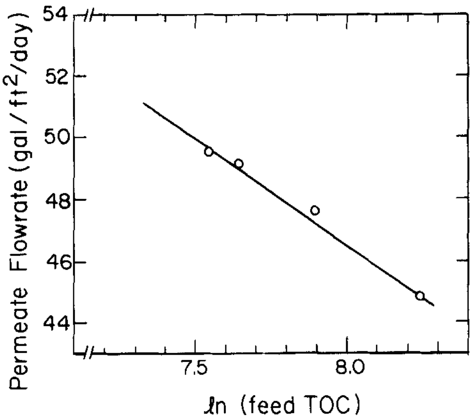

FIGURE 12. Ultrafiltration of yarn scouring water by a dynamically formed SiO_2 membrane: dependence of steady state permeate flowrate on TOC concentration of the yarn scouring water feed. Ultrafiltration conditions: Pressure - 17 psig; Q_R = 2 gpm for 1 ft^2 membrane module; T = 60°C.

behavior with an activation energy of 2 to 6 kcal/mole. Thus, the rapid increase in flux between 55 to 60°C must involve an additional phenomenon. One possible explanation is that at 50°C the oil emulsion is too near its phase separation point. It therefore becomes unstable near the membrane boundary due to concentration polarization and slowly begins to deposit some semi-solid paraffin on the membrane surface. (It will be recalled that gross phase separation occurs below ∿45°C). As the temperature is increased, the oil emulsion droplets not only back diffuse faster but also achieve stability, so that deposition layers are avoided, thereby enhancing the flowrate.

Since the permeate had a very low turbidity (Table 2), most of its TOC may be assumed to arise from organic surfactants used to aid the scouring process. For this reason, they would not be expected to interfere (and would likely aid) the re-use of permeate in the scouring process. Scouring tests in small kettles containing 15 L of permeate appear to confirm this expectation, at least for those yarns and dyes tested to date.

CONCLUSIONS

The pleated thin channel crossflow module has the ability to function on solutions contaminated with up to 800 ppm of particulates, which is a factor of 30 larger than that possible with the spiral wound design.

Due to the low pressure drop in the permeate compartment, the pleated crossflow module is especially suitable for microfiltration and can be operated in the constant permeate flow rate mode for this purpose.

Modules containing dynamically formed silica ultrafiltration membranes can be created on microporous supports in the pleated crossflow configuration. Such modules can be used for the ultrafiltrative recovery of hot water for re-use in yarn scouring processes, thereby contributing to a saving of both water and energy.

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REFERENCES

1. D. McBain, Chemistry in Britain 12, 231 (1976).
2. Private communication from P. Johnston, Gelman Sciences. Sedimentation analysis was performed by R. Hall, Micron Data Laboratories, on a Sedisperse E-1.
3. B. R. Breslau, E. A. Agranat, A. J. Testa, S. Messinger, and R. A. Cross, Chem. Eng. Prog. 71, 74 (1975).
4. A. G. Collins, "Enhanced Oil Recovery Injection Waters," in Oil Field Subsurface Injection of Water, ASTM STP 641, C. C. Wright, D. Cross, A. G. Ostroff, and J. R. Standord (eds.) (1977), pp. 2-23.
5. W. L. Griffith, G. B. Tanny, and A. L. Compere, Dev. Ind. Microbiology 20, 743 (1979).
6. H. P. Grace, A.I.Ch.E. 2, 307 (1956).
7. W. F. Blatt, A. Dravid, A. S. Michaels, and L. Nelson, Membrane Science and Technology, J. E. Flinn (ed.), Plenum Press, New York (1970), p. 47.
8. R. Rautenbach and K. Rauch, Int. Chem. Eng. 18, 417 (1978).
9. N. C. Beaton, Textile Institute and Ind., 361 (1975).
10. A. J. C. Pearson, C. A. Anderson, and G. F. Wood, J.W.P.C.F. 48, 945 (1976).
11. G. B. Tanny, Sep'n. and Pur. Methods 7, 283 (1978).
12. G. B. Tanny, in preparation.
13. D. Freilich and G. B. Tanny, Desaln. 27, 233 (1978).
14. R. L. Goldsmith, D. A. Roberts, and D. L. Bune, J.W.P.C.F. 46, 2183 (1974).