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USE OF FLUIDISED BEDS AS TURBULENCE PROMOTORS IN TUBULAR MEMBRANE SYSTEMS

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SUMMARY

In this paper results are presented of our investigations on the use of fluidised beds as turbulence promotors in tubular membrane systems. The fluidised bed will be compared with other types of turbulence promotors like static mixers and spiral wires. Attention will be given to mass transfer, to possible damage of the membrane surface and to the energy-efficiency of the fluidised bed. Mass transfer is favourably effected at superficial velocities as low as 1 cm/sec. Damage of the membrane surface could not be observed for glass particles equal to or smaller than 0,7 mm diameter. For a fluidised bed the same mass transfer coefficient as in the empty tube can be reached with only about 5% of the circulation energy.

INTRODUCTION

In hyper- and ultrafiltration processes solute concentrations at the membrane surface can be considerably higher than in the bulk of the feed solution. This phenomenon is called concentration polarization. It is caused by the fact that the rejected solute builds up a concentration gradient for back-diffusion into the bulk of the solution, which counteracts the convective transport of solute to the membrane. Adverse effects of the increased membrane wall concentra-

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tion are reduced quality and quantity of the product water and the possible occurrence of fouling and extra flux-decline. These adverse effects can be diminished by applying high superficial liquid velocities or turbulence promotors.

The first mathematical description of the concentration polarisation phenomenon in pressure driven membrane processes was given by Sherwood *et al.* [1] in 1963. Various aspects related to it were studied by o*"*ers later on [2-12]. Fig. 1a shows the concentration curve near the surface of an ultra- or hyperfiltration membrane in the case of a turbulent flow, while no precipitation occurs. The boundary layer is considered to be a thin film, which separates the membrane surface from the turbulent core (film model).
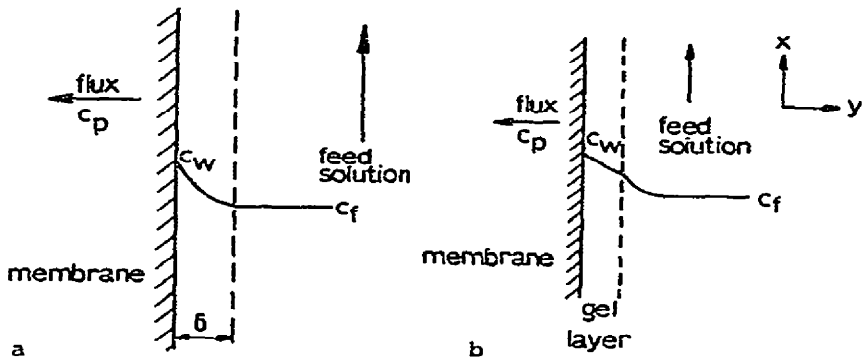

Fig. 1. Schematic view of concentration polarisation in the boundary layer adjacent to the membrane surface: a) in hyperfiltration processes; b) in ultrafiltration processes with gel layer build-up.

The concentration near the wall is higher than in the bulk of the liquid, resulting in decreased flux because of lower effective pressure $(P-\Delta\pi)$, decreased observed (overall) rejection, possible precipitation of fairly soluble solutes and possible shortened membrane-life. In addition, for ion-exchange membranes the intrinsic rejection itself decreases with increasing salt concentration [13, 14].

In ultrafiltration processes where larger molecules are processed and osmotic pressure is negligible, the situation is somewhat different from that in hyperfiltration processes. As a consequence of the low back diffusion velocity of the rejected molecules the solute concentration at the interface increases and a gel layer may be formed on the membrane surface (Fig. 1b). As in scaling, this gel

layer lowers the flux through the membrane considerably.
Since several applications of both ultra- and hyperfiltration pro-
cesses are greatly affected by concentration polarisation, several
authors studied the reduction of its deleterious effects by using
turbulence promotors. In this study we used fluidised bed parti-
cles, since this turbulence promotor might also be able to remove
mechanically the gel layer formed in ultrafiltration processes.

THEORY

Neglecting a concentration profile parallel to the membrane surface
one can describe the concentration build-up at the membrane surface
as a one-dimensional flow problem for which under steady-state con-
ditions the nett flux in any place is given by

$$J_w c = J_w \cdot c_p - D \cdot \frac{dc}{dy} \tag{1}$$

Integration of eq. (1) with boundary condition $c = c_f$ at $y = \delta$,
yields for $c = c_w$ at $y = 0$

$$\theta = \frac{c_w}{c_f} = \frac{\exp.[J_w \ \delta/D]}{R_i + (1-R_i)\exp.[J_w \ \delta/D]} \tag{2}$$

in which the intrinsic rejection is given by

$$R_i = (c_w - c_p)/c_w \tag{3}$$

During the experiments it is only possible to measure

$$R_{obs} = (c_f - c_p)/c_f \tag{4}$$

and R_{obs} will only approach R_i if c_w approaches c_f, at infinite
mass transfer. The film thickness δ is taken equal to that in the
case of mass transfer to an impermeable wall, the influence of the
water flux through the membrane being neglected as a first approxi-
mation, because of the small ratio (J_w/U) [31]. The film thickness
results from the mass transfer coefficient k:

$$k = \frac{D}{\delta} \tag{5}$$

Substitution of eqs. (4) and (5) in eq. (2) gives

$$\ln \left[\frac{1 - R_{obs}}{R_{obs}}\right] = J_w/k + \ln \left[\frac{1 - R_i}{R_i}\right] \tag{6}$$

Using the general relation for mass transfer

$$St = const. \; Re^{-m} . \; Sc^{-n} \tag{7}$$

with $Sc = \nu/D$ and $St = k/U$, k can be eliminated from eq. (6) to obtain

$$\ln \left[\frac{1 - R_{obs}}{R_{obs}}\right] = const. \frac{J_w}{U} . \; Re^{m} . \; Sc^{n} + \ln \left[\frac{1 - R_i}{R_i}\right] \tag{8}$$

R_i can now be determined by plotting $\ln \left[\frac{1 - R_{obs}}{R_{obs}}\right]$ versus $\frac{J_w}{U}. \; Re^{m}$ and extrapolating towards $U = \infty$. The concentration polarisation modulus θ can then be calculated by

$$\theta = \frac{1 - R_{obs}}{1 - R_i}$$

In this study we used fluidised beds as promotors. In a fluidised bed consisting of a liquid and solid particles the particles move at random. The drag forces on the solid particles are a function of the liquid velocity and the porosity of the bed. Gravity and drag forces are balanced, so the particles will not be hydraulically transported and the bed occupies a predictable volume depending on system-parameters.

The expected enhanced mass transfer is caused by the irregular flow of the liquid between the particles. The movement of the particles itself seems not to contribute to the mass transfer directly [33]. The impuls of the particles, however, may cause an erosive action, which removes a gel or fouling layer in ultra- or hyperfiltration processes. Possible erosion is mentioned by King and Smith [43].

Turbulence promotors

Methods to minimize the concentration polarisation modulus θ, (eq. 2), or to increase the mass transfer coefficient k, are the use of higher flow rates or of very narrow channels, stirrers etc. Several of the authors mentioned below conclude that the use of turbulence

promotors gives great cost advantage [10,15]. The most fruitful area for further work on this subject was suggested to be the strongly fouling situations [15].

The objective of a turbulence promotor is to enhance the convective flow and to induce turbulences, thereby increasing the mass transfer. Since the action of the turbulence promotors does not necessarily involve turbulent flow, the name convection promotors is also sometimes used [18]. Good turbulence promotors introduce no stagnant regions, they do not damage the membrane and they act continuously [14-23]. The use of moving polyurethane sponge balls is also a well-known method to combat fouling in ultrafiltration processes [24,25]. Hamer [26,27] used moving glass spheres with a diameter nearly as large as the inside diameter of the tubular membranes.

Better than these large spheres could be a dispersed bed of many small particles, that constantly bomb the membrane surface. By this mechanism a solid-fluid fluid bed removes deposited matter from the membrane surface and reduces scaling and fouling in addition to reducing concentration polarisation. In their investigation Csurny *et al*. [28] used fluid beds with lead shot, stainless steel filings, stainless steel particles, pellets cut from stainless steel wire, tungsten particles and spherical glass beads. The metal particles soon gave corrosion products in the test unit and none of the beds employed were successful when applied to dynamically formed membranes or cellulose acetate (C.A.) ones. The authors concluded that their survey did not lead to a convincing evaluation pro or con. Later on both Lai [10] and Lolachi [17] proved the feasibility of a fluid bed in improving the membrane performance. In spite of these studies several questions remained unanswered, especially questions concerning the possible damage of the membrane by the fluid bed, the optimum diameter of the spheres and the application of fluid beds in ultrafiltration processes, where previous studies are unknown. Hence we studied the effects of fluid beds as turbulent promotors on the membrane performance, in both hyperfiltration (reverse osmosis) and ultrafiltration processes.

EXPERIMENTAL

Fluid bed equipment

The test equipment (supplied by WAFILIN B.V., Hardenberg, The Netherlands) was easy to handle and very suitable for our experiments, which necessitated frequent replacement of membranes. The mass transfer experiments were carried out with commercial membranes made from cellulose acetate (C.A.) or polyacrylonitrile (PAN). Each test module contained seven 1.5 meter polyvinyl chloride tubes in parallel, each equipped with a tubular membrane, one of which usually being used and the others blocked. The membranes were cast on the inside of a non-woven support tube. During the experiments we used both PAN ultrafiltration membranes (18 mm i.d.) and C.A. hyperfiltration ones (12 mm i.d.), made after Manjikian [29], by using a 1:1 blend of Eastman Kodak E 383-40 and E 398-6S polymers and a curing temperature of 80°C. Two similar types of apparatus were used, one for the ultrafiltration experiments at about 4.5 atm

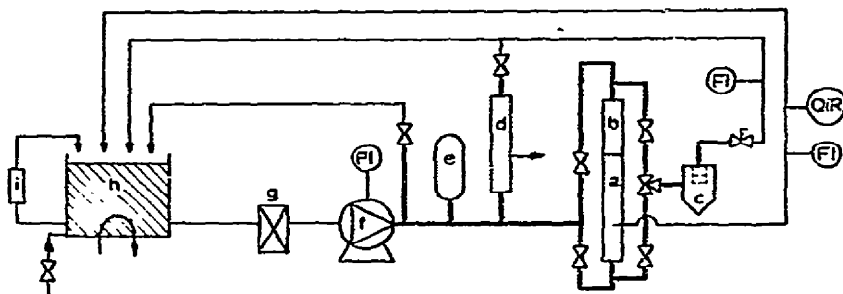and one for hyperfiltration experiments at 40 atm (Fig. 2).

Fig. 2. Hyperfiltration apparatus; a) membrane module; b) blind module without membranes; c) particles collector vessel; d) reference module; e) accumulator; f) pump; g) two filterite filters of 10μ; h) feed solution with cooling; i) ultraviolet sterilizer.

Both units contained two modules: one with membranes below (a, Fig. 2) and a blind one on top (b). At the bottom of the lower modules a perforated plate distributor was installed.

The hyperfiltration unit was equipped with an ultraviolet sterilizer (i) and two filterite 10 micron filters (g) to reduce the effects of bacterial growth and fouling during long term experiments. The feed could be recycled through a stainless steel particle collector (c), back to the supply vessel (h), which was thermostated by a cooling spiral. Parallel to the fluid bed equipment and connected to the same pump the membrane performance of a reference membrane without fluid bed could be measured (d) under the same feed conditions as the module with the fluid bed. The fluid bed particles used were Ballotini glass spheres of varying diameter. For the erosion-experiments a similar installation was used in which 20 parallel tubes of 1.5 m length followed by a blind transparent module of 0.2 m each was used. The glass particles were kept in place by appropriate filters on the bottom of the membrane tube and at the top of the blind module. The flow through each tube was controlled by a needle valve so that the upper side of the bed was visible in the blind tube.

RESULTS AND DICUSSION

Mass-tranfer experiments

Mass-transfer experiments have been carried out in the hyperfil-

tration unit using particle diameters of 0.4; 0.5; 1.0; 1.3 and
2.0 mm, fluidized at different velocities.
The behaviour of the fluidised bed can be described [30] by the em-
pirical correlation

$$U = U_i \, \epsilon^n \tag{9}$$

In Table 1 the experimental values of U_i and n for several parti-
cles are given. For particles larger than one tenth of the tube-
diameter, equation (9) does not describe the fluidised bed behav-
iour, because of non-homogenity of the bed at low U-values. For
higher velocities the bed becomes homogenous again, and the porosi-
ty then follows the given formula. For ϵ going to unity it follows
that U_i equals the free falling-velocity of a particle in the tube.

TABLE 1

Constants to be used in eq. (9) for tube-diameter of 12 mm

dp (mm)	U_i (cm/sec)	n	ρ (kg/m^3)
0.4	6.5	3.85	2890
0.5	7.7	3.50	2900
1.0	13.7	2.36	2670
1.3	15.0	2.55	2870
2.0	19.5	2.65	2500

At the start of an experiment the whole membrane tube was filled
with a packed bed. After the system was pressurized the axial ve-
locity gradually increased until the minimum fluidisation velocity
was reached. The system was kept under these conditions overnight,
whereupon the rejection and flux were measured at different axial
velocities. In Fig. 3 the dependence of rejection on the superfi-
cial velocity (U_o) is shown. Due to relatively small changes in
$(P - \Delta\pi)$ in these experiments at low salt concentrations (3000 ppm),
the flux of the fluid bed promoted membranes did not vary much with
U_o and it was generally the same as in the module without fluid bed
apart from the individual differences between the membranes.
The results in Fig. 3 show that application of a fluid bed enables
us to reduce the axial velocity to 5-20% of the value in the expe-
riments without turbulence promotion. The maximum found in Fig. 3
can be related to a maximum in mass transfer normally observed for
fluid beds. When the axial velocity through a packed bed increases,

the bed will expand; further increase of U_o will increase the bed height and bed porosity, while at very high U_o-values hydraulic transport of the particles will occur. Hence at high velocities, when all particles have left the tube, the R_{obs} vs U_o curve of the fluid bed promoted membrane will coincide with that of the unpromoted membrane.
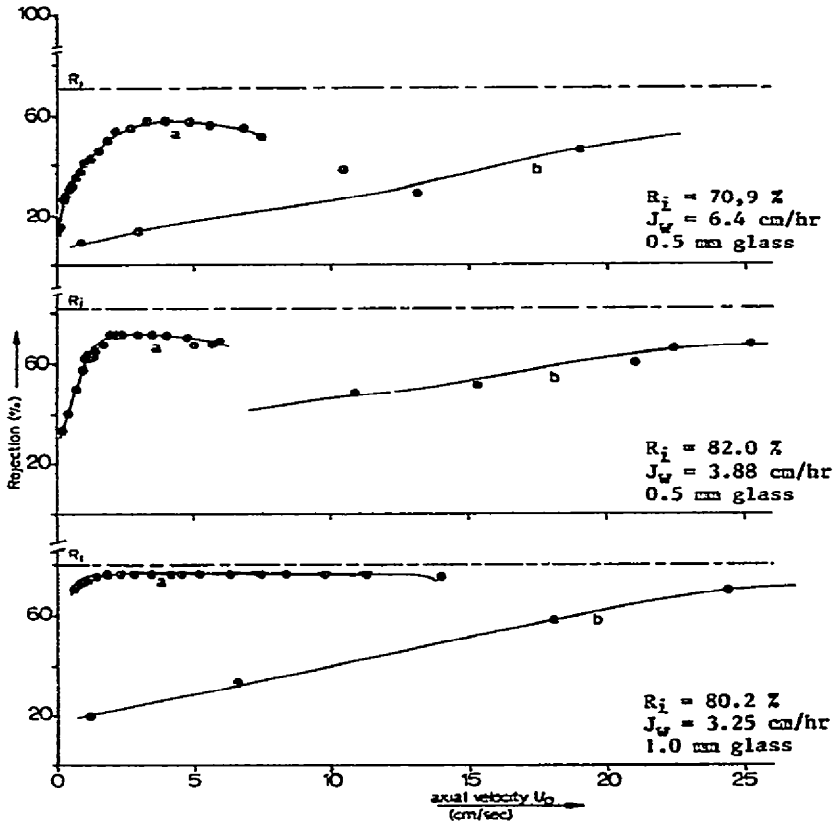

Fig. 3. Rejections observed with (curves a) and without (curves b) fluid bed as turbulence promotor.

The tubular membranes used here, are permeable to water, giving a radial flow which is uncommon in homogeneous fluid beds. On the assumption, however, that the influence of this radial flow on the hydrodynamics of the axial flow may be neglected, the following equations for mass transfer between a wall and the fluid bed are valid [31].

$$Re = \frac{d_H U}{v\epsilon} \; ; \; d_H = \frac{d}{1 + \frac{3}{2}(1-\epsilon)\frac{d}{dp}} \; ; \; St = \frac{k.\epsilon}{U} = \frac{k}{U_o} \; ; \; Sc = \frac{v}{D}$$

$$
\begin{aligned}
St &= 0.455 \; Re^{-0.44} Sc^{-0.70} \quad \text{for} \quad 1.98 \text{ mm} < d_p < 2.85 \text{ mm} \\
St &= 0.275 \; Re^{-0.38} Sc^{-0.70} \quad \text{for} \quad 0.53 \text{ mm} < d_p < 1.98 \text{ mm} \\
& \qquad\qquad\qquad\qquad\qquad \text{and} \quad 5.80 \quad < Sc < 2100 \\
& \qquad\qquad\qquad\qquad\qquad\qquad\;\; 0.40 \quad < \epsilon < 0.90
\end{aligned}
\tag{10}
$$

Eq. (10) gives the relation between mass transfer and liquid veloc-
ity under the assumption that the normal flow through the membrane
wall does not influence the mass transfer film thickness; see also
Bird, Stewart and Lightfoot [40].
With the help of eq. (6) the values of the Stanton versus Reynolds
numbers are obtained from the experimental data. In the calculations
a correction was made for salt-enrichment and decreased axial veloc-
ity. In Fig. 4 it is shown that there is a reasonable agreement be-
tween theory and experiments. Only at low velocities the points
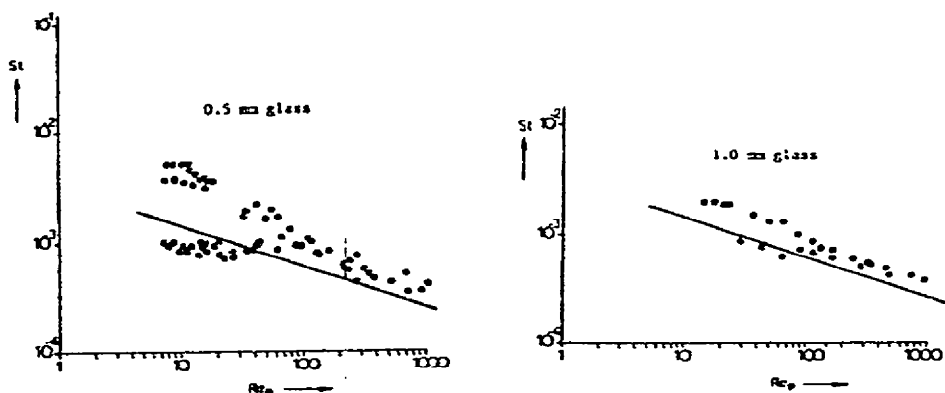tend to diverge, possibly due to the transmembrane flux.

Fig. 4. Mass-transfer in a fluid bed promoted tubular membrane module using C.A.
membranes (40 atm); ———: theoretical (eq. (10)); ●●●: experimental.

Although the influence of the transmembrane flux was not considered
before, the decrease in axial flow caused by the application of the
fluid bed must necessarily affect the ratio (J_w/U_o). Schlichting
[35] and Thomas [32] observed that the influece of the transmembrane
flux on the hydrodynamic behaviour of the boundary layer becomes
important above a (J_w/U) ratio of $1.18*10^{-4}$.

Our main conclusion is that there is an important increase in mass-transfer also at low velocities when applying a fluidised bed. This effect can be influenced by a proper choice of particle diameter and specific weight.

As we assumed that a fluid bed might have an important cleaning action on the membrane we did some ultrafiltration experiments in which formation of a gel-layer is expected.

Ultrafiltration experiments

In the ultrafiltration experiments tubular polyacrylonitrile membranes were applied. Polyethylene glycol was used with a molecular weight of 4,000,000 to make a 500 ppm feed solution, since it was expected that this solute would form a gel-layer on the membrane surface. During the experiments we used either three tubes pro module (2.0 mm particle diameter experiments) or all tubes (0.5 and 1.0 mm particle diameter experiments). With the exception of the experiment with 1.0 mm particle diameter, all experiments started without a fluid bed in the test module. At the end of each experiment the membranes were observed both by eye and with a scanning electron microscope. All membranes were covered with a visible gel-layer which could be removed by rinsing with pure water. Scanning Electron Microscope (SEM) photographs showed, however, that except when large fluid bed particle diameters were used, the gel-layer persisted on the membrane surface. This layer seriously hindered the study of membrane damage after simple cleaning. The persistence of a gel-layer after application of a fluid bed might be possible when the particles do not have enough momentum to remove the highly swollen, elastic polyethylene glycol gel in the boundary layer.

The SEM-photographs also revealed the presence of bacteria, which could not be eliminated completely.

Flux results of the experiments with 0.5; 1.0 and 2.0 mm particle diameters are shown in Fig. 5. The results are plotted on a double logarithmic scale as is usually done [38]. The dotted part of the curves represents the fluxes when the membrane did not contain a fluid bed. Because of the low pressures being applied (about 4.5 atm) membrane compaction is small and the observed flux decline with time can be ascribed to gel-layer build-up. Breakdown of the gel-layer results in a higher flux as shown for curves b) and c). The low flux and the absence of an increase in flux when a 0.5 mm particle fluid bed is applied, indicates that the momentum of the 0.5 mm particles (curve a) is insufficient te remove the gel-layer.
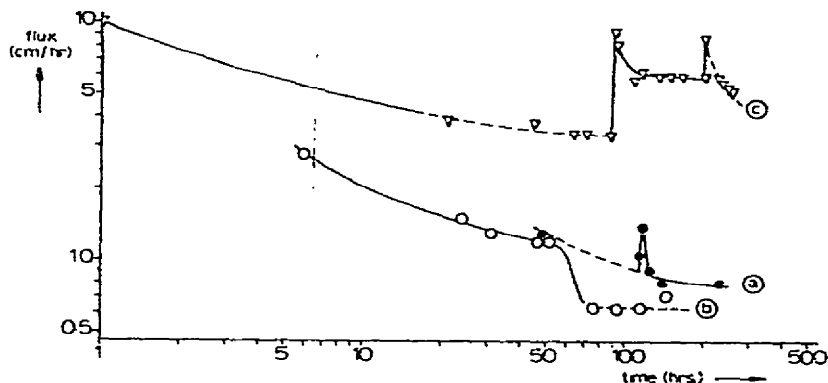
Fig. 5. Flux measurements on polyacrylonitrile membranes, with and without fluid
bed. a) 0.5 mm; b) 1.0 mm; c) 2.0 mm. ——— without fluid bed; ———
with fluid bed.

It can be concluded that the self-cleaning behaviour of filtration
systems containing a fluid bed turbulence promotor is insufficient-
ly effective for the smallest particles. The fluxes in all fluid
bed experiments with the larger particle diameters tend to be high-
er, but definite conclusions on the practical value of these results
cannot yet be drawn. Future experiments with various solutes are
necessary to study the influence of the fluid bed on rejection and
flux under practically important circumstances.

Damage of the membrane surface by fluid bed particles

The continuous bombardment of the fluid bed particles on the mem-
brane surface may result in damage of the thin homogeneous skin of
asymmetric membranes generally in use. Although both Lai [10] and
Lolachi [17] reported that no damage of the surface occurred, their
conclusions are based on just a few results from experiments of
short duration and hence are uncertain. Csurny [28], however, ob-
served that his irregularly shaped particles adhered to the dynamic
membrane surface tenaciously. Also membrane performance was worse
when applying a fluidised bed then an empty tube.
Because of the great importance for the actual life time of the mem-
branes, we studied this problem more extensively.
As mentioned before, the gel-layer build-up on polyacrylonitrile
membranes, when a solute of high molecular weight was used, made
the study of membrane damage difficult. As far as the membrane sur-
face was made visible, no damage by 0.5 and 1.0 mm diameter parti-
cles was observed. On one membrane, however, slide patterns of a
spherical indenter were found. Identical patterns on polymeric sur-
faces were reported for wear tests by Bethune [36] and Lawn and
Wilshaw [37]. The only way these patterns can be formed is by a
particle scraping over the surface. These patterns can be formed
when the bed is rapidly started up. The patterns found had indeed

the same direction as the length of the tube. The magnitude of the
deleterious effects of membrane damage by glass spheres depends to
a large extent on the skin thickness. For the polyacrylonitrile
membranes the skin is about two microns thick, while for the cel-
lulose acetate membranes it is about 2000 Å. Hence the occurrence
of damage is likely to be more easily observed in hyperfiltration
experiments with cellulose acetate membranes in which no gel-layer
build-up occurs.

Preliminary experiments showed no measurable damage for the 0.5 mm
particles, little damage for the 1.0 mm particles and considerable
damage for the larger ones (Fig. 6). It was observed in these expe-
riments that the flux remained almost constant, while the rejection
decreased. These experiments were carried out consecutively, so no
conclusive statements can be made. For better comparison, experi-
ments were carried out with a set-up consisting of 20 parallel ver-
tical tubes, all connected to one high-pressure feed line, but in-
dependently flow-controlled. Two consecutive experiments were done:
one in which drinking water of the city of Enschede was hyperfil-
tered without any treatment between the tap and the pump, apart
from a cartridge-filter (10μ), and a second experiment in which a
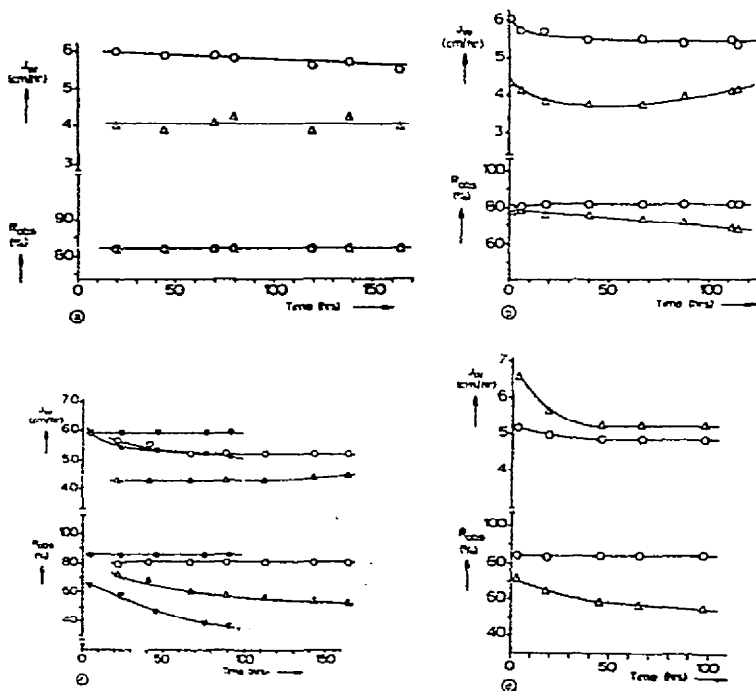3000 ppm NaCl-solution was recirculated through the system (closed
loop).

Fig. 6. Influence of a fluid bed on the performance of C.A. hyperfiltration mem-
branes compared with reference C.A. membranes tested without fluid bed.
a) fluid bed with 0.5 mm diameter; b) 1.0 mm; c) 1.3 mm; d) 2.0 mm.
(o,● unpromoted membranes; Δ,▲ fluid bed promoted membranes).

Assuming that at any time any membrane-spot has an equal chance to be damaged, the relation for the decrease in undamaged area A reads

$$- \frac{dA}{dt} = b \, A \tag{11}$$

from which follows that

$$A = A_o \exp(-bt) \tag{12}$$

Assuming that damaged surface gives no rejection at all, and that the flux through the damaged surface is the same as for the intact part, it follows that

$$R_t = R_o \exp(-bt) \tag{13}$$

Experiments indeed show this relationship to be valid. During the experiments the flux showed to be a slight function of time also, therefore this variable was plotted in the figures too. A typical example of the results is given in Fig. 7.
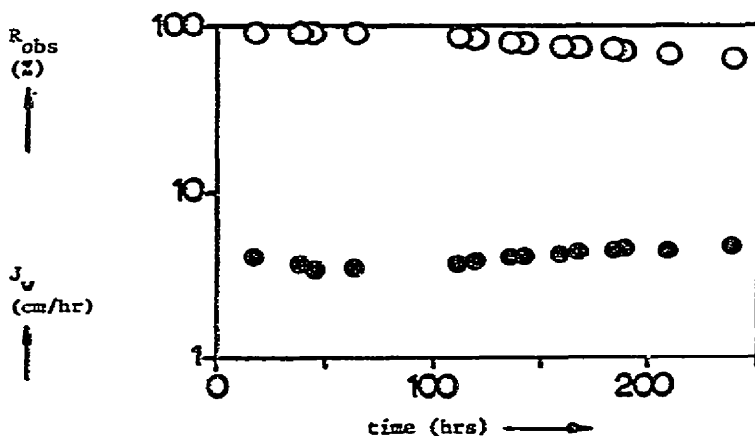

Fig. 7. Typical curves for damage-experiments; d_p ≈ 2.0 mm, ε ≈ 50%, closed loop experiment; o: rejection; •: flux.

The results are given in Table 2. From these rejection data there seems to be a trend of increasing damage with increasing particle

diameter, although the numbers for the particles equal to or smaller than 0.7 mm, do not show any damage at all.

TABLE 2

Deterioration numbers, b from eq. (13), for parallel damage-experiments

porosity	dp (mm)	$b*10^4$ (hr^{-1})	
		closed loop experiment	drinking water experiment
0.50	.4	1.08	-3.43
	.5	0.52	-0.65
	.7	-0.12	-0.59
	1.0	0.04	0.77
	1.3	2.78	-
	2.0	23.4	
0.60	.4	-0.25	-
	.5	0.77	3.90
	.7	0.58	0.97
	1.0	1.20	0.41
	1.3	10.8	9.00
	2.0	-	
0.80	.5	0.64	-3.80
	.7	1.30	-1.00
	1.0	5.23	8.10
	2.0	30.2	16.8
	3.0	4.62	
1.00	-	-0.03	-0.24
	-	-0.18	-0.12

Use of large fluid bed particles offers great advantage by an increased mass transfer and increased ability of the system for self-cleaning. These larger particles, however, damage the hyperfiltration membranes to an unacceptable level when dp ≳ 1 mm. Hence fluid beds with 0.7 mm particles should be considered as the upper limit, when C.A. membranes are used. This conclusion is not in conflict with the results of Lai [10] and Lolachi [17]. The results on deterioration found by Csurny et al. [28] with other kinds of particles cannot be compared here, partly due to the irregular particles used and partly by the different type of damage that occurs for a dynamically formed membrane by fluid bed particles. The membranes of ultrafiltration experiments in which either glass (1.3 and 2.0 mm), stainless steel (2.0 mm) or lead (3.0 mm) spheres were used, were also visually studied. On none of these membranes a gel-layer was observed, and damages were visible.

Energy efficiency

Turbulence promotors will be applied to mass-transfer systems, when e.g. recovery of mass or lowering of energy input can be improved. Many authors [18,41,44] working on turbulence promotors only mention results on rejection and flux as a function of flow-velocity, without taking care of the additional pressure-drop caused by the promotor. On the other hand there is the misunderstanding that the necessary extra pressure-drop causes serious energy-losses in the system [42].

When pressure-drop in the system is low compared to the absolute pressure in the system, the relation between pressure and velocity is not important, but the resulting decreased velocity will influence the system-design: for the same product-quantity and recovery the system will be shorter, resulting in a pressure-drop which is not necessarily higher than for the empty-tube system; for such a system membranes must be operated more in parallel.

When the pressure-drop in the system is of the same order of magnitude as the required absolute pressure, such as is the case in many ultra-filtration systems, a good comparison between the promoted and the unpromoted case is on the basis of required energy to obtain a certain mass-transfer. This energy per unit of time is expressed as the product of pressure-drop and quantity of displaced liquid:

$$E = U_o \, \pi \, d^2 \, \Delta p \tag{14}$$

So we propose to compare energy losses at equal mass transfer:

$$\frac{E_{promotor}}{E_{empty\ tube}} = \frac{U_o' \Delta p' (promotor)}{U_o \Delta p (empty\ tube)}$$

in which U_o' and $\Delta p'$ correspond to the same mass-tranfer coefficient as resulting from U_o in the empty tube of the same diameter. The advantage of such a comparison is a quick selection of promotors: if the ratio is larger than or equals 1, application of the promotor is a waste. If the ratio is smaller than 1, it depends on the ultimate design whether a promotor should be applied, since a parallel-system may require additional piping and insertion of the promotor results in extra costs of material and man-hours. As most authors do not report both mass-transfer and pressure-drop, the energy ratio can only be estimated for some promotors.

So for the Kenics Static Mixers, described by Pitera and Middleman
[18] a low ratio results for velocities between 1 and 10 cm/sec in
a tube of 13.5 mm diameter. For a better determination of the ratio
as a function of the velocity more data are necessary. Also the work
of Dejmek *et al.* [39], who included pressure-drop, did not allow a
determination of the energy ratio.

Thomas *et al.* [16] give combined graphs for mass-transfer and pres-
sure-drop for three different inserted or attached spiral wires. For the
continuous runner in a 6 mm tube the energy ratio is smaller than 1
for velocities smaller than 0.6 m/s in the promoted tube, corres-
ponding with an empty-tube velocity of 1.35 m/s. For higher veloc-
ities the ratio becomes as high as 10. For the partial runner the
ratio is smaller than 1 for 0.95 m/s in the promoted case (2.3 m/s
in the empty tube). The attached spiral wire can be advantageous at veloc-
ities smaller than 1.6 m/s in the promoted case (3.8 m/s in the
empty tube). For velocities higher than the ones given the ratio is
larger than 1, except for the attached spiral wire, where the ratio re-
mains about 1 ($\pm$ 10%) for higher velocities. For the fluidised beds
the mass-transfer is given as a function of the velocity, eq. (10).
The pressure-drop equals the extra weight per unit of surface of
the fluidised particles. From this the ratio can be calculated. The
results are given in Fig. 8 where the energy ratio and the empty-
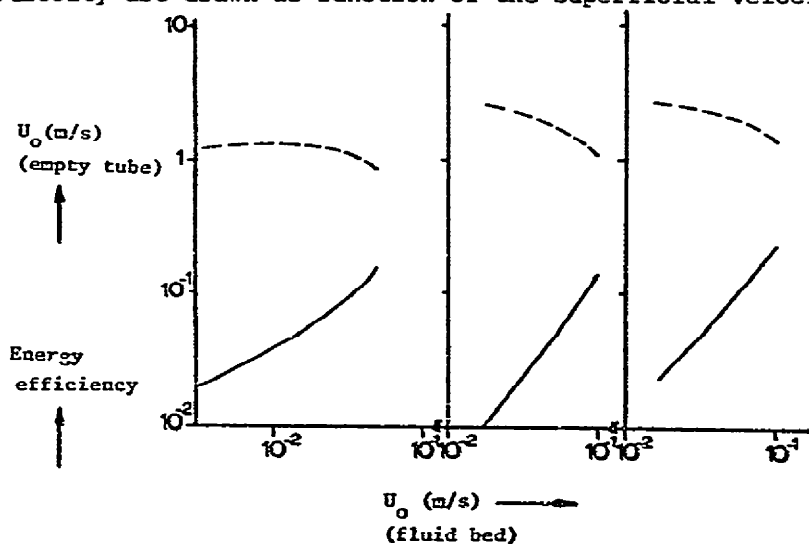tube velocity are drawn as function of the superficial velocity in

Fig. 8. Energy efficiency and corresponding empty-tube velocity as a function of
fluid-bed superficial velocity in a 14 mm tube. ——— efficiency;
——— empty tube velocity.

the fluidised bed. Over the whole range the energy ratio is in the
order of magnitude of 1-20%; for the maximum mass-transfer in the
bed it is smaller than 5%. Assuming that the mass-transfer is sat-
isfactory, most promotors may be advantageous (based on energy-con-
sumption) within a limited range of fluid velocities, and hence li-
mited to certain values of the mass-transfer coefficient.

CONCLUSIONS

The study described here shows that a fluid bed can be effective in
combatting concentration polarisation in tubular membrane modules.
When the bed is rapidly started up care must be taken to avoid
crack patterns of sliding (spherical) indenters. The fluid bed was found
to be useful both for hyperfiltration applications and ultrafiltra-
tion purposes. Accurate rejection data for the latter, however, must
become available. A proper choice of the particle size is essential
in order to prevent membrane damage by the fluid bed particles. Es-
pecially the asymmetric cellulose acetate membranes (with an ul-
trathin skin) can be deteriorated by particles larger than 0.7 mm
diameter.
No membrane damage of the polyacrylonitrile membranes was observed
with smaller beads; with larger particles of both glass and metals
membrane damage was observed. The fluid beds used were unable to
remove the gel-layer completely (0.5 and 1.0 mm glass) or to remove
the gel-layer without damaging the surface (1.3 and 2.0 mm glass).
Application of a fluid bed enables the use of low axial velocities.
For very low velocities, however, the behaviour of the mass-trans-
fer cannot be predicted accurately. Application of most turbulence-
promotors can be advantageous, based on energy consumption.

NOMENCLATURE

A	undamaged surface area	(m^2)
A_0	same at t = 0	(m^2)
b	constant	(t^{-1})
c	concentration	(mol/l)
c_f	concentration in feed	(mol/l)
c_p	concentration in product	(mol/l)
c_w	concentration at the interface membrane/feed	(mol/l)
d	inside diameter of the tubular membrane	(m)
d_H	hydraulic diameter	(m)
d_p	particle diameter	(m)
D	diffusion coefficient	(m^2/s)
E	energy per unit of time	$(m^3\ atm/s)$
J_w	water flux through the membrane	(m/s)

k	mass-transfer coefficient	(m/s)
m	constant	(-)
n	constant	(-)
P	applied pressure	(atm)
Δp	pressure drop	(atm)
Re	Reynolds number $(Ud_H/v\varepsilon)$	(-)
Ri	intrinsic rejection $(1-(c_p/c_w))$	(-)
R_0	rejection at t = 0	(-)
R_{obs}	observed rejection $(1-(c_p/c_f))$	(-)
R_t	rejection at t = t	(-)
Sc	Schmidt number (v/D)	(-)
St	Stanton number (k/U)	(-)
t	time	(s)
U	interstitial velocity	(m/s)
U_i	free falling velocity in tube	(m/s)
U_0	superficial velocity	(m/s)
y	distance from the membrane surface	(m)

Greek symbols

δ	film thickness	(m)
ε	voidage fraction	(-)
∂	concentration polarisation modulus = (c_w/c_f)	(-)
v	kinematic viscosity	(m^2/s)
$\Delta\pi$	osmotic pressure difference across the membrane	(atm)
ρ	specific gravity	(kg/m^3)

REFERENCES

1. T.K. SHERWOOD, P.L.T. BRIAN AND R.E. FISCHER, M.I.T. Desalination Res. Lab. Report 295-1 (1963).
2. W.N. GILL, L.J. DERZANSKY AND M.R. DOSHI in: Surface and Colloid Science 4, E. Matijević, Ed., Wiley Interscience, New York (1971).
3. W. PUSCH in: Reverse osmosis membrane research, H.K. Lonsdale and H.E. Podall, Eds., Plenum Press, New York (1972) pp. 299-317.
4. M.C. PORTER, I & EC Prod. Res. Develop. 11 (1972) 234.
5. H. STRATHMANN, Chemie-Ing.-Techn. 44 (1972) 1160.
6. H. STRATHMANN, Chemie-Ing.-Techn. 45 (1973) 825.
7. H.K. LONSDALE in: Industrial processing with membranes, R.E. Lacey and S. Loeb, Eds., Ch. VIII, p. 123, Wiley-Interscience, New York (1972).
8. S. SOURIRAJAN, Reverse osmosis, Logos Press, London (1970).
9. O. DEJMEK, Thesis, Lund Institute of Technology, Lund, Sweden (1975).
10. J. LAI, Thesis, Montana State University, Bozeman, Montana, U.S.A. (1971).
11. S. KIMURA AND S. NAKAO, Desalination 17(3) (1975) 267.
12. E. FRIEDRICH AND R. TIETZE, Chemische Technik 26(9) (1974) and 26(11) (1974) 688.
13. D.G. THOMAS, I & EC Fund. 11 (1972) 302.
14. D.G. THOMAS, I & EC Proc. Des. Delop. 7 (1968) 397.
15. D.G. THOMAS, W.L. GRIFFITH AND R.M. KELLER, Desalination 9 (1971) 33.
16. D.G. THOMAS, P.H. HAYES, W.R. MIXON, J.D. SHEPPARD, W.L. GRIFFITH AND R.M. KELLER, Environ. Sci. Technol. 4 (1970) 1129-1136.
17. H. LOLACHI, O.S.W. R & D report nr. 843, U.S. Dept. of the Interior (1973).

18. E.W. PITERA AND S. MIDDLEMAN, I & EC Proc. Des. Develop. 12 (1973) 52.
19. T.J. KENNEDY, L.E. MONGE, B.J. McCOY AND R.L. MERSON, AIChE Symp. no. 132, 69 (1973) 81.
20. T.J. KENNEDY, R.L. MERSON AND B.J. McCOY, Chem. Eng. Sci. 29 (1974) 1927.
21. W.L. THAYER, L. PAGEAU AND S. SOURIRAJAN, Canadian J. Chem. Eng. 53 (1975) 422.
22. R.A. SHAW, R. DELUCA AND W.N. GILL, Desalination 11 (1972) 189.
23. Rev-O-pack, Commercial membranes, Van Aartsen, Haarlem, The Netherlands.
24. WAFILIN B.V., Commercial membranes, Hardenberg, The Netherlands.
25. J.D. SHEPPARD AND D.G. THOMAS, AIChE 17 (1971) 910.
26. E.A.G. HAMER, U.S. Patent 3, 425, 562 - 4 Febr. (1969).
27. E.A.G. HAMER AND R.L. KALISH, Paper presented at the 2nd O.S.W. Symp. on Reverse Osmosis, Miami, Florida (1969).
28. J. CSURNY, J.S. JOHNSON Jr, K.A. KRAUS, H.O. PHILIPS, W.G. SISSON AND C.G. WESTMORELAND in: Biennial Progress Report for the period 15/3-1968-15/3-1970, G.E. Moore and J.S. Johnson Jr. Eds. p. 270, Oak Ridge National Laboratory, Oak Ridge, Tennessee (1973).
29. S. MANJIKIAN, I & EC Prod. Res. Develop. 6, (1967) 23.
30. J.F. RICHARDSON AND W.N. ZAKI, Trans. Inst. Chem. Eng. 32, (1954) 35.
31. T. MIZUSHINA, S. TAKESHITA, J. YOSHIZAWA AND I. NAKAMAE, J. Chem. Eng. Japan 5 (1972) 361.
32. D.G. THOMAS, I & EC Fund. 12 (1973) 396.
33. J.W. SMITH AND D.H. KING, Can. J. Chem. Eng. 53 (1975) 41.
34. F. COEURET, P. LE GOFF AND F. Vergens, Proc. Int. Symp. on Fluidization, Eindhoven p. 537-552 (1967).
35. H. SCHLICHTING, Boundary layer theory, McGraw-Hill Book Co., New York (1968).
36. B. BETHUNE, J. Mater. Sci. 11 (1976) 199.
37. B. LAWN AND R. Wilshaw, J. Mater. Sci. 10 (1975) 1049.
38. J.D. SHEPPARD AND D.G. THOMAS, Desalination 8 (1970) 1.
39. P. DEJMEK, B. FUNFTEG, B. HALLSTRÖM AND L. WINGE, J. Food. Sci. 39 (1974) 1014.
40. K.B. BIRD, W.E. STEWART AND E.N. LIGHTFOOD, Transport Phenomena, Wiley (N.Y. (1960).
41. C. PERI AND W.L. DUNCKLEY, J. Food Sci. 36 (1971) 395.
42. F. FORBES, The Chem. Eng. (Jan. 1972) 29.
43. D.H. KING AND J.W. SMITH, Can J. Chem. Eng. 45 (1967) 329.
44. A.L. COPAS AND S. MIDDLEMAN, Ind. Eng. Chem. Proc. Res. Dev. 13(2) (1974) 143.