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The transversal flow microfiltration module. Theory, design, realization and experiments*

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Abstract

This study shows that the permeate flux in microfiltration can be increased without increasing energy consumption, by using the hollow fiber membranes themselves as turbulence promoters This is realized by using shell-side fed hollow fibers, which are placed perpendicular to the feed stream This module type is called the transversal flow module The theoretical benefits of such a module have been calculated by applying the model proposed by Zydney and Colton to predict the permeate flux Furthermore the technical realization of a module with a filtering area of 0 5 m^2 as well as the first preliminary microfiltration results will be discussed

Keywords microfiltration, module design, concentration polarization, turbulence promotion, transversal flow

Introduction

In membrane filtration in general and in crossflow microfiltration especially a severe flux decline in the permeate flow can appear When operating an installation with a very fouling feed stream, containing e g yeast, the flux can drop to 1% of the clean water flux For other kinds of pressure driven membrane processes the flux can vary from 50% for reverse osmosis to 10% for ultrafiltration So especially for microfiltration it can be useful to control and to decrease this flux decline

Flux decline can be caused by several factors, e g concentration-polarization, adsorption, gel-layer formation and plugging of the pores [1] All these factors can be considered as a resistance against transport of a permeate from a feed solution across a membrane (Fig 1)

The plugging resistance (R_p) depends on the particle dimensions and on the pore dimen-

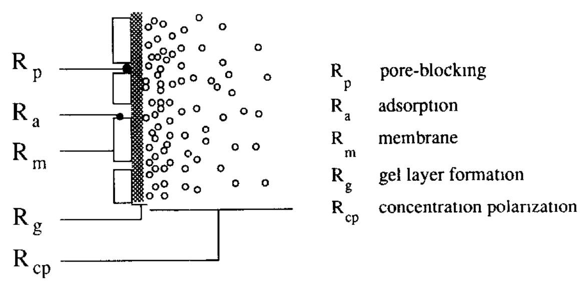

Fig 1 Sketch of a membrane with various resistances

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sions The adsorption resistance (R_a) depends on the properties of the particle material and the membrane material. The membrane resistance (R_m) is a function of the pore radius, the porosity and the turtuosity The gellayer resistance (R_g) and the concentration-polarization resistance (R_{cp}) depend on the composition of the feed stream, the process conditions and the module performance

This work deals with reducing the resistance due to concentration polarization (R_{cp}) by changing the module geometry, under the assumption that no gel-layer is formed ($R_g=0$)

Theory

Various models are used to predict the permeate flux in crossflow microfiltration Examples are: the models of Schock [2], of Blake [3] and of Schultz and Ripperger [4] The models of Schock and Blake describe the permeate flux by using the force balance on a single particle in the feed stream. The model of Schultz and Ripperger uses the pressure drop across the membrane to calculate the permeate flux These models do not predict the permeate flux in relation to concentration-polarization.

The model proposed by Zydney and Colton [5] describes the permeate flux by using the concentration-polarization model. It extends the existing models with describing the influence of particle diffusion near the membrane wall At steady state, the boundary layer is in a dynamic equilibrium If it is approximated by a stagnant film, the rate of convection of particles normal to the membrane is balance by diffusion back into the bulk

$$J(x)C=D\frac{\mathrm{d}C}{\mathrm{d}y} \tag{1}$$

where $J(x)$ is the local permeate flux in the flow direction, C the particle concentration, D the particle diffusion coefficient, and y the distance to the membrane This expression is integrated over the boundary layer thickness $\delta(x)$ to give:

$$J(x)=\frac{D}{\delta(x)}\ln\frac{C_w}{C_b}=k(x)\ln\left(\frac{C_w}{C_b}\right) \tag{2}$$

where $k(x)$ is the local mass transfer coefficient, C_w the particle concentration near the membrane wall and C_b the particle concentration in the bulk. According to Zydney and Colton for microfiltration the local mass transfer coefficient can be described as

$$k(x)=0\ 538\left(\frac{D^2\bar{\gamma}}{x}\right)^{1/3} \tag{3}$$

where x is the axial coordinate along the channel and $\bar{\gamma}$ is the average wall shear rate at the surface of the channel. Integration of the local mass transfer coefficient along the length of the channel yields an expression for the length-averaged flux

$$J=\frac{1}{L}\int_0^L J(x)\mathrm{d}x=0.807\left(\frac{D^2\bar{\gamma}}{L}\right)^{1/3}\ln\left(\frac{C_w}{C_b}\right) \tag{4}$$

The effective diffusivity, arising from particle migrations and rotations, governs particle transport in crossflow microfiltration This diffusivity increases with the square of the particle radius (a) and linearly with the shear rate Above a particle concentration of 20% it is constant with a value of

$$D=0\ 03\,a^2\bar{\gamma} \tag{5}$$

If eqn (5) is incorporated in eqn (4) the length-averaged flux is given as

$$J=0\ 078\left(\frac{a^4}{L}\right)^{1/3}\bar{\gamma}\ln\left(\frac{C_w}{C_b}\right) \tag{6}$$

So the permeate flux J is a function of:
- particle radius a
- particle concentration near the wall C_w
- particle concentration in the bulk C_b

– channel length L.
– average shear rate $\bar{\gamma}$.

The theory of Zydney and Colton has some limitations:

– It can only predict the maximum pressure-independent permeate flux
– It does not predict the flux decline in time
– It assumes the particle concentration in the feed stream to be constant along the module This only applies for a high feed flow.

However, according to eqn (6) the permeate flow can be enhanced by·

– Changing the feed stream: enlarging the particle radius or decreasing the particle concentration.
– Enlarging the shear rate.
– Reducing the channel length

Some remarks on these topics·

• Changing the feed stream is rather impractical in most applications. Only in very special applications particle radius can be increased by flocculation [5].
• The shear rate can be enlarged by increasing the pressure drop over the module.

$$\bar{\gamma}=\frac{\Delta P_{\mathrm{mod}}A_{\mathrm{d}}}{\eta A_{\mathrm{w}}} \tag{7}$$

The shear rate is a function of the pressure drop over the module (ΔP_{mod}), the area of the cross-section (A_{d}), the wetted area of the module (A_{w}) and of the feed stream viscosity (η) The pressure drop depends on the feed flow velocity and on the module design. A higher pressure drop over the module can be realized by enlarging the feed flow (Φ) (by using recirculation) Increasing both the feed flow and pressure drop implies the energy consumption (W) to increase linearly to the pressure drop *and* to the feed flow.

$$W=\Delta P_{\mathrm{mod}}\Phi \tag{8}$$

A better way to increase the permeate flux is to increase the pressure drop over the module while maintaining a low feed flow velocity, i e by changing the module design.

• A short channel length means decreasing the length along which a boundary layer is build-up This can be done in two ways reducing the module length or changing the module design An example: most membrane manufacturers build longitudinal modules with a length of approximately 1 0 m By using two modules of 0.5 m instead of one module of 1 0 m the permeate flow should increase 26%, according to the theory of Zydney and Colton.

$$J_{0\,5}=\left(\frac{1}{0.5}\right)^{1/3} J_1=1\ 26J_1 \tag{9}$$

In this work we have modified the module design Hollow fiber membrane modules can be of a tube-side fed type or a shell-side fed type The shell-side fed module can be operated in longitudinal flow (feed flow along the fiber) or in transversal flow (feed flow perpendicular to the fiber) A schematic drawing of a transversal flow module is presented in Fig 2.

The basic principle of a transversal flow module is well known and was investigated by· Yang and Cussler [6], Akzo international research [7], Rhône-Poulenc [8], Standard Oil [9]

The transversal design has basically the same benefits as the corrugated-plate membrane modules, which were realized by Van der Waal et al. [10,11]. In this modules, turbulence is increased by the corrugations, resulting in high values of mass transfer, especially in the region

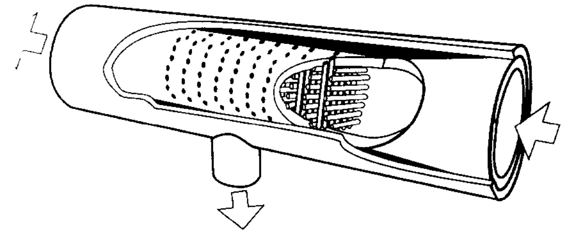

Fig 2 Sketch of a transversal flow microfiltration module

of reattachment of the streamlines (Fig. 3).

The expected theoretical benefits of the transversal flow module can be calculated according to eqn (6) The permeate flux J is a function of known parameters
- The particle radius
- The concentration
- The channel length

and of unknown parameters
- The particle concentration near the membrane wall
- The average shear rate

Zydney and Colton estimated the concentration C_w for rigid particles to be 74%, assuming a hexagonal close-packing The average shear rate depends on the pressure drop across the module and on the module dimensions (eqn 7) For longitudinal flow modules the pressure drop across the module can be calculated using

$$\Delta P_{\mathrm{mod}} = \xi \frac{L}{d_i} \frac{\rho v^2}{2} \tag{10}$$

The friction factor ξ is a function of the Reynolds number [12] Brauer [13] derived an equation to predict the pressure drop over transversal flow heat exchangers

$$\Delta P_{\mathrm{mod}} = \frac{N}{2} \xi_t \frac{\pi^3 Re_h^2}{16} \frac{\rho v^2}{d^2} \frac{1}{\left(\frac{s_t}{d} - \frac{\pi}{4}\right)^3} \tag{11}$$

where N is the number of grids, i.e. the number of hollow fibers placed after each other in the module, Re_h is a function of the mean velocity and of the hydraulic diameter d_h, the friction factor is a function of the Reynolds number and of the relation s_t/d, ν is the kinematic viscosity of the feed stream

The module type to compare the transversal flow module with, is a tube-side fed longitudinal flow module This module is produced by X-Flow [14] Equation (6) is solved for a range of Reynolds numbers for these two module types The solution used for the calculations is a suspension of 0 5% (by volume) rigid particles. The particle diameter is 0 5 μm

X-Flow R05-PVC

Membrane area·	A_{mem}	$= 0\ 5\ \mathrm{m}^2$
Module length	L	$= 0\ 5\ \mathrm{m}$
Internal fiber diameter	d_i	$= 1\ 55\ \mathrm{mm}$
Number of fibers	250	

Transversal flow module, yet to produce

Membrane area	A_{mem}	$= 0\ 474\ \mathrm{m}^2$
Module length	L	$= 0\ 21\ \mathrm{m}$
Outer fiber diameter	d	$= 2\ 1\ \mathrm{mm}$
Internal module diameter:		62 mm
Number of grids	N	$= 100$
Transversal pitch	s_t	$= 4\ 2\ \mathrm{mm}$

Both modules have the same microfiltration fibers poly-ether sulfon with a pore diameter of 0 2 μm produced by X-Flow

According to the theory of Zydney and Colton the modules have the same performance for low Reynolds numbers (Fig 4), and with increasing Reynolds number the permeate flux of a transversal flow module increases much more than the flux of a longitudinal flow module

Although the transversal flow module shows

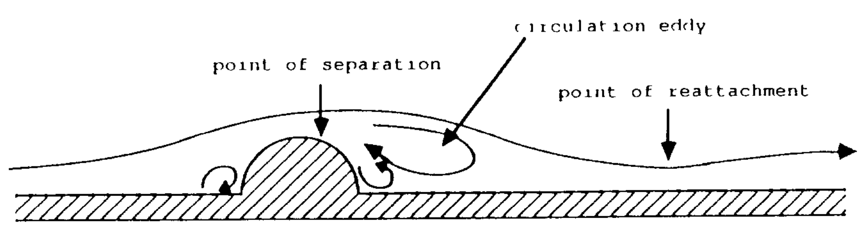

Fig 3 Impression of streamlines for flow over corrugated plates

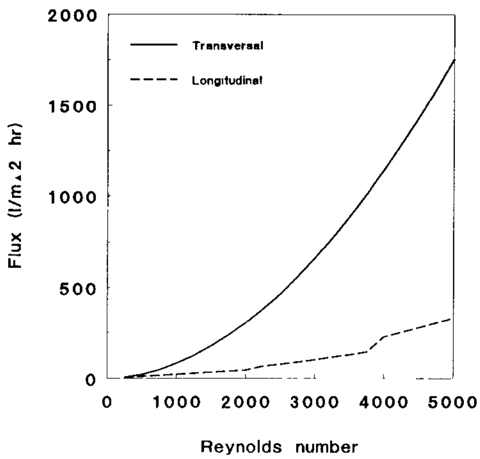

Fig 4 Calculated flux as a function of Reynolds number for transversal flow and longitudinal flow modules

a better performance than the longitudinal flow module, this is not convincing to potential industrial customers. These data have to be translated to costs In general there are two groups of costs
- the investment costs
- the operating costs

Investment costs are linear to the number of square meters of membrane area required to produce one cubic meter of permeate per hour.

$$A_{sp}=\frac{1}{J} \tag{12}$$

The operating costs are represented by the specific power consumption, i.e. the power consumption per cubic meter of permeate

$$W_{sp}=\frac{\Phi \Delta P_{mod}}{A_{mem}J} \tag{13}$$

Both factors, specific membrane area as well as specific power consumption are functions of the Reynolds number (Fig 5). The specific membrane area decreases and the specific power consumption increases with increasing Reynolds number This implicates that increasing the membrane area (so investment costs) will decrease the power consumption (so the oper-

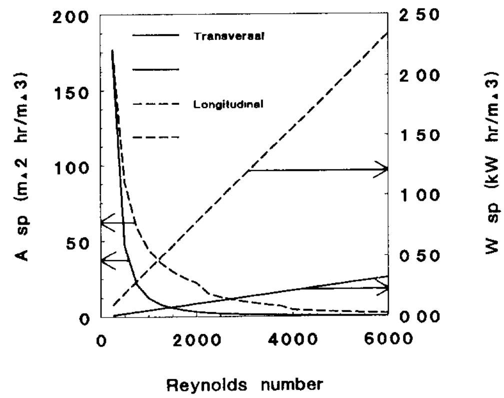

Fig 5 Calculations of specific membrane area and specific power consumption as a function of Reynolds number for transversal flow and longitudinal flow modules

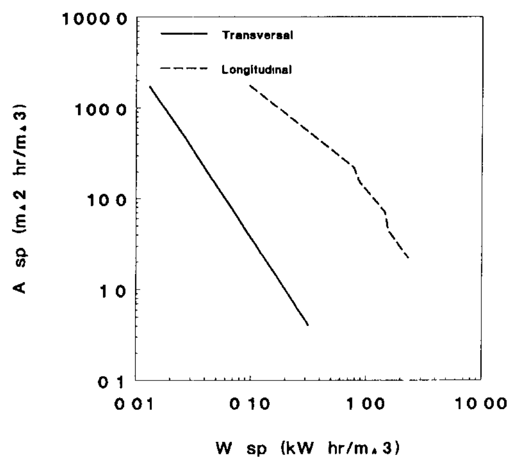

Fig 6 Calculations of specific power consumption versus specific membrane area for transversal flow and longitudinal flow modules

ating costs). Combining the two lines results in a graph displaying the energy consumption versus the specific membrane area (Fig. 6). An operating point at the bottom left means less total costs for a given membrane area. Moving to the right and upwards means increasing total costs Because of both axis being logarithmic this graph can only be used in a qualitative way, that is to compare the modules At this point we can conclude it could be profitable to use transversal flow modules instead of longitudinal flow modules

Fig 7 Photograph of series of realized transversal flow modules

Transversal modules

A photograph showing a chronological order of module series realized is shown in Fig. 7 The production technique is a centrifugal casting technique similar to the technique described by Nichols [9]

Experiments

The experimental set-up was a closed loop circulation system, controlling the operating pressure, the feed flow velocity and the temperature (Fig 8) The maximum feed flow is 8 m^3/hr and the maximum operating pressure is 4.5×10^5 Pa A personal computer is used to record the measured operating pressure, the pressure drop over the module, the feed flow velocity and the permeate flux

The experiments have been carried out with polystyrene globules This is a monodisperse material which has no interactions with the membrane material The particle diameter is 0 5 μm and the bulk concentration is 0.5% (by volume) The results are given in Table 1. These values are the average of 5 experiments each and are recalculated to the specific power con-

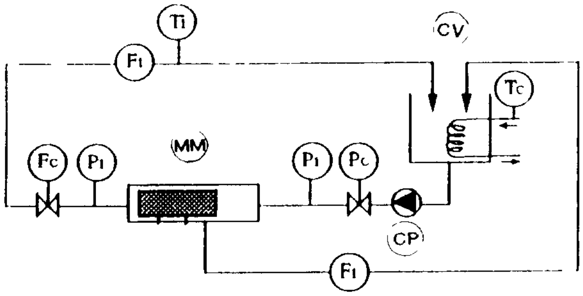

Fig 8 Flowdiagram of experimental set-up CV circulation vessel, CP circulation pump, MM membrane module, Pc pressure control valve, Pi pressure indicator, Fc flow control valve, Fi flow indicator, Ti temperature indicator, Tc temperature control

TABLE 1

Comparison of longitudinal flow and transversal flow modules in microfiltration of 0 5 vol % later globules

Re	*Φ* (m³/hr)	ΔP_{mod} (×10⁵ Pa)	Permeate flux (l/m²-hr)	A_{sp} (m²-hr /m³)	W_{sp} (kW-hr /m³)
Longitudinal flow module					
2400	4 0	0 67	211	4 74	0 706
3100	5 0	0 83	216	4 63	1 067
3700	6 0	1 35	204	4 90	2 206
Transversal flow module					
1500	3 0	0 07	295	3 39	0 042
2200	4 5	0 15	440	2 27	0 090
2400	5 0	0 17	430	2 33	0 116

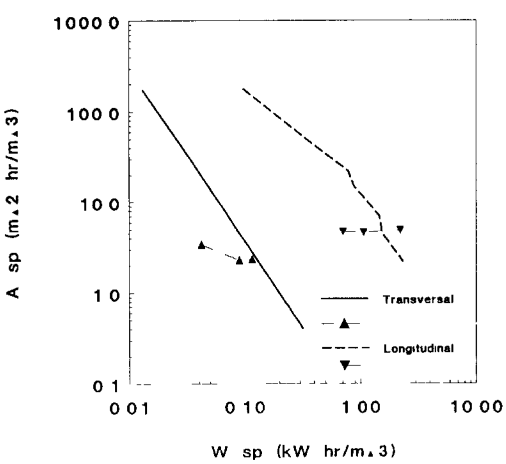

Fig 9 Comparison of experimental values on calculations

sumption and to the required membrane area and plotted in Fig 9

Discussion

The recorded data of the transversal flow module fit nicely to the predictions The longitudinal flow module is better than predicted. The reason for this effect can be that the boundary layer which is build up from the entrance of the module is limited to a maximum. This maximum is related to the internal radius of the hollow fiber Beyond the point at which this maximum is reached the rest of the module has a boundary layer with a constant height. The theory of Zydney and Colton predicts the boundary layer to increase all along the module to a very high value at the rear end of the module

Because of the effect of a limited boundary layer height the cutting in two pieces of a longitudinal flow module will result in a smaller increase in permeate flux than predicted by the theory

The experiments have been reconfirmed with experiments carried out with a 2% (by weight) yeast concentration (particle diameter approx 6 μm)

Reynolds number	Permeate flux Longitudinal flow	Permeate flux Transversal flow
4600	27 l/m²-hr	51 l/m²-hr

Probably because of a gel-layer formation and because of particle adsorption it is not possible to fit these values to the theory

Further research will be extended to other testing materials, like pigments and emulgated oil, commonly used in the industry Another point of research will be optimizing the process conditions. The effects of temperature and of backflushing will be investigated

Conclusions

(1) It is possible to produce transversal flow microfiltration modules in series by using a centrifugal casting technique developed in our laboratory.

(2) There is an overall fair agreement between the experimental data and the permeate flux predictions of the model proposed by Zydney and Colton

(3) Transversal flow microfiltration modules require less membrane area and/or less energy consumption than longitudinal flow modules to produce a given amount of permeate

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List of symbols

a	particle radius (m)
A_d	area of the cross-section (m^2)
A_{mem}	membrane area (m^2)
A_{sp}	specific membrane area (m^2-sec/m^3)
A_w	wetted area of a module (m^2)
C	particle concentration (-)
C_b	particle concentration in the bulk (%)
C_w	particle concentration near the wall (%)
D	particle diffusion coefficient (m^2/sec)
d	outer fiber diameter (m)
d_h	hydraulic diameter (m)
d_i	inner fiber diameter (m)
J	permeate flux (m^3/m^2-sec)
$J(x)$	local permeate flux (m/sec)
$k(x)$	local mass transfer coefficient (m/sec)
L	channel length (m)
N	number of grids in a transversal module (-)
Re_h	hydraulic Reynolds number (-)
s_t	transversal pitch (m)
W	power consumption (W)
W_{sp}	specific power consumption (W-sec/m^3)
x	axial coordinate (m)
y	distance to the membrane (m)

Greek

$\bar{\gamma}$	average shear rate (1/sec)
$\delta(x)$	local boundary layer thickness (m)
ΔP_{mod}	pressure drop over the module (Pa)
η	dynamic viscosity (Pa-sec)
ν	kinematic viscosity (m^2/sec)
ξ	friction factor (-)
ξ_t	transversal friction factor (-)
ρ	gravity (kg/m^3)
Φ	feed flow (m^3/sec)

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