DEXTRAN SEPARATION FROM SYNTHETIC RAW SUGAR SOLUTIONS BY ULTRAFILTRATION, AND ANALYSIS OF MEMBRANE FOULING MECHANISMS

Arwut Promraksa[1], Chalongsri Flood[2*], Philip Schneider[3], and Adrian Flood[4]

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Abstract

Dextran separation from synthetic raw sugar solutions was studied using regenerated cellulose and polyethersulfone membranes in stirred and unstirred batch ultrafiltration cells. It was found that the percent rejection of the 5000 molecular weight cut-off regenerated cellulose membrane is larger than that of other membranes investigated, and fouling of the regenerated cellulose membranes occurs more slowly than that of the polyethersulfone membranes. The percent rejection in the membrane separation can be improved by high speed agitation, which is likely due to lower levels of concentration polarization. The permeate flux decay for various transmembrane pressures in both unstirred and stirred filtration was used to predict the mechanism of the membrane fouling. A dextran cake layer is formed in unstirred dead-end filtration, with the cake compressibility in the range of 0.52-0.54, while the complete pore blocking model can be used to describe the fouling mechanism in stirred ultrafiltration. Membrane resistance determination was used for the initial flux data with the result that the resistances of the polyethersulfone membrane and the regenerated cellulose membrane are equal to 1.80×10^{13} m^{-1} and 3.36×10^{13} m^{-1} respectively.

Keywords: Dextran, membrane separation, stirred cell ultrafiltration, cake compressibility, fouling

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[1] *School of Engineering and Resources, Walailak University, Nakhon Si Thammarat 80161, Thailand.*
[2] *School of Chemical Engineering, Suranaree University of Technology, Nakhon Ratchasima 30000, Thailand. E-mail: chalong@sut.ac.th*
[3] *School of Engineering, James Cook University, Townsville 4811, Australia.*
[4] *School of Chemical and Biomolecular Engineering, Vidyasirimedhi Institute of Science and Technology, Rayong 21210, Thailand.*
[*] *Corresponding author*

Introduction

Background

Raw sugar is an important product for both export and the domestic production of refined sugar. Raw sugar from sugar cane is typically produced in areas with warmer climates (such as Brazil, Australia, South Africa, and Thailand, among others), while cold climate countries, particularly in Europe and North America, use sugar beet for raw sugar production. The two sources for sugar require differences in processing, particularly due to differences between cane and beet in their respective material properties, impurity types, and contents.

The process streams for producing raw sugar from cane initially involve a process of shredding and crushing the cane in a series of mills, with the cane being washed with hot water at 80°C to assist in the extraction of the sucrose-rich juice. The liquid product from the mills contains a number of organic and inorganic impurities and suspended particles. One of the most significant impurities that may be present is dextran. The presence of dextran in the raw sugar processing is due to microbiological activity during cane growth, harvesting, and storage (Cuddihy *et al*., 2001). Dextran is commonly produced from the sucrose present in cane by the *Leuconostoc* species of bacteria, which are common in soil. It is a homopolysaccharide of glucose characterized by a high fraction of α(1-6) linkages. The structure of dextran consists of a basic straight-chain polymer of α(1-6) linked glucose units, with some branches linked by α(1-2), α(1-3), or α(1-4) glycosidic bonds (Cuddihy *et al*., 2001; Rauh *et al*., 2001). The molecular weight of dextran is typically in the range of a few thousand to several million Daltons. The molecular weight and degree of branching in the dextran that is produced has been discussed for a particular *Leuconostoc* species (Kim *et al*., 2003). It has been known for many years that dextran adversely affects the process of sugar manufacture (and particularly refined sugar manufacture) in several ways such as sucrose loss, reduced accuracy of pol analysis, reduction of the evaporation rate, and the crystallization rate of sucrose (Eggleston and Harper, 2006), and deterioration of the crystal shape (Faria *et al*., 2003). Dextran in the sugar milling process is usually measured using units of ppm/Brix which indicates *x* ppm of dextran per mass of total solids (largely sugar).

Ultrafiltration is an alternative separation method to remove dextran from the raw sugar production. It is probably most economical to use it to remove dextran from the clarified juice, since this stream has a relatively low content of dissolved solids (approximately 15% Brix as compared to around 65% Brix after the evaporators) and thus should be able to operate at a relatively high flux due to a lower viscosity. The clarified juice stream has already had most of the other impurities removed through the clarification process, and this should reduce the problem of membrane fouling in comparison to the juice from the mills.

The major limiting factors of membrane filtration, such as permeate flux decline, percent rejection, and membrane fouling are commonly dependent on the molecular weight cutoff (MWCO) in relation to the size of the solute molecule (and other molecules present in the liquid) and type of membrane. In the ultrafiltration process the MWCO of the membrane is usually characterized by a term of 90% rejection of globular proteins by the membrane (Porter, 1990). The rejection percent can be defined as:

$$\text{Rejection (\%)} = \frac{\left(C_f - C_p\right)}{C_f} \times 100 \qquad (1)$$

where C_f and C_p represent the concentrations of the feed solution and permeate, respectively (Yoon *et al*., 2006). Membrane fouling is generally attributed to solute molecule accumulation on the membrane surface, and adsorption and precipitation of small solutes in the membrane pores.

Concentration polarization is the phenomenon of solute molecule accumulation near the membrane surface during the filtration process due to the flux of solvent through the membrane. The increased membrane surface concentration results in a significant decrease in the value of the mass transfer flux in a period of membrane use.

Membrane configurations can be categorized as dead-end and cross-flow filtrations. The solute molecules deposited on the membrane surface in cross-flow filtration are normally removed by the cross-flow velocity; however, some molecules may be sufficiently strongly adsorbed as to not be removed, and hence the cross-flow velocity can affect the degree of membrane fouling. In this study we have used stirred cell ultrafiltration equipment with a stirring blade placed immediately adjacent to the membrane surface (and sweeping the entire surface during half a revolution) to remove solute molecules from the surface of the membrane. This is still fundamentally different from a cross-flow system, which operates at a steady state (in the absence of fouling): the stirred cell still has varying solute concentrations above the membrane surface, even when there is no fouling. This is likely in most applications to result in a significantly time-dependent effect on the viscosity of the liquid passing through the membrane, and at the surface of the membrane. This increase in the viscosity of the liquid at the surface of the membrane with respect to time would cause a corresponding decrease in the rate of fouling molecule removal from the membrane surface due the change in the properties of the boundary layer at the surface of the membrane. However, in the experiments presented here, the main determinant of the viscosity of the solution on the retentate side of the membrane is the concentration of sucrose since this component is at 15% Brix, which is very much larger than the concentration of the dextran content (at ppm/Brix). Since sucrose is a small molecule relative to the pore sizes of the membranes used in the study (which have an MWCO greater than an order of magnitude larger than this molecule), the concentration of sucrose in the retentate side will not increase during the ultrafiltration, and therefore the viscosity of this solution and the rate of removal of molecules from the membrane surface by the stirrer blade will be approximately constant. This allows fouling models for cross-flow filtration (having constant rates of fouling molecule removal from the membrane surface) to be used with confidence in the system we investigated. The experiments showed that these models fitted the data obtained from the fouling experiments to a high degree of precision.

The objective of this research is to study the membrane separation of dextran from raw juice to determine the maximum achievable percent rejection and fluxes, to measure membrane resistances for regenerated cellulose (RC) and polyethersulfone (PES) membranes, and to understand the mechanisms of fouling in these systems.

Mathematical Models of Membrane Fouling

In this study, several models were analyzed to interpret the phenomena of membrane fouling observed during the experiments. Both pore blocking models and the cake filtration models have been investigated. There are several modes of fouling mechanism depending on the solute molecule size and shape in relation to the membrane pore size distribution, and the chemical interactions between the solute and the membrane material (Field *et al*., 1996).

Complete pore blocking: Complete pore blocking occurs when the solute molecule reaches an open pore at the surface of the membrane and blocks the pore entrance, sealing the pore closed.

Partial pore blocking: Partial pore blocking occurs when the solute molecule occupies a fraction of the pore entrance causing a reduction in the permeate flux without totally sealing the pore.

Cake filtration: Cake filtration occurs when an accumulation of solute molecules occurs over the entire surface of the membrane, increasing the resistance to flow through the membrane.

Internal pore blocking: Internal pore blocking occurs when a solute molecule that cannot be rejected by the pore entrance is adsorbed or trapped on the pore wall, thus reducing the flux through the pore, and encouraging further pore blocking. This form of fouling cannot be mitigated by flow across the surface of the membrane (or stirring at the surface, as in this study) as the blocking mechanism is internal to the membrane.

In the case of unstirred dead-end filtration, the classical constant pressure dead-end filtration equation (Hermia, 1982) has

been presented in a general form through the set of differential equations defined by Field *et al*., 1995:

$$-\frac{dJ}{dt}\left(J^{n-2}\right)=kJ \qquad (2)$$

where the exponent n and the physical meaning of the constant k depend on the mechanism of fouling.

However, the permeate flux decline in a dilute solution of the retained solute in a stirred cell filtration unit can be described by the cross-flow filtration model that is based on the classical constant pressure dead-end filtration equation (Hermia, 1982) and has been proposed earlier (Field *et al*., 1995)for a cross-flow system. The assumption in the dilute solution system is that the removal rate of the fouling molecules, B, may be considered to be constant (and related to the membrane porosity and rate of removal of particles on the membrane surface), and the rate of erosion of the cake and the back flux factor, which have been presented earlier (Field *et al*., 1995), are not a function of time.

$$-\frac{dJ}{dt}\left(J^{n-2}\right)=k(J-J^*) \qquad (3)$$

J^* is a critical flux which should not be exceeded if fouling is to be avoided. In this study J^* was considered the limiting flux ($J_{\lim}$) for large time periods; the constant k and index n take different values depending upon the fouling mechanism.

Complete Pore Blocking (n = 2)

When particles are larger than the pore size, a portion of the membrane surface used for filtration is blocked by means of pore sealing (de Barros *et al*., 2003). In the case of filtration of a dilute solution in a stirred cell, the equation can be expressed using the same model as cross-flow filtration since the fouling removal flux will be constant. The solution to the fouling model is

$$J=J_{\lim}+(J_0-J_{\lim})e^{-k_2t} \qquad (4)$$

where J is the time-dependent permeate flux, and J_0 is the initial permeate flux (t = 0). In unstirred filtration, the limiting flux ($J_{\lim}$) can be considered to be zero, since there is no foulant removal, resulting in the time-dependent flux being

$$J=J_0e^{-k_2t} \qquad (5)$$

Partial Pore Blocking (n = 1)

As in the previous section, an open pore of the membrane can be sealed by fouling by solute molecules; however, if the fouling molecule is smaller than the pore diameter, then each molecule does not necessarily block a pore completely. In the dilute solution stirred filtration case, the solution to the fouling model (with n = 1) is

$$J=\frac{J_0}{J_0k_1t+1} \qquad (6)$$

For the case of unstirred filtration, the resulting equation is

$$k_1t=\frac{1}{J_{\lim}}\left[\ln\left(\frac{J_0-J_{\lim}}{J_0}\cdot\frac{J}{J-J_{\lim}}\right)\right] \qquad (7)$$

Cake Filtration (n = 0)

The cake filtration model is used when macromolecules, which cannot enter the pores, have accumulated on the membrane surface. During the cake formation the overall resistance is composed of a cake resistance and a membrane resistance (de Barros *et al*., 2003). The cake resistance is normally dependent on the cake materials via the compressibility of the cake. The equation representing the model for stirred filtration of a dilute solution can be written as

$$J=\frac{J_0}{\sqrt{J_0^2k_0t+1}} \qquad (8)$$

and the solution of the model for unstirred filtration is given as

$$k_0t=\frac{1}{J_{\lim}^2}\left[\ln\left(\frac{J}{J_0}\cdot\frac{J_0-J_{\lim}}{J-J_{\lim}}\right)-J_{\lim}\left(\frac{1}{J}-\frac{1}{J_0}\right)\right] \qquad (9)$$

Internal Pore Blocking (n = 1.5)

In this model, the pore volume decreases due to either molecule deposits or adsorption on the pore wall. The membrane resistance increases as a consequence of pore size reduction. If internal pore blocking occurs, the fouling mechanism becomes independent of cross-flow velocity and there is no limiting value of the flux (de Barros *et al*., 2003) . For this reason, the solution of both the stirred and unstirred filtration models can be expressed by the same equation

$$\frac{1}{J^{0.5}} = \frac{1}{J_0^{0.5}} + k_{1.5}t \tag{10}$$

However in this study, all of the stirred cell results (with stirring) have a non-zero limiting flux, and hence it is clear that the internal pore blocking mechanism is not evident.

Cake Compressibility and the k Constant in Model Fitting

The constant k in the fouling models of the previous section can be used to study the mechanism of fouling. If cake formation occurs, the compressibility of the cake will be a factor determining the permeate flux decline during the filtration process. The classical empirical equation for flow through a dead-end filter is derived from Darcy's Law (Equation (11)). The permeate flux J is determined by the combination of the membrane resistance R_m and the cake resistance R_c relative to the transmembrane pressure ΔP (which includes the pressure drop across the fouling layer) divided by the solution viscosity μ (Lodge *et al*., 2004).

$$J = \frac{\Delta P}{\mu\left(R_m + R_c\right)} \tag{11}$$

R_c increases proportionally to the dry cake mass accumulation on the membrane surface:

$$R_c = \alpha \cdot m_c \tag{12}$$

α is defined as the specific cake resistance per unit mass and increases as a power law function with the transmembrane pressure, as given by

$$\alpha = \alpha_0 \cdot \Delta P^s \tag{13}$$

where S is the cake compressibility.

The physical meaning of the constant k for each blocking models has been proposed (Hermia, 1982) and simplified (Field *et al*., 1995). It is seen that the constant k in the complete pore blocking model is a linear function of the transmembrane pressure ΔP following the relation

$$k_2 = \frac{\sigma J_0}{\varepsilon_0} = \frac{\sigma \Delta P}{\mu \varepsilon_0 R_m} \tag{14}$$

where σ and ε_0 are defined as a blocked area per unit volume of filtrate and the clean membrane porosity, respectively. The constant k in the partial pore blocking model does not relate to the pressure driving force; it is given by

$$k_1 = \frac{\sigma}{A_0} \tag{15}$$

where A_0 is the initial area of pores.

If a cake is formed during the filtration process, the constant k will become a power law function of pressure in terms of the cake compressibility S. The relation can be expressed as

$$k_0 = \frac{\alpha k_c}{J_0 R_m} = \alpha_0 \mu k_c \Delta P^{S-1} \tag{16}$$

where k is the cake filtration constant.

Experimental

Materials

In this work, a high purity commercial refined sugar (approximately 99.9% purity) was used as a source of sucrose, and a high fraction dextran (approximately 250000 kDa) and a low fraction dextran (approximately 60000-90000 kDa) from Acros Organics (Fisher Scientific UK Ltd., Loughborough,

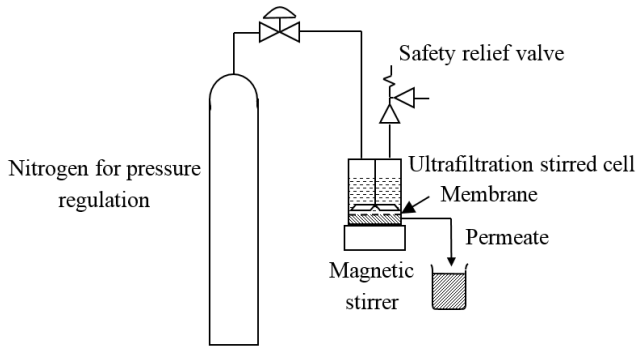

Figure 1. Schematic diagram of the stirred cell ultrafiltration unit used in the study

Table 1. Membrane separation of dextran in 15% dissolved solid of sucrose

Membrane material	MWCO (kDa)	Operating condition	Flux after long-term fouling (mL/m^2s)	Percent rejection
Regenerated cellulose (RC)	30	1 bar, 200 rpm	9.34	27.8
RC	10	1 bar, 200 rpm	8.48	86.5
RC	5	1 bar, 200 rpm	1.12	100
RC	5	2 bar, 200 rpm	2.23	100
RC	5	3 bar, 200 rpm	3.21	98.3
Polyethersulfone (PES)	5	1 bar, 200 rpm	5.25	98.3
PES	5	2 bar, 200 rpm	8.48	94.5
PES	5	3 bar, 200 rpm	8.48	90.2
PES	5	1 bar, 100 rpm	4.44	97.6
PES	5	2 bar, 100 rpm	6.76	91.9
PES	5	3 bar, 100 rpm	7.93	80.9

UK) were used as the solute component in the filtration. Chemicals required for the dextran determination in the Roberts test were ACS grade, as specified in the industrial standard test (Chen and Chou, 1993) such as absolute ethyl alcohol, sodium hydroxide, copper sulfate, sodium citrate, sodium sulfate, celite filter aid, phenol, trichloroacetic acid, and sulfuric acid.

Commercial membranes of the 2 materials in various pore sizes (RC MWCO 5000, 10000, and 30000, and PES MWCO 5000) were purchased from Merck Millipore Corporation, Billerica, MA, USA.

Methods

The methods used can be divided into 2 categories; the methods to determine the extent of dextran separation by ultrafiltration, and the methods to elucidate the fouling mechanism. Before the first category, the molecular weight distributions of both high fraction dextran and low fraction dextran were determined by gel permeation chromatography. The average values of the low and high fraction dextran molecular weight are 53000 Da and 136000 Da, respectively.

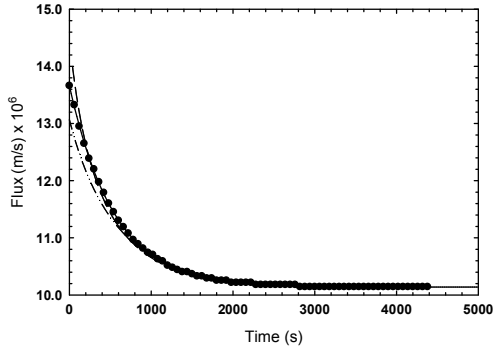

Figure 2. Flux decay and fouling mechanism at different transmembrane pressures in stirred cell membrane filtration (100 rpm) using a PES membrane at 4 bar: (●) experimental, (—) complete pore blocking (Equation 4), (—··—··) partial pore blocking (Equation 6), (———) cake filtration (Equation 8)

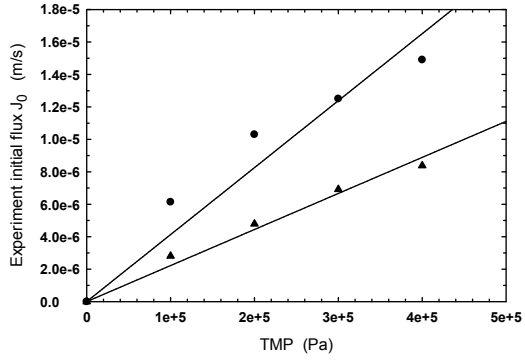

Figure 3. Membrane resistance determination (Equation 11) of 5000 MWCO polyethersulfone (●) and 5000 MWCO regenerated cellulose (▲)

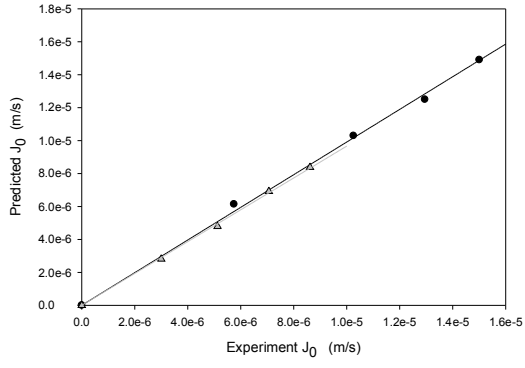

Figure 4. The accuracy of the predicted initial flux from the best fitted model of 5000 MWCO polyethersulfone (●): fitted line (—) and 5,000 MWCO regenerated cellulose (▲): fitted line (—)

Dextran separation by ultrafiltration was performed on a lab scale batch stirred cell (400 mL) membrane unit from Merck Millipore Corp, shown in Figure 1. The feed solution was prepared from 5000 ppm/Brix (a higher than average value that is commonly found in the raw sugar production process, but indicative of a process with problems in cane input to the system) of an equal weight mixture of high fraction dextran and low fraction dextran in a 15% Brix sucrose solution. The membranes in the work used 2 materials (RC and PES) and various pore sizes. The transmembrane pressure and the stirrer speed were also varied from 1 bar to 3 bar, and 100 rpm and 200 rpm, respectively. Dextran concentrations in the permeate and feed solutions were determined using the Roberts' test, which has been described in the Cane Sugar Handbook (Chen and Chou, 1993).

In the fouling model fitting, the permeate flux decline measurement was performed in a constant pressure stirred cell ultrafiltration mode by using both a 5000 MWCO RC membrane and 5000 MWCO PES membrane. For each membrane type, the transmembrane pressure was varied from 1 bar to 4 bar and the magnetic stirrer was operated in both stirred mode and unstirred mode. The permeate was collected in a receiver on a digital balance and the volume flux was calculated from the rate of increase in weight. Fouling models were fitted using the program SigmaPlot 9.0 (2004, Systat Software, Inc., Chicago, IL, USA).

Results

Membrane Separation of Dextran

The permeate flux after long-term fouling (after a steady permeate flow was achieved) and the percent rejection on the dextran separation by using ultrafiltration processes are shown in Table 1. Note that the feed solution is composed of 5000 ppm/Brix of dextran and 15% Brix of sucrose.

These results show that the permeate flux after long-term fouling of the PES membrane is higher than that of the RC membrane under the same conditions. If the differences in the dextran cake resistance and the resistance due to blocking of dextran molecules between the 2 types of membranes can be neglected, then the membrane resistance is sufficient to explain this result. In addition, the trend of the permeate flux after long term fouling increases with increasing the transmembrane pressure and increasing the magnetic stirrer speed.

In the comparisons of the percent rejection it can be seen that the dextran rejection by the RC membrane is higher than that of the PES membrane of the same pore size. The transmembrane pressure and the stirrer speed also have a strong influence on the percent rejection, and in particular a lower stirrer speed will result in a smaller value of the percent rejection because of the concentration polarization effect.

Model Fitting

The permeate flux decline for each condition was fitted to the fouling models using SigmaPlot 9.0. For unstirred dead-end filtrations, where fouling results in the flux eventually decaying to zero, the models were fitted with 2 parameters, a rate constant (k) and an initial flux (J_0). For stirred cell filtrations, where a finite limiting flux was achieved due to the removal of the fouling layer, the models were fitted using 3 parameters, a rate constant (k), an initial flux (J_0), and a limiting flux (J_{lim}). An example of a flux decay curve is shown in Figure 2.

The cake filtration model adequately represents the overall fouling mechanism in unstirred dead-end filtration using both PES and RC membranes. This is a result of the dextran molecules rejected by the membranes accumulating on the membrane surface without dispersion by agitation.

In filtration with stirring at 100 rpm, the fouling behavior is controlled by the complete pore blocking model for both membrane materials. However, the model fitting in the 2 figures does not distinguish between 2 fouling mechanisms (cake filtration and complete pore blocking) very well, so the fouling mechanism will be reconsidered in a later section based on an analysis of more complete mathematical descriptions of the behavior. The behavior of the same membrane filtrations at 200 rpm showed similar results; however, the magnitude of the flux reduction due to fouling was smaller and the rate of fouling was also lower. The

results for the RC membranes at 200 rpm and transmembrane pressures of 1 and 2 bar did not transmembrane pressures of 1 and 2 bar did not display sufficient levels of fouling to enable models to be fitted to a suitable level of significance in the fitted parameters.

Membrane resistance determination by Darcy's Law

The initial fluxes through both the PES and RC membranes were plotted against the transmembrane pressure to determine the membrane resistances as shown in Figure 3. The accuracy of the predicted initial flux from the best fitted model for the PES membrane and for the RC membrane is 99% and 97%, respectively. The comparison between the experimental results and the fitted model is very good. In addition, the accuracy of the predicted initial flux compared to the measured initial flux points for both membranes is shown in Figure 4.

The membrane resistance can be calculated from Darcy's law (in the absence of the cake resistance term since the results are initial flux values) using the viscosity of the permeate and the slope of the plots in Figure 3. The resistances of the PES membrane and of the RC membrane are equal to 1.80×10^{13} m^{-1} and 3.36×10^{13} m^{-1}, respectively. The lower resistance is the reason for the higher flux

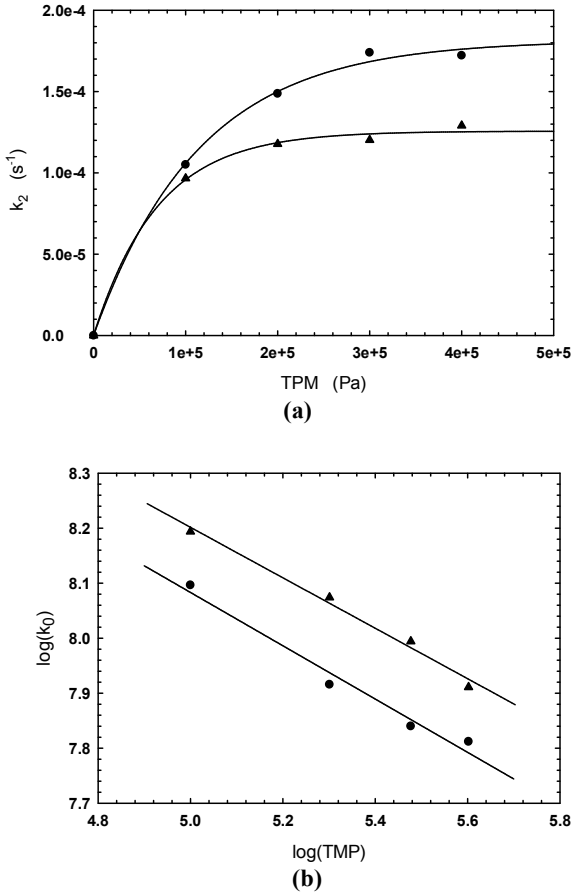

Figure 5. The relationship between the constant *k* of (a) the complete pore blocking model (Equation 14) and (b) the cake filtration model (Equation 16); and the transmembrane pressure for unstirred dead-end filtration: polyethersulfone (●) and regenerated cellulose (▲)

through the PES membrane in the membrane separation of dextran.

Fouling Mechanism Prediction by Using the Constant *k* from Model Fitting

The constants for both the complete pore blocking model (k_2) and the cake filtration model (k_0) were plotted relative to both the transmembrane pressure and the log of the transmembrane pressure to predict the mechanism of the membrane fouling. The plots used to test the fouling mechanism in unstirred filtration and stirred filtration (at 100 rpm) are shown in Figure 5 and Figure 6, respectively.

Figure 5 demonstrates that the complete pore blocking model is not correct in this instance because the constant k_2 is not a linear function of the transmembrane pressure. The constant k_0 is a power law relation to the pressure (Figure 5(b)), so it is clear that the fouling mechanism is cake filtration. Moreover, the compressibility (S) of the dextran cake layer and the constant coefficient ($\alpha_0 \mu k_c$) can be calculated from the slope and the intercept of the plot of k_0, respectively. The cake compressibilities of both membrane materials are not significantly different because the constant depends only on the characteristic of the dextran molecules fouling the membrane, not on the membrane itself. The

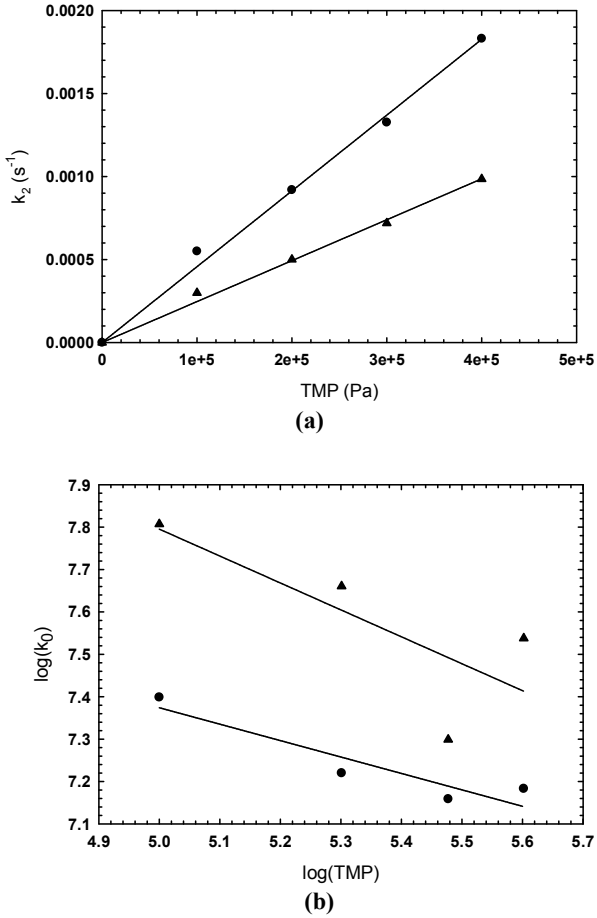

Figure 6. The relationship between the constant *k* of (a) complete pore blocking model (Equation 14) and (b) cake filtration model (Equation 16); and the transmembrane pressure for stirred cell filtration at a speed of 100 rpm: polyethersulfone (○) and regenerated cellulose (Δ)

cake compressibility is between 0.52-0.54, and the constant coefficient is equal to 3.16×10^{10}.

However if the ultrafiltration was performed with magnetic stirring at 100 rpm, the fouling mechanism becomes complete pore blocking, as exhibited in Figure 6. Moreover a qualitative comparison in terms of the fouling of both membranes can be deduced, suggesting that the fouling of the PES membrane is faster than that of the RC membrane, as evidenced by the magnitude of the coefficient (k_2).

Conclusions

Dextran separation from a synthetic raw juice solution of 15% Brix could be achieved using an RC membrane with an MWCO 5000 Da, with the process being able to achieve a high percent rejection and a suitable flux. In addition, it was found that the percent rejection and the permeate flux can be improved by increasing the stirrer speed, which in an industrial cross-flow unit would equate to having a higher cross-flow velocity on the retentate side of the module. Although the permeate flux of the 5000 MWCO PES membrane is higher than that of the RC membrane, its percent rejection is smaller and its performance relating to fouling is not as good.

The resistances of the PES membrane and the RC membrane are equal to 1.80×10^{13} m^{-1} and 3.36×10^{13} m^{-1}, respectively. Moreover the accuracy of the predicted initial flux from the best fit model of the PES membrane and the RC membrane was 99% and 97%, respectively.

In this study it was found that the fouling due to dextran molecules (mean molecular weight 53000 Da) in unstirred dead-end ultrafiltration is cake filtration. The compressibility of the dextran cake layer is in the range of 0.52-0.54. When filtration is performed with stirring at 100 rpm or greater, the fouling mechanism becomes complete pore blocking.

List of Symbols

A_0 initial area of pores (m^2)
C_f concentration of the feed solution (kg m^{-2})
C_p concentration of the permeate (kg m^{-3})
J permeate flux (ms^{-1})
J_0 initial flux (ms^{-1})
J_{lim} limiting flux (ms^{-1})
k phenomenological coefficient (units depend on mechanism)
k_c cake filtration constant (kg m^{-3})
m_c cake mass per unit area (kg m^{-2})
n general index (units depend on mechanism)
ΔP transmembrane pressure (Pa)
R_c cake resistance (m^{-1})
R_m membrane resistance (m^{-1})
S cake compressibility
t filtration time (s)

Greek Letters

α specific cake resistance per unit mass (m kg^{-1})
α_0 empirical constant (m $(kgPa^n)^{-1}$)
ε_0 clean membrane porosity
μ dynamic viscosity (Ns m^{-2})
σ blocked area per unit volume of filtrate (m^{-1})

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