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The influence of feed flow channel diameter on frictional pressure drop, membrane performance and process cost in full-scale tubular ceramic membranes

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ABSTRACT

Tubular ceramic membranes with many, small-diameter feed flow channels have the advantage of being less costly per unit membrane area. For this reason, membranes with many small channels are often favored over those with fewer larger channels. We have theoretically studied how the diameter of the feed flow channels influences the frictional pressure drop, the membrane performance (flux and retention) and the process cost by performing basic pressure drop calculations and using experimental data from bench-scale experiments. The investigation was carried out on a model microfiltration process consisting of the separation of yeast cells from polyethylene glycol (PEG) macromolecules. For a membrane with 2.5 mm channels the average flux and PEG retention were 112 L/m^{2}h and 22%, respectively, differing significantly from a membrane with 6.0 mm channel diameter (131 L/m^{2}h, 17%), under the same hydrodynamic conditions. The choice of channel diameter also has a considerable impact on the process cost. While the costs were similar for membranes with 6.0 and 3.8 mm diameter channels, they were about 55% higher using a membrane with 2.5 mm channels. This high cost was mainly attributed to the high frictional pressure drop along the membrane, which increases the energy required for pumping.

Keywords: Microfiltration; Tubular membrane; Channel diameter; Frictional pressure drop; Cake formation; Cost estimate

1. Introduction

Tubular ceramic membranes are especially suitable in applications where the feed stream contains a relatively high proportion of large particles, and where the membranes are exposed to extreme pH and temperature. In general, the size of the feed flow channels in tubular ceramic membranes is larger than in alternative membrane modules (Cheryan, 1998). This more open channel geometry reduces the risk of blockage of the feed channels, reducing the need for costly pretreatment before microfiltration.

The cost per unit membrane area of a tubular ceramic element usually decreases with the number of parallel feed flow channels in the element. It is therefore common to use membranes with more densely packed, narrow channels in applications where channel blockage is not a problem. However, reducing the diameter of the feed flow channels leads to an increase in frictional pressure drop along the membrane element. This is undesirable as it increases the energy required for pumping, and thus has a direct impact on the operating cost (Nordin and Jönsson, 2010). Moreover, the frictional pressure drop can cause considerable differences in the transmembrane pressure (TMP) between the inlet and outlet of tubular membrane elements, affecting the flux during both ultrafiltration and microfiltration (Nordin and Jönsson, 2010, 2009; Piry et al., 2008; Yeh et al., 2010,2004; Yeh, 2002).

Furthermore, the frictional pressure drop can promote the formation of cake layers on the membrane surface, resulting in poorer separation by the membrane. Piry et al. (2008) observed a markedly higher retention of proteins at the inlet than at the outlet of a tubular ceramic membrane during microfiltration of skim milk, due to the formation of a cake layer. Several fundamental studies using various binary mixtures of a cake-forming substance and macromolecules confirmed

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that cake layers formed at high TMP could increase macromolecule retention during microfiltration (Hwang et al., 2006; Hwang and Cheng, 2003; Jiraratananon et al., 1998; Kawakatsu et al., 1993).

The choice of channel diameter in tubular ceramic membranes is thus not straightforward. The objective of this work was therefore to investigate how the diameter of the feed flow channels influences the frictional pressure drop, the membrane performance (i.e. the flux and macromolecule retention), and the cost of a full-scale microfiltration process. Bench-scale microfiltration experiments were carried out using model solutions of polyethylene glycol (PEG), an aqueous suspension of washed yeast cells, and a mixture of PEG and washed yeast. The aim of these experiments was to study how the high TMP at the inlet of full-scale tubular membranes influences flux, cake layer formation, and the retention of PEG macromolecules. The experimental results were then used to predict the membrane performance and the cost of full-scale processes using membranes differing in the number and diameter of feed flow channels.

2. Materials and methods

2.1. Preparation of the PEG solutions and yeast suspension

Five different solutions were used in this investigation: three aqueous PEG solutions containing 5, 10, and 15 g PEG/L, an aqueous yeast suspension containing 10 g yeast/L, and a mixture of yeast and PEG at concentrations of 10 and 5 g/L, respectively. The PEG solutions were prepared from dry PEG (PEG 10, Merck KGaA, Darmstadt, Germany) by dissolution of the PEG flakes in deionized water. The molecular mass of PEG 10 is in the range 9 to 11.25 kg/mol.

Dry yeast of the strain *Saccharomyces cerevisiae* (Jästbolaget AB, Sweden) was used for the preparation of the yeast suspension. It is well known that dry yeast contains considerable amounts of extracellular polymeric substances (EPS) that originate from yeast metabolism. These substances can adsorb onto the pore walls, leading to extensive fouling of microfiltration membranes (Beier and Jonsson, 2009; Foley et al., 1995; Hughes and Field, 2006; Negaresh et al., 2007; Ye et al., 2005). In this investigation, the EPS were removed from the dry yeast prior to the preparation of the yeast suspension, using a procedure consisting of three consecutive washing steps with deionized water. Sixty grams of dry yeast was mixed with 2.4 L deionized water, and centrifuged for 15 min at 3700 rpm in a laboratory centrifuge (model C4.12, Jouan S.A., Saint-Herblain, France). Most of the supernatant (2.1 L) was removed from the centrifuged suspension and the same amount of deionized water was added to the sediment for the next washing step. After three washing steps the yeast sediment was mixed with deionized water to obtain a total volume of 6 L yeast suspension for the bench-scale experiments. Analysis (described in Section 2.3.3) revealed that about 99% of the initial amount of EPS had been removed using this washing procedure.

Mixtures of washed yeast and PEG were prepared by preparing a yeast suspension at a concentration of 10 g/L, according to the procedure described above, and then adding the required amount of dry PEG flakes to the yeast suspension. The pH of the solutions containing yeast cells was about 4.2.

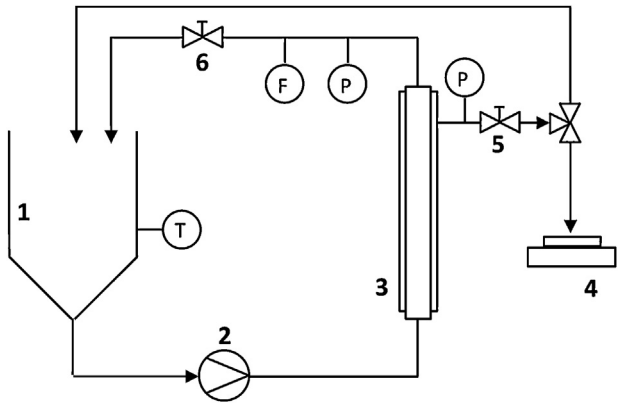

Fig. 1 – Schematic illustration of the experimental setup. 1: feed tank, 2: circulation pump, 3: membrane module, 4: electronic balance, 5: permeate valve, 6: retentate valve, P: pressure transmitters, T: temperature probe, F: flow meter.

2.2. Membrane filtration equipment

2.2.1. Membrane

The experiments were performed with a tubular, ceramic membrane manufactured by Atech Innovations GmbH, Gladbeck, Germany. The membrane was made of α-alumina oxide and consisted of a single feed channel with an inner diameter of 6 mm, a wall thickness of 2 mm, and a channel length of 250 mm. The nominal pore size of the membrane was 0.2 μm, and the total filtration area 47 cm^2.

2.2.2. Experimental setup

The tubular membrane was installed in the filtration setup shown schematically in Fig. 1. The setup consisted of a feed tank (1), a centrifugal pump (CRN2-40 AAA, Grundfos AB, Angered, Sweden) (2) for circulation of the feed solution, and a membrane module (3). Pressure transmitters (type 4 AP-30-020/064, JUMO GmbH & Co. KG, Fulda, Germany) were installed on the retentate side and permeate side of the membrane, and a rotameter was installed after the module to measure the cross-flow. A Pt100 temperature probe allowed measurement of the temperature in the feed tank, and an electronic balance (PL6001-S, Mettler-Toledo Inc., Columbus, OH) (4) was used to determine the permeate flow through the membrane. The TMP was calculated as the pressure difference between the retentate and permeate pressure, and was adjusted with a valve (5) on the permeate side. The cross-flow velocity was regulated by changing the pump speed with a frequency converter (SL 750-1, Scandialogic AB, Sweden) and adjusting a valve (6) installed after the membrane module. All data obtained from the instruments were recorded using LabView software (National Instruments Co., Austin, TX).

2.3. Experimental procedure

2.3.1. Membrane cleaning

Before starting an experiment, the membrane was cleaned for 2 h with a 1 wt% solution of the alkaline cleaning agent Ultrasil 11 (Ecolab AB, Älvsjö, Sweden). The TMP, the cross-flow velocity and the temperature during cleaning were 0.5 bar, 3.5 m/s and 50 °C, respectively. The system was rinsed thoroughly with deionized water after cleaning, and the pure water flux of the membrane was measured at various TMPs. If single alkaline cleaning was not sufficient to recover the pure