Contents lists available at ScienceDirect

Chemical Engineering Research and Design

journal homepage: www.elsevier.com/locate/cherd

Reduction of energy demand by use of air sparging during ultrafiltration of alkali-extracted wheat bran hemicelluloses

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ARTICLE INFO

Article history:
Received 4 January 2018
Received in revised form 18 July 2018
Accepted 1 August 2018
Available online 8 August 2018

Keywords:
Two-phase gas–liquid flow
Ultrafiltration
Hemicelluloses
Wheat bran
Flux enhancement

ABSTRACT

The flux during membrane filtration can be enhanced by the use of a two-phase gas–liquid flow. This has been shown to be an energy-efficient alternative to increasing the cross-flow velocity. In this work, air sparging was used to increase the flux during ultrafiltration of alkali-extracted wheat bran hemicelluloses. Batch filtration was performed in a pilot unit with a ceramic ultrafiltration membrane with a nominal cut-off of 10 kDa. Parametric studies with and without air sparging were performed at temperatures of 30 °C, 50 °C and 80 °C and cross-flow velocities of 1, 3, 5 and 7 m/s. The limiting flux was not affected by air sparging at 30 °C, while a slight increase was observed at 50 °C and a considerable increase was obtained at 80 °C. Air sparging reduced the energy demand per m^3 permeate produced during dead-end batch ultrafiltration at 80 °C and 1 m/s from 0.96 kWh/m^3 to 0.51 kWh/m^3.

1. Introduction

It is well-known that two-phase gas–liquid flow can improve both heat transfer (Groothuis and Hendal, 1959) and mass transport (Horvath et al., 1973). The first attempts to utilize a two-phase flow, i.e. gas sparging, to increase the flux in membrane filtration, was reported in the early 1990s (Cui and Wright, 1994; Lee et al., 1993). Since then, several studies have been carried out on gas sparging, often with model solutions of dextran (Cui and Wright, 1994; Cui and Wright, 1996; Ghosh and Cui, 1999; Cheng, 2002; Cheng and Wu, 2003; Smith and Cui, 2004; Cheng and Lin, 2004; Taha et al., 2006), yeast (Lee et al., 1993; Sur and Cui, 2001; Hwang and Hsu, 2009) or bentonite (Cabassud et al., 1997; Mercier et al., 1997; Laborie et al., 1998). The effect of gas injection in these trials has varied, from having no effect at all on the flux (Ducom and Cabassud, 2003), to a 5-fold increase in flux (Fadaei et al., 2007).

The influence of air sparging during ultrafiltration (UF) of lignocellulosic material has, to the best of our knowledge, not been investigated previously. Hemicelluloses are high molecular mass biopolymers that can be extracted from plant biomass such as wheat bran, and then concentrated and purified using UF, for use as barrier films (Hansen and Plackett, 2008; Zhu Ryberg et al., 2013) and hydrogels (Carvajal-Millan et al., 2012; Maleki et al., 2015). The high molecular weight of hemicelluloses makes the viscosity of the solution high, and thus the UF flux is low (Krawczyk et al., 2011). Previous investigations of UF of wheat bran solutions have shown that the flux can be improved by the removal of suspended matter by prefiltration using dead-end filtration (Krawczyk et al., 2013; Thuvander et al., 2014) and partial degradation of the hemicelluloses using heat treatment (Arkell et al., 2013).

The aim of this study was to investigate the influence of gas sparging on flux during UF of hemicelluloses extracted from wheat bran. The hemicellulose solution was prefiltered with diatomaceous earth, and gas sparging by air injection was applied during UF at 30 °C, 50 °C and 80 °C.

2. Theory

Extensive research has been applied to the theoretical modelling of the UF process. The most widely used models for UF are the gel polarization model (Michaels, 1968; Blatt et al., 1970; Nakao et al., 1979; Fane et al., 1981; Bhattacharjee et al., 1996; Wijmans et al., 1984):

$$J = k \ln\left(\frac{C_m - C_p}{C_b - C_p}\right) \quad (1)$$

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https://doi.org/10.1016/j.cherd.2018.08.001

the osmotic pressure model (Wijmans et al., 1984; Kedem and Katchalsky, 1958; Goldsmith, 1971; Wijmans et al., 1985; Nabetani et al., 1990; Prádànos et al., 1995):

$$J = \frac{\Delta P - \Delta\pi}{\mu_p \cdot R_m} \tag{2}$$

and the resistance-in-series model (Wijmans et al., 1985; Chudacek and Fane, 1984; van den Berg and Smolders, 1989; Roy and De, 2015):

$$J = \frac{\Delta P}{\mu_p \cdot (R_m + R_c + R_{cp})} \tag{3}$$

where J is the flux, k the mass transfer coefficient, C_m, C_p and C_b the concentrations at the membrane surface, in the permeate and in the bulk solution, ΔP the transmembrane pressure (TMP), $\Delta\pi$ the osmotic pressure difference across the membrane, μ_p the viscosity of the permeate and R_m, R_c and R_{cp} the filtration resistance of the membrane, the cake layer on the membrane surface, and the concentration polarization layer.

The mass transfer coefficient k in Eq. (1) is related to the operating conditions in the system by the dimensionless correlation between the Sherwood number (Sh), the Reynolds number (Re) and the Schmidt number (Sc):

$$Sh = \frac{k \cdot d_h}{D} = a \cdot Re^b \cdot Sc^c \tag{4}$$

where

$$Re = \frac{d_h \cdot CFV \cdot \rho_b}{\mu_b} \tag{5a}$$

and

$$Sc = \frac{\mu_b}{\rho_b \cdot D} \tag{5b}$$

where d_h is the hydraulic diameter of the flow channel, D the diffusion coefficient, a, b and c parameters depending on the operating conditions, CFV the cross-flow velocity, and ρ_b and μ_b the density and dynamic viscosity of the bulk solution.

For laminar flow, the Leverque correlation can be used (Cheryan, 1998):

$$Sh = 1.86\,(Re \cdot Sc)^{0.33} \tag{6a}$$

For turbulent flow, the Chilton–Colburn correlation can be used (Cheryan, 1998):

$$Sh = 0.023 \cdot Re^{0.8} \cdot Sc^{0.33} \tag{6b}$$

The mass transfer coefficient is dependent on whether the flow is laminar or not, and is influenced by the CFV, as described by Eqs. (7a) and (7b)

$$k \propto CFV^{0.33} \quad (\text{laminar flow}) \tag{7a}$$

$$k \propto CFV^{0.8} \quad (\text{turbulent flow}) \tag{7b}$$

The influence of temperature on k is:

$$\text{none} \quad (\text{laminar flow}) \tag{8a}$$

$$k \propto \left(\frac{\rho_b}{\mu_b}\right)^{0.47} \quad (\text{turbulent flow}) \tag{8b}$$

In addition to its influence on k, the temperature also affects the flux through the permeate viscosity, μ_p, as shown in Eqs. (2) and (3).

An increase in membrane flux can thus be achieved by increasing the CFV and the temperature. An alternative way of improving the flux, with a smaller energy demand, is to use gas sparging. Gas–liquid two-phase flow creates hydrodynamic instabilities in the membrane flow channels that limit the thickness of the boundary layer, reducing concentration polarization and the formation of filter cakes on the membrane surface. The flow pattern models used in vertical tubes are bubble flow, plug flow, slug flow, churn flow and annular flow (Wibisono et al., 2014). Bubble flow is the model used at low gas/liquid ratios and annular flow at high gas/liquid ratios. The gas/liquid ratio, θ, is defined as:

$$\theta = \frac{u_G}{u_L + u_G} \tag{9}$$

where u_G and u_L are the superficial velocity of the gas and the liquid. The transition between flow patterns is influenced not only by the gas and liquid velocities, but also by the characteristics of the liquid, such as density, viscosity and dissolved substances (Wibisono et al., 2014).

3. Materials and methods

3.1. Hemicellulose solution

Hemicelluloses were extracted from destarched wheat bran with a concentrated sodium hydroxide solution. The hemicellulose solution was dead-end filtered with 2 wt% diatomaceous earth (Dicalite 4500, Dicalite, Burney, CA, USA) as a filter aid to remove colloidal material (Krawczyk et al., 2013). The total solids content was about 40 g/L, of which about 30 g/L was NaOH, and about 7 g/L hemicelluloses (Thuvander et al., 2014).

3.2. Membrane

The membrane used in the experiments was a 10 kDa, tubular ceramic α-Al_2O_3 membrane, with an active layer of TiO_2 (Atech Innovations GmbH, Gladbeck, Germany). The 1000 mm long membrane had 7 parallel 6 mm diameter feed flow channels. The total membrane area was 0.132 m^2.

3.3. Equipment

The experiments were performed in a pilot unit with injection of air between the feed pump and the membrane module, as shown in Fig. 1. The set-up included two 200-L tanks. Tank 1 was used as the feed tank during the experiments, and tank 2 was used during start-up and membrane cleaning. The outlet of the pipe returning the retentate to tank 1 was positioned below the liquid surface to avoid mixing of air and liquid in the tank. The temperature, T, the pressure at the inlet (P_1), outlet (P_2) and on the permeate side of the membrane (P_3), and the feed, Q_{feed}, and permeate, Q_p, flow were measured and stored in a computer. The air flow, Q_{gas}, was measured manually using a rotameter. The TMP was calculated from the relation:

$$TMP = \frac{P_1 + P_2}{2} - P_3 \tag{10}$$