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Modeling prediction of the process performance of seawater-driven forward osmosis for nutrients enrichment: Implication for membrane module design and system operation

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ABSTRACT

Seawater-driven forward osmosis (FO) has been successfully applied in wastewater nutrient recoveries in laboratory-scale studies. In this study, modeling simulations were performed to gain a better understanding on the performance of a large-scale FO with two practically applied module configurations, a plate-and-frame module and a submerged hollow fiber module. The mathematical models were derived based on the mass balance and permeate flux model, taking into account the water and solute bidirectional transportation and the influence of internal concentration polarization. Iterative method was adopted to solve the highly non-linear and implicit equations in the models. The models were afterward applied to simulate a seawater-driven FO process for enriching nitrogen and phosphorous in wastewater. The simulation results show that approximately 30–40% of bulk osmotic pressure difference is used as the effective driving force in the processes with both plate-and-frame and submerged hollow fiber modules. The FO process is terminated at the osmotic equilibrium state, which is independent of the module configuration. However, a submerged hollow fiber module can meet the equivalent performance with much more compact module dimension compared with a plate-and-frame module. In addition, the influences of membrane module dimensional and operational parameters, including channel height, membrane length, assistant hydraulic pressure, and inlet cross-flow velocity in feed and draw stream on process performance, were further discussed in a plate-and-frame module. These developed models can be applied to design membrane modules and optimize the operational conditions in scale-up FO processes in future studies and applications.

1. Introduction

Forward osmosis (FO), an osmotic energy driven membrane process, has been recently receiving significant research and application interest due to its lower energy consumption and membrane fouling propensity compared with nanofiltration (NF) and reverse osmosis (RO) that operate at very high hydraulic pressure. Hence, an increasing number of studies and applications of FO have been reported during the past decade, including involvement in seawater and brackish water desalination [1–3], commercial materials production [4–7], and wastewater treatment [8–10].

One practical application of FO is to enrich nitrogen and phosphorous in various wastewater sources for further nutrients recovery [8,11–15]. High levels of phosphate retention (> 90%) and improvements in nitrogen retention (up to 80%) from nutrient-rich wastewater source (e.g., urban source-separated urine and digested sludge) and dilute wastewater source (e.g., treated municipal wastewater) have been achieved in laboratory-scale FOs [11–14]. However, laboratory experiments were typically carried out with small membrane area, the influence caused by the complicated feed and draw solutes flow condition was a rarely concern, and the results were limited to predict the performance of FO within a large-scale process. Inside a practical FO process, the driving force determines the transmembrane migration of substances and, meanwhile, strongly depends on the changes of concentration caused by the mass transfer between feed and draw solution [16]. Deeper understanding of these issues as well as their proper consideration in module design and process operation will remarkably enhance the engineering and economic performance of FO processes. Model-based simulation, design and control provide valuable methods and tools to approach this challenge. Extensive studies using modeling approaches have been implemented to meet various application purposes such as seawater

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desalination [17–19], FO membrane bioreactor (FoMBRs) [16,20], and fertilizer drawn FO (FDFO) [21,22]. Most of these studies focused on the transfer behavior of water and draw solutes in FO depending on their specific application goals. Few researchers have simultaneously identified the performance of feed solute concentration in a scale-up FO process [23]. To apply FO on nutrient recovery from wastewater, it is essential to simultaneously depict the behaviors of nutrient solutes, draw solutes as well as net water within integral models.

Concentration polarization phenomena play critical roles in impacting the performance of FO on both water permeation [24–26] and solute migration [27,28]. Specifically, the composite or asymmetric structures of FO membranes give rise to a significant variation of concentration profile inside the porous support layers, referred to as the internal concentration polarization (ICP), substantially declining the driving force across the active layers. During a practical FO process, the overall performance will be further influenced by the ICP phenomenon, which sensitively responds to the concentration of feed solution and dilution of draw solution. This eventually adds more complexity to quantitatively predict the performance of FO in a scale-up process. In contrast, the influence of the external concentration polarization (ECP) is generally surmounted by the ICP, and it can be minimized by applying spacers and optimizing the flow condition inside membrane modules [18,20].

Module selection and design is of paramount importance for promoting the process performance of membrane technology. There are two typical membrane configurations that can be applied in FO modules, namely, flat sheet membranes and tubular membranes. The fabrication and applications of both membrane configurations have been widely reported in the literature [2,7–9]. A systematical comparison helps to understand the properties of each module configuration and is valuable for selecting a suitable module for a particular target in future applications of FO technology.

The overall goal of this study is to perform modeling simulation for a scale-up FO process with two commercially used module configurations, the plate-and-frame module and the submerged hollow fiber module. The mathematical models consist of transmembrane water flux, reverse solute diffusion, and feed solutes permeation with consideration of the impact from ICP. Furthermore, the influence of dimensional and operational parameters such as channel height, membrane length, assistant hydraulic pressure, and feed and draw stream inlet velocity combinations in a plate-and-frame module is discussed to provide knowledge on optimizing the module design and process operation in FO applications.

2. Modeling

2.1. Water flux and bi-directional solutes flux in FO

Water transportation through the selective layer of FO membrane was generally described using the solution-diffusion model [17,18,21,29,30], where water flux in FO is driven by the integration of effective osmotic pressure difference (i.e., draw osmotic pressure minus feed osmotic pressure) and hydraulic pressure difference across the selective layer. It is worth mentioning that in general understanding, hydraulic pressure is not the major driving force for FO process, while, within a practical FO process, hydraulic pressure difference may be provided across the membrane due to the pressure drop occurred in each stream. Furthermore, assistant hydraulic pressure, as an effective approach to enhance the performance of the FO process, is worthy to be taken into account in the discussion of this work. Hereby, a comprehensive expression of the transmembrane water flux is given as:

$$J_W = A(\Delta\pi_{\mathrm{eff}} + \Delta P) \tag{1}$$

where A is the water permeability coefficient, $\Delta\pi_{\mathrm{eff}}$ the effective osmotic pressure difference across membrane selective layer, and ΔP the hydraulic pressure difference across the membrane.

According to the Morse equation [31], the osmotic pressure of diluted solution can be directly related to the concentration of the solution:

$$\pi = nzcRT = Hc \tag{2}$$

where n is the dimensionless van't Hoff factor, z the valence, c the mole concentration, R the ideal gas constant, T the thermodynamic temperature, and H the osmotic proportionality coefficient of the solute. According to the previous literature [29], ICP is the main reason that causes the "lower-than-expected" flux in FO, and an expression of effective osmotic pressure accounting for ICP is given by [24,26]:

$$\Delta\pi_{\mathrm{eff}} = \frac{\pi_{\mathrm{Hi}}\exp(-K_d J_W) - \pi_{\mathrm{Low}}}{\frac{B_d}{J_W}\left[1 - \exp(-K_d J_W)\right] + 1} \quad (\text{AL-FS, diluted ICP}) \tag{3}$$

$$\Delta\pi_{\mathrm{eff}} = \frac{\pi_{\mathrm{Hi}} - \pi_{\mathrm{Low}}\exp(K_d J_W)}{\frac{B_d}{J_W}\left[\exp(K_d J_W) - 1\right] + 1} \quad (\text{AL-DS, concentrative ICP}) \tag{4}$$

where π_{Hi} and π_{Low} are the osmotic pressures in bulk of draw and feed solution, respectively, B_d the draw solute permeability coefficient, and K_d the resistance to draw solute diffusion within membrane porous support layer, given by:

$$K = \frac{\tau t}{D_s \varepsilon} \tag{5}$$

where τ, t and ε are the tortuosity, thickness and porosity of the porous support layer, respectively, and D_s the solute diffusion coefficient in water.

Unlike the pressurized membrane processes, the FO process contains a bi-direction solutes diffusion across the membrane, reverse draw solute diffusion and forward solute flux, due to its double solutions system. In particular, the bi-direction solutes diffusion plays a significant role in a feed-concentration-aimed FO process. In this model, the reverse draw solute diffusion and forward solute flux are simultaneously considered.

The reverse draw solute diffusion, J_{DS}, has been reported as a function of A, B_d, and the osmotic proportionality coefficient of the draw solute, H_d [32,33], as expressed in the following equation:

$$\frac{J_{DS}}{J_W} = \frac{B_d}{AH_d} = \xi_d \tag{6}$$

where ξ_d is the reverse flux selectivity of draw solute in FO.

Conversely, the forward solute flux from the feed to the draw solution is driven by the concentration gradient across the active layer of the membrane and is given considering the influence of the ICP as follows [27]:

$$J_S = \frac{B_f}{1 + B_f/J_W} c_f \quad (\text{AL-FS}) \tag{7}$$

$$J_S = \frac{B_f\exp(J_W/K_f)}{1 + B_f\exp(J_W/K_f)/J_W} c_f \quad (\text{AL-DS}) \tag{8}$$

where B_f is the feed solute permeability of the membrane, K_f the resistance to feed solute diffusion within the membrane porous support layer, and c_f the solute concentration in the bulk of the feed solution.