Journal of Membrane Science 476 (2015) 311–320

Contents lists available at ScienceDirect

Journal of Membrane Science

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

Scale-up characteristics of membrane-based salinity-gradient power production

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ARTICLE INFO

Article history:
Received 20 May 2014
Received in revised form
8 October 2014
Accepted 10 October 2014
Available online 18 November 2014

Keywords:
Salinity gradient power
Pressure-retarded osmosis
Reverse electrodialysis
Renewable energy

ABSTRACT

The controlled mixing of streams with different salinity is a potential route for clean and renewable base-load power generation. Here, a comprehensive process model has been developed for pressure-retarded osmosis (PRO) accounting for full-scale system losses such as viscous dissipation, external mass transfer and equipment efficiency. Also, an existing model for reverse electro-dialysis (RED) is adapted to account for analogous full-scale system losses. The models are used to predict practical power densities and process efficiencies. The projected power density for PRO (using best available membranes) is much lower than generally predicted by extrapolation of experimental data. For example, a power density of 4 W/m^2 extrapolated from laboratory experiments actually yielded negative power at full-scale. The maximum power density for PRO is doubled as the hydraulic energy recovery (HER) efficiency is increased from 90% to 99%. Furthermore, the operating pressure, load voltage, and crossflow velocities typically applied in laboratory studies appear much too high to be practical in full-scale PRO and RED systems. Notably, RED systems exhibit a lower system size required for achieving a given degree of mixing, compared with PRO. For both processes, maximum energy efficiency does not occur at thermodynamic equilibrium due to hydraulic losses. Finally, maximum power density appears to be an inadequate parameter for assessing full-scale PRO/RED process feasibility because both processes could produce the same maximum power density, yet exhibit different power outputs and efficiencies and system sizes.

1. Introduction

Clean, renewable energy may be produced by controllably mixing streams of different salinity [1,2]. This is achieved by employing membranes that facilitate selective transport of either solute or solvent, resulting in different process characteristics with various advantages and disadvantages. Pressure-retarded osmosis (PRO) and reverse electro-dialysis (RED), the two most popular salinity gradient power (SGP) technologies, are discussed in detail in the literature (see, for example, Refs. [1,3–9]). While both PRO and RED involve mixing dilute and concentrated solutions, the energy conversion mechanism achieved via mixing is fundamentally different. PRO involves the diffusion of water molecules across a semipermeable membrane from a dilute feed stream into a pressurized concentrated stream, augmenting its volume upon dilution. The volume-augmented feed is then passed through a hydro-turbine where the mechanical energy is converted into electricity. RED involves the diffusion of salt ions across charge-selective ion exchange membranes, creating an ionic flux that is converted into an electron flux at an electrode surface, and power is harnessed through an applied load in an external circuit.

Previous modeling work on SGP technologies has focused on the maximum power density produced through PRO or RED. However, maximum power density alone is an insufficient metric for designing a large-scale, realistic SGP system. Even in an ideal system, maximum power density will only be achieved at the system inlet where the concentration gradient is the largest, but downstream mixing of the streams inevitably lowers the power density. Conversely, energy efficiency, defined as the ratio of produced-to-available energy, is maximized at the point of complete mixing, which would theoretically occur at the system outlet. However, complete mixing would require infinite system size (or residence time). Therefore, in order to evaluate the scale-up of SGP technologies, it is essential to first understand how system-level power density and energy efficiency (and therefore total power output) change with module length (a proxy for system size).

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http://dx.doi.org/10.1016/j.memsci.2014.10.023
0376-7388/

A full-scale process model accounting for changes in velocity, pressure, and concentration along the length of the flow channels can provide a direct, quantitative indication of these important process metrics. To the best of our knowledge, a detailed process model has yet to be published for PRO. For RED, a potentially full-scale process model has already been proposed [10]; however, it was employed for the purpose of comparison with experimental measurements on a small-scale system and was not used to evaluate full-scale performance. Moreover, the model did not consider external mass transfer, which significantly impacts the performance at low crossflow velocity [2]. A recent study by Vermaas et al. applied a process model to evaluate changes in energy recovery and power density over relatively short lengths, but did not account for membrane resistance, viscous dissipation, or concentration polarization [11]. Furthermore, detailed comparisons of PRO and RED previously published have considered the maximum power density achievable, but did not consider how both energy efficiency and average power density would comparatively scale with system size.

The purpose of this paper is to identify how full-scale performance varies with the main process parameters. In particular, this is the first time that PRO performance has been evaluated by a full-scale process model. Note that the two processes modeled are envisioned with different system dimensions and operating principles; hence, directly comparing PRO and RED is naturally a difficult task. Therefore, the results presented herein are designed to illustrate the potential of a full-scale process model and the extension of that model to practical scale-up considerations. The results are not intended to be a prescriptive statement on which process is technologically superior. Furthermore, although PRO and RED can be envisioned as multi-stage processes, we only consider a single stage (single hydro-turbine and single electrode for PRO and RED, respectively). Finally, only co-current cases are considered here; cross-current or counter-current configurations could produce different results.

2. Model formulation

In this section, one-dimensional process models are developed for both PRO and RED, in the spirit of previous efforts to model full-scale reverse osmosis systems [12–15]. In this framework, cross-sectional variations of velocity and concentration in the flow channels are not explicitly solved. Instead, mass transfer correlations are used to account for external concentration polarization (in the case of PRO, internal concentration polarization is accounted for as well), and a friction factor is used to account for viscous dissipation. This results in a significant computational simplification, producing a coupled system of non-linear ordinary, rather than partial, differential equations. Through employing this flexible approach, operational parameters can be more rapidly adjusted and evaluated over a wider range as compared with more computationally intensive approaches such as computational fluid dynamics (CFD). The one-dimensional model is the first step toward approximating scale-up of salinity gradient energy and requires much less computational power than more complex modeling techniques. While over short lengths CFD can be of great use, over longer system lengths (many meters) it is not computationally rational to apply two- and three-dimensional approaches. We note that good correlation between one-dimensional models and experimental data has been reported in the past for the fouling of reverse osmosis systems [16]. In the present formulation, other inefficiencies are included, such as co-ion transport and salt leakage, hydraulic losses within the module, and machine inefficiencies (pumps, hydro-turbine, hydraulic energy recovery device).

Mass and momentum balances applied to each channel yield the system of governing ordinary differential equations for transport in each process. Eqs. (1)–(3) describe changes in crossflow velocity $u(x)$, solute concentration $c(x)$, and pressure $p(x)$, with position x (module length) in the dilute and concentrated channels, *viz.*,

$$\frac{du}{dx} = \pm\frac{2J_w}{h} \tag{1}$$

$$\frac{d(uc)}{dx} = \pm\frac{2J_s}{h} \tag{2}$$

$$\frac{dp}{dx} = -\frac{\lambda\rho u^2}{2d_h} \tag{3}$$

where h is the PRO or RED module channel height, ρ is the density of water, d_h is the hydraulic diameter of the channel, J_w is the water flux across the membrane, J_s is the salt flux, and λ is the friction factor coefficient. An exhaustive study was conducted in order to identify useful friction factor correlations in the literature. Many correlations have been developed for different flow geometries and Reynolds numbers. Here we apply a friction factor correlation which is relevant to the range of conditions modeled in this study [17]. The friction factor coefficients used in determining the hydraulic losses within each channel were calculated through

$$\lambda = \omega_1 - \frac{\omega_2}{Re^{\omega_3}} \tag{4}$$

where ω_1, ω_2, and ω_3 are fitting parameters obtained from computational fluid dynamics [17]. For the case of circular spacers with a diameter half the channel height and 4.5 mm filament spacing, the values of these parameters are $\omega_1 = 0.42$, $\omega_2 = 189.29$, and $\omega_3 = 1$. Please see the Supporting information for more details on our approach to the friction factor correlation.

A graphical representation of a single PRO or RED membrane "cell" is shown in Fig. 1. Note that in Eqs. (1) and (2), negative and positive signs signify mass loss and gain, respectively, occurring in the dilute and concentrated channel. Furthermore, we define the dilution ratio, i.e., the ratio of dilute to concentrated channel flow rates, as

$$d_r = \frac{q_d}{q_c}. \tag{5}$$

The dilution ratio has a fundamental impact on the energy efficiency achieved during mixing, as well as the total power extracted. It is distinctly a feature of large-scale implementation; such effects have not received attention in the laboratory-scale literature.

2.1. Water and salt fluxes

2.1.1. PRO

Water flux in PRO is a function of the bulk solute concentrations (and hence, the osmotic pressure), salt diffusivity D, external mass transfer coefficient k_c, universal gas constant R, temperature T, and membrane permeability to water, A, and salt, B [5]. In order to account for the impact of non-ideality in the osmotic pressure, a correction factor must be applied. The actual osmotic pressure (π_{actual}) and the ideal osmotic pressure (π_{ideal}) can be related through the concentration-dependent osmotic coefficient (ϕ),

$$\phi = \frac{\pi_{actual}}{\pi_{ideal}} = \frac{(RT/\overline{V})\ln a_w}{2RTc} \tag{6}$$

where the numerator is the actual osmotic pressure and considers the activity of water (a_w), and thus non-ideality, and the denominator is the idealized van't Hoff approximation for osmotic