Increasing the power density and reducing the levelized cost of electricity of a reverse electrodialysis stack through blending

Adam M. Weiner\textsuperscript{a}, Ronan K. McGovern\textsuperscript{a}, John H. Lienhard V\textsuperscript{a}

\textsuperscript{a}Center for Clean Water and Clean Energy, Department of Mechanical Engineering
Massachusetts Institute of Technology, Cambridge, MA 02139 USA

Abstract

We increase the power density of a reverse electrodialysis (RED) stack by blending the low salinity feed with a higher salinity stream before the stack entrance. This lowers the capital cost of the system and the resulting levelized cost of electricity, enhancing the viability of RED renewable energy production. Blending increases the power density by decreasing the dominating electrical resistance in the diluate channel as well as the effective resistance caused by concentration polarization, but not without sacrificing some driving potential. To quantify this trade-off and to evaluate the power density improvement blending can provide, a one-dimensional RED stack model is employed, validated with experimental results from the literature. For a typical stack configured with a feed velocity of 1 cm/s, power density improvements of over 20% and levelized cost of energy reductions of over 40% are achievable, provided the salinity of the available river water is below 200 ppm. Additional cost reductions are realized through back-end blending, whereby the diluate exit stream is used as the higher salinity feed. Additionally, improvements from blending increase for higher feed velocities, shorter stack lengths, and larger channel heights.

Keywords: reverse electrodialysis, salinity gradient power, renewable energy, power density

1. Introduction

The objective of this study is to quantify the viability of blending as a design approach for improving the power density of future RED stacks installed at promising locations around the world. An examination of these locations suggests that salinity gradient power production through reverse electrodialysis (RED) could potentially provide 1 TW of clean, uninterrupted power globally [1]. However, the future competitiveness of RED as a renewable energy technology will depend upon the achievement of significant reductions in capital cost, through lower membrane prices and higher power densities [2].

As shown in a recent study of the financial feasibility of reverse electrodialysis [2], capital cost - driven by the gross power density of the stack - is the dominant contributor to the levelized cost of electricity produced by RED. Thus, raising the power density represents the greatest potential for enhancing RED viability.

One effective method for raising the power density is by reducing the electrical resistance through the stack. As evidenced by Fig. 1, significant improvements may be made by reducing the dominant diluate or low conductivity stream resistance $\bar{r}_d$. This reduction may be accomplished in any of the following three ways: by reducing the height of the diluate channel; by reducing the spacer shadow effect; or by increasing the diluate conductivity through blending a high salinity stream with the river water feed [3]. Both reducing the diluate channel height and reducing the spacer shadow effect have been studied extensively in the literature and can significantly improve the power density [4–6]. However, neither method is free from trade-offs. Reducing the diluate channel height increases the effective concentration polarization resistance [6]. Although raising the diluate conductivity through blending has its own set of trade-offs, it significantly reduces both the diluate and effective concentration polarization surface resistances, and should be given careful consideration. With blending, a fundamental trade-off is made between minimizing the diluate resistance and maximizing the driving potential for charge transport by optimizing the diluate salinity. As shown in Fig. 2, by blending a portion of the higher salinity stream with the river water before the RED stack entrance, the salinity of the river water may be increased. Optimization of the amount of blending allows the power density of the stack to be maximized.
Figure 2: One embodiment of blending (front-end blending) involves the continuous withdrawal of pretreated river water and seawater and the continuous blending of a portion of the seawater with the river water before feeding the streams to the stack concentrate and diluate inlets.

Figure 1: Total resistive losses through a typical RED stack include contributions from the concentrate channel $r_c$, anion and cation exchange membranes $r_{AEM}$ and $r_{CEM}$, effect of concentration polarization $r_{CP}$, as well as from the diluate channel $r_d$. Reduction of the diluate channel and effective concentration polarization surface resistances holds the greatest potential for improving RED power density. The case shown assumes a diluate feed from the Rhone River (339 ppm salinity \cite{7}), a channel height of 100 $\mu$m, stack length of 10 cm, and feed velocity of 1 cm/s. Here, each contributing surface resistance is averaged over the RED length.

Currently, there are many studies in the literature illustrating the trade-off between diluate resistance and driving potential, but there are no studies devoted to quantifying the power density gains made through blending. Weinstein and Leitz \cite{8} and Lacey \cite{9} modeled and computed the optimal diluate concentration in a zero-dimensional RED stack (a stack of infinitesimal length) with a seawater concentrate stream. Similarly, Veerman et al. \cite{10} showed that the local power density in a one-dimensional stack (a stack of finite length) initially increases and reaches a maximum with respect to increasing local diluate concentration within the stack. The same trade-off is found in using electrodialysis for desalination. McGovern et al. \cite{11, 12} showed that removing salt from higher diluate salinity feeds significantly reduces the capital cost, because the resistance is lower.

Other studies have examined different feed waters for salinity gradient power generation, but none have considered blending the feed water as a design approach \cite{2, 13, 14}. As an example, Daniilidis et al. \cite{2} experimentally investigated RED power density and efficiency across a wide range of feed water salinities beyond river and seawater applications showing that power density continues to increase at very large salinity gradients despite reductions in permselectivity.

In a 2009 study of the various power output limitations in an RED stack, Dlugolecki et al. \cite{3} mentioned blending as a possible means for reducing the resistance of the diluate channel, without specifically quantifying the potential power density improvements. The study cites early RED work by Weinstein and Leitz \cite{8} from whose results it may be inferred that blending 600 ppm river water with seawater in a stack of infinitesimal length \footnote{An infinitesimal stack model does not account for streamwise variations in concentration as the diluate and concentrate travel through the RED channels. The present results are based on a one-dimensional model that incorporates the strong influence of streamwise changes.} could improve RED power density by upwards of 30%. Since Weinstein and Leitz’s zero-dimensional study, significant advancements have been made in improving RED technology and stack design and in understanding and modeling the loss mechanisms as well. Absent from the current literature is a quantification of blending improvements since this progress.

In our study, we analyze the viability of blending...
in the context of the current modeling methods. First, we analyze blending using a one-dimensional model which accounts for streamwise variations in salinity. In Sect. 3.2, we show that, when inflowing river water is at 600 ppm, blending results in a minimal power density gain. Second, we analyze blending using a model which includes concentration polarization effects. We show in Sect. 4.1 that blending also reduces the effective concentration polarization resistance.

We also analyze the viability of blending in the context of recent designs. As an example, RED membrane resistances have decreased by over 90% since early RED development and no longer dominate resistive losses (see Fig. 1). Additionally, channel heights have decreased ten-fold. We then extend the study by quantifying how blending may impact future stack configurations (see Sect. 4.3), concluding that improvements increase with shorter residence times and larger channel heights.

Lastly, we propose and analyze a blending configuration in which the diluate feed is recycled, see Fig. 3. In this configuration (termed back-end blending with diluate recirculation), recycling of the diluate feed reduces pretreatment system capital, operating, and energy costs - an additional benefit. We briefly analyze and discuss the cost advantages over front-end blending in Sect. 3.3.

Figure 3: Back-end blending with diluate recirculation. The advantages are reductions in pretreatment capital, operating, and energy costs.

2. Methods

To quantify the gross power density improvements achieved through blending, we model a single-cell RED stack accounting for salinity variations in the streamwise direction. We model three stack designs proposed in the literature, each characterized by different feed velocities - 0.5 cm/s [15], 1 cm/s [4], and 1.25 cm/s [16]. All three designs have 100 µm channel heights and 10 cm stack lengths. For each design, we maximize the gross power density with respect to the load resistance and inlet diluate salinity. We then compare this power density to the power density achieved with unblended river water (with the load resistance optimized) to evaluate the gross power density improvement.

To quantify the cost advantages of back-end blending over front-end blending, we model the RED net power density and levelized cost of electricity for the front-end and back-end blending cases, noting that for fixed velocities, the pumping power will be the same. As in the gross power density case, we repeat this analysis for three different feed velocities, keeping the stack length fixed.

2.1. Gross power density model

The RED gross power density model is based on the approach taken by Weiner et al. [15]. As in [15], we fix the diluate and concentrate channel heights \( h \) and set them equal, as well as set the diluate and concentrate inlet feed velocities \( V_i \) equal. This simplifies the stack design. Figure 4 shows the cell pair circuit diagram, which is divided into \( N \) discrete segments for modeling stream-wise variations in electromotive force (EMF) \( \epsilon_n \), diluate resistance \( R_{d,n} \), and concentrate resistance \( R_{c,n} \). The variations result from changes in salinity along the length due to salt and water transport across the membranes. For this analysis, we model 20 stack segments (\( N = 20 \)). Additionally, we neglect variations in membrane resistance as well as the existence of ionic shortcut currents [17].

![Circuit diagram](image)

Figure 4: A circuit model for the one-dimensional, unsegmented-electrode RED stack which accounts for streamwise variations in salinity.

The local EMFs \( \epsilon_n \) are computed from the local chemical potential differences across the membranes [18]:

\[
\epsilon_n = \frac{t_s}{F} \left( \mu_{c,n}^\circ - \mu_{d,n}^\circ \right) + \frac{t_w}{F} \left( \mu_{c,n}^w - \mu_{d,n}^w \right)
\]

(1)

where \( t_s \) is the salt transport number, \( t_w \) is the water transport number, \( F \) is Faraday’s constant, \( \mu_{c,d}^\circ \) is the local salt chemical potential at the membrane surface
on the concentrate side, and \( \mu_{d,n} \) is the local salt chemical potential at the membrane surface on the dilute side. Blending reduces the difference between \( \mu_{d,n} \) and \( \mu_{d,n} \), driving the local EMFs down. This detrimental effect is a trade-off against reduced stack resistance brought about by the increased dilute salinity.

We model concentration polarization using a convection-diffusion approach [19] (dilute example shown):

\[
C_{d,m,n} - C_{d,n} \approx \frac{2h}{\text{Sh}_{d,n}} \frac{j_{D,n}}{F} \frac{(\bar{T}_c - t_{cu})}{D_{\text{NaCl}}} \tag{2}
\]

where \( C_{d,m,n} \) is the local diluate concentration at the membrane (an input to \( \mu_{d,n} \)), \( C_{d,n} \) is the local diluate concentration in the bulk, \( j_{D,n} \) is the local current density (see Eq. 8 below), \( h \) is the channel height, \( t_{cu} \) is the counter-ion transport number (\( \approx 0.5 \) for anions and cations), and \( D_{\text{NaCl}} \) is the diffusion coefficient of salt through the bulk. \( T_{cu} \) is the integral counter-ion transport number [20]:

\[
\bar{T}_c \approx t_{cu} + \frac{1}{2}
\tag{3}
\]

\( \text{Sh}_{d,n} \) is the local Sherwood number [21] (dilute example shown):

\[
\text{Sh}_{d,n} = K_w \text{Re}_{D_{d,n}}^{1/2} \text{Sc}_{d,n}^{1/3} \tag{4}
\]

where \( K_w \) is the Kuroda constant, \( \text{Re}_{D_{d,n}} \) is the local Reynolds number based on the hydraulic diameter \( D_h \), and \( \text{Sc}_{d,n} \) is the local Schmidt number.

When the difference in concentration between the bulk and membrane is small compared to the diluate concentration, one can approximate concentration polarization as a linearized ohmic surface resistance (see Appendix B):

\[
\bar{r}_{\text{CP,n}} = 2t_{cu} hRT \frac{(\bar{T}_c - t_{cu})}{F^2 \text{Sh}_{d,n} D_{\text{NaCl}}} \frac{1}{C_{d,c} + \frac{1}{C_{d,n}}} \tag{5}
\]

Equation 5 reveals that increased blending (and by extension, increased \( C_{d,m,n} \)) drives this effective concentration polarization resistance lower.

The other resistance sources are combined in the total ohmic surface resistance \( \bar{r}_{\text{tot,n}} \) [16]:

\[
\bar{r}_{\text{tot,n}} = \bar{r}_{\text{AEM}} + \bar{r}_{\text{CEM}} + \bar{r}_{d,n} + \bar{r}_{e,n} \tag{6}
\]

Membrane resistances \( \bar{r}_{\text{AEM}} \) and \( \bar{r}_{\text{CEM}} \) generally decrease with increasing salinity (and therefore blending) [5], but they represent a small fraction of the total stack resistance. Consequently, we neglect any such variations.

The main benefit of blending is a reduction in the diluate resistance. The local diluate channel resistance \( \bar{r}_{d,n} \) is modeled as [16]:

\[
\bar{r}_{d,n} = \frac{h}{\epsilon^2 \kappa_{d,n}} \tag{7}
\]

where \( \epsilon \) is the spacer porosity and \( \kappa_{d,n} \) is the diluate conductivity - the product of diluate concentration \( C_{d,n} \) and solution molar conductance \( \Lambda_{d,n} \) [22]. The increase in conductivity with \( C_{d,n} \) is slightly less than linear, as solution molar conductance \( \Lambda_{d,n} \) decreases with concentration (see Fig. 5).

![Dilute conductivity](image)

Figure 5: Dilute conductivity \( \kappa_d \) as a function of concentration, \( c_d \), or salinity, \( S_d \).

The local concentrate channel resistance \( \bar{r}_{e,n} \) is modeled analogously to the diluate resistance although the impact of blending on this resistance is negligible.

Combining the local EMFs and resistance sources, provides the local current density:

\[
j_{D,n} = \frac{\epsilon_n - \phi_{\text{stack}}}{\bar{r}_{\text{tot,n}}} \tag{8}
\]

where \( \phi_{\text{stack}} \) is the total stack voltage. The total stack voltage is derived from Kirchoff’s Current Law:

\[
\phi_{\text{stack}} = \sum \frac{E_{\text{seg}}}{R_{\text{seg}}} \cdot \frac{A_n}{A_L} \tag{9}
\]

where \( A_n \) is the area of a section and \( R_L \) is the load resistance.

The local molar salt and water fluxes \( J_{s,n} \) and \( J_{w,n} \) transported into the diluate channel are modeled as in Fidalgo and Moresi [23]:

\[
J_{s,n} = \frac{j_{D,n}}{F} + L_s \left( C_{d,m,n} - C_{d,n} \right) \tag{10}
\]

\[
J_{w,n} = \frac{j_{D,n}}{F} - L_w \left( \pi_{m,n} - \pi_{d,n} \right) \tag{11}
\]

where \( L_s \) is the overall salt permeability (in m/s), \( L_w \) is the overall water permeability (in mol/(bar-m²-s)), and \( \pi_{m,n} \) is the local osmotic pressure at the membrane surface [24]. Finally, the gross power density \( P_{D,g} \) is given by:
2.2. Net power density model

The net power density \( P_{D,\text{net}} \) is modeled as the gross power density less the stack pumping power density \( P_{D,s} \) and the pumping power required through the pretreatment system \( P_{D,\text{PT}} \) [15]. The stack pumping power density is modeled as:

\[
P_{D,s} = \frac{\phi_{\text{stack}}^2}{R_{L,\text{wl}}}
\]  

(12)

where \( \phi \) is the stack width and \( l \) is the stack length.

In our analysis, the load resistance \( R_L \) is continuously optimized with respect to the gross power density [15].

2.2. Net power density model

The net power density \( P_{D,\text{net}} \) is modeled as:

\[
P_{D,\text{net}} = P_{D,s} - P_{D,\text{PT}} = \frac{K_p \mu V^2}{h} - \frac{K_p h h \mu V^2}{h}
\]  

(13)

where \( K_p \) is a fitted parameter (accounting for the diluate and concentrate feeds) and \( \mu \) is the feed viscosity. The stack pumping power model was fit to experimental data reported by Vermaas et al. [4]. The pretreatment system is modeled after a setup implemented by Post et al. [1], featuring coarse-media filtration with two drum filters. The required pumping power density through the system \( P_{D,\text{PT}} \) is modeled as:

\[
P_{D,\text{PT}} = \left[ \rho_f (1 - X_{\text{blend}}) + \rho_c \right] \frac{g h H h}{\tau}
\]  

(14)

where \( \rho \) is the feed density, \( X_{\text{blend}} \) is the mass fraction of exiting diluate diverted for blending, \( g \) is the gravitational body acceleration, \( H \) is the system head loss, and \( \tau \) is the residence time (stack length divided by feed velocity). Equation 14 shows how back-end blending reduces the pumping power density required for pretreatment.

2.3. Modified cost model

When back-end blending as opposed to front-end blending is implemented, additional cost reductions are realized from recycling a portion of the already-pretreated diluate stream. To assess the impact of back-end blending on the levelized cost of electricity, we model the LCOE as in Weiner et al. [15], with a slight modification to account for this recirculation:

\[
\text{LCOE} = \frac{1}{P_{D,\text{net}}} \left[ \frac{K_{\text{mem}}}{\text{CAF}} + (2 - X_{\text{blend}}) \frac{K_{\text{PT}} h h}{\tau \text{CAF}} \right]
\]  

(15)

where \( K_{\text{mem}} \) is the RED stack capital cost figure in $/m^2$, \( K_{\text{PT}} \) is the pretreatment system capital cost figure in $/(m^2/day)$, and CAF is the capital amortization factor [20]:

\[
\text{CAF} = \frac{1}{r} \left[ 1 - \left( \frac{1}{1 + r} \right)^T \right]
\]  

(16)

We assume a plant life \( \Gamma \) of 20 periods (one period is one year) and an annualized cost of capital \( r \) of 6% [1].

The constants used throughout the model are compiled in Appendix D. To determine the salt and water permeabilities, salt transport number, and Kuroda constant, the model was fit to experimental data [4], as in Appendix A. All equations were solved using the quadratic approximation method in Engineering Equation Solver [25].

3. Results

3.1. The optimal inlet diluate salinity

Figure 6 shows the computed inlet diluate salinity which maximizes the gross power density of an RED stack for three RED systems proposed in the literature.
3.2. Power density improvements through blending

Figure 7 quantifies how the gross power density improvement varies with the available river water salinity up to 600 ppm. For the Rhone River (339 ppm), blending can improve the gross power density of the stack (with a feed velocity of 1 cm/s) by over 9%. As shown in Fig. 6, the optimal power density is 2.06 W/m².

The gross power density improvement increases significantly at low river water salinities, primarily because the diluate resistance and effective concentration polarization resistance are proportional to the inverse of the diluate concentration (see Eqs. 5 and 7). The diluate resistance is extremely high at low salinities.

Figure 8 shows the resulting percent reduction in the levelized cost of electricity by simply employing front-end blending. In the Rhone River case, cost reductions of over 21% are achievable. Because the levelized cost of electricity is dominated by the stack capital cost, the expected capital cost reductions are nearly equivalent to the capital cost reductions in LCOE shown here. Despite these sizable cost reductions, electricity production with reverse electrodialysis remains significantly more costly than production with other renewable energy technologies.

For each feed velocity, there is a small but finite river water salinity at which, without blending, the gross power density is too low for the system to produce a positive net power density and the levelized cost of electricity approaches infinity. Consequently, there is a sharp rise in cost reductions brought about by blending for low salinity river water feeds. At higher salinity river water feeds, the cost reductions from blending diminish rapidly as gross power density improvements diminish.

3.3. Back-end vs. front-end blending

Figure 9 compares the total percent reduction in the levelized cost of electricity by employing back-end blending to the total percent reduction in cost by employing back-end blending with diluate recirculation. For all feed velocities and across all river water salinities, back-end blending offers additional cost reductions. These reductions are primarily due to the drop in the required pretreatment pumping power density which increases the net power density (see Eq. 14). Pretreatment capital and operating costs drop as well.

At low river water salinities, back-end improvements approach front-end improvements as the net power density approaches zero and reductions using either a back-end or front-end configuration approach 100%. At river water salinities approaching the optimal inlet diluate salinity, neither form of blending offers significant cost reductions. In general, back-end blending offers higher additional cost reductions with increasing feed velocity, because the difference between the outlet and inlet diluate salinities decreases with increasing velocity and a larger recirculation mass fraction, $X_{\text{blend}}$, is required. See Appendix C for plots of the required blending mass fraction $X_{\text{blend}}$ in both the front-end and back-end blending cases.
4. Discussion

4.1. Effect of blending on average resistances and average resistive losses

To illustrate the impact blending has on surface resistance contributions within the stack, we consider the example case of an RED stack installed on the Rhone River (339 ppm). As shown in Fig. 10 below, blending reduces the total stack surface resistance $r_{tot}$ by about 46%. This arises from reductions in the average dilute channel resistance $r_{d,avg}$ (by 53%) and the average effective concentration polarization resistance $r_{CP,avg}$ (by about 53%). The average concentrate channel resistance $r_{c,avg}$ remains relatively constant (membrane resistances, $r_{AEM}$ and $r_{CEM}$, are held constant at $1 \, \Omega \, \text{cm}^2$). Surface resistance profiles are averaged over the stack length.

![Figure 10: A comparison of the total stack surface resistance $r_{tot}$ and the major resistance sources which comprise it, before and after blending](image)

4.2. Blending improvements at promising estuaries around the world

In Table 1, we evaluate the power density improvements gained through blending for RED stacks installed at promising river mouth systems throughout the world. Locations with low salt intrusion lengths [13] and mild climates where reduced sedimentation and biological activity drive pretreatment costs [26] down are most promising. Additionally, rivers with estuaries in close proximity to large population centers and in regions with favorable renewable energy policies show the most promise.

Data on river water salinities is taken from Gaillardet et al. [7] who references the GEMS/WATER Global Register of River Inputs. The Register compiles dissolved ion measurements from various sources. Each measurement is collected near the river mouth and upstream of any ocean influence and is averaged over at least five years [27]. As we are using TDS for salinity, we do not account for the effect of multivalent ions on RED power density [28]. Additionally, we assume an available seawater salinity of 35,000 ppm, although ocean salinities along coastlines may range between 30,000 and 40,000 ppm [29].

<table>
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<th>$P_{D,E}$ ↑</th>
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<tr>
<td>Rhone</td>
<td>339 ppm</td>
<td>9%</td>
<td>30%</td>
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</table>

Table 1: Percent power density improvements and levelized cost of electricity reductions through back-end blending at promising locations around the world, assuming an available seawater salinity of 35,000 ppm and a feed velocity of 1 cm/s

4.3. Impact on future designs

The power density improvements that can be gained through blending are sensitive to design parameters such as feed velocity, stack length, and channel height. Considering that RED stack designs may change with time, we evaluate how power density improvements change with respect to these design parameters.

![Figure 11: The sensitivity of blending improvements to the stack length over a wide river water salinity range](image)
Figure 12 shows increasing improvements from blending with increasing channel height. For the Rhone River case (339 ppm river water), increasing the channel height from 100 μm to 200 μm increases blending improvements by about 10 percentage points. Improvements are most sensitive to channel height, because the diluate resistance scales with channel height directly (see Eq. 7). The 100 μm channel height represents the optimal channel height with respect to the net power density identified in the literature (excluding pretreatment energy and capital costs) [4].

5. Conclusions

Our analysis of the optimal inlet diluate salinity for current RED stack designs shows that for nearly all river mouth systems deemed suitable for salinity gradient power, blending significantly improves the gross power density and reduces the levelized cost of electricity. In the case of the RED stack sourcing river water from the Rhone (339 ppm and 1 cm/s feed velocity), a 9% increase in gross power density and 21% reduction in the levelized cost of electricity are achievable through front-end blending.

Further cost reductions may be realized by implementing back-end blending with diluate recirculation. In the case of the Rhone, a 30% total reduction in the levelized cost of electricity may be realized. Considering that stack designs may evolve with time, we predict that configurations with higher feed velocities, smaller stack lengths, and larger channel heights will benefit more from blending.

6. Acknowledgements

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7. References

Nomenclature

**Roman Symbols**

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<th>Definition</th>
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**Greek Symbols**

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**Subscripts**

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</tr>
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</tr>
<tr>
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<tr>
<td>i</td>
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<tr>
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<td>load</td>
</tr>
<tr>
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</tr>
<tr>
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<tr>
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</tr>
<tr>
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<td>net</td>
</tr>
<tr>
<td>PT</td>
<td>pretreatment</td>
</tr>
<tr>
<td>p</td>
<td>pressure</td>
</tr>
<tr>
<td>r</td>
<td>river water</td>
</tr>
<tr>
<td>s</td>
<td>stack</td>
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</tbody>
</table>

**Nomenclature**

- **LCOE**: levelized cost of electricity, $/kWh
- **P**: power density, W/m²
- **ρ**: pressure, Pa
- **R**: universal gas constant, J/mol K
- **R**: resistance, Ω
- **r**: annuity depreciation rate
- **r**: membrane surface resistance, Ω m²
- **Re**: Reynolds number
- **S**: salinity, ppm
- **Sc**: Schmidt number
- **Sh**: Sherwood number
- **T**: temperature, K
- **t**: transport number
- **V**: feed velocity, m/s
- **w**: stack width, m
- **X**: mass-based blending fraction

**Subscripts**

- **aq**: aqueous
- **avg**: average
- **blend**: blended
- **c**: concentrate or critical
- **cu**: counter-ion
- **CEM**: cation exchange membrane
- **CP**: concentration polarization
- **D**: density
- **D_h**: hydraulic diameter
- **d**: dilute
- **eq**: equivalent
- **g**: gross
- **i**: inlet
- **L**: load
- **m**: membrane surface
- **n**: segment number
- **NaCl**: salt solution
- **net**: net
- **PT**: pretreatment
- **p**: pressure
- **r**: river water
- **s**: stack
Appendix A. Model validation

We validate the model as in Weiner et al. [15], with experimental results reported by Vermaas et al. [4]. The salt transport number \( t_s \), spacer porosity \( \varepsilon \), and Sherwood correlation constant \( K_m \) are the fit parameters (see Table D.2). In Fig. A.1, we show the fit results for the average ohmic surface resistance \( \bar{r}_{ohm} \), which includes the sum of the average diluate surface resistance \( \bar{r}_d \), concentrate surface resistance \( \bar{r}_c \), and membrane resistances \( \bar{r}_{AEM} \) and \( \bar{r}_{CEM} \).

![Figure A.1: Validation of the model with respect to the average ohmic surface resistance at various flow rates. The root mean squared error in the fit is 0.62 \( \Omega \) cm\(^2\).](image)

In Fig. A.2, we show the fit results for the equivalent stack surface resistance \( \bar{r}_{eq} \). The equivalent stack surface resistance includes the sum of the average ohmic surface resistance \( \bar{r}_{ohm} \), the average effective concentration polarization resistance \( \bar{r}_{CP} \), and an effective resistance caused by streamwise concentration variations \( \bar{r}_{C} \) [4]. The equivalent stack surface resistance is computed using the following equation:

\[
\phi_{\text{stack}} = \epsilon_i^* - \bar{r}_{eq} j D \text{tot} \quad (A.1)
\]

where \( \epsilon_i^* \) is the local EMF at the inlet without considering concentration polarization losses. The total current density is determined from load resistance matching. The load resistance \( R_L \) is set equal to \( \bar{r}_{eq}/wL \).

In Fig. A.3, we show the fit results for the gross power density \( P_{D,g} \), and in Fig. A.4 we show the fit results for the stack pressure drop. The constant \( K_p \) in the pumping power model was fit to the experimental results reported by Vermaas et al. [4].

Appendix B. Derivation of the linearized, effective concentration polarization surface resistance

Concentration polarization reduces the local EMFs in RED by increasing the salt concentration at the
membrane wall in the diluate channel, while decreasing the concentration at the membrane wall on the concentrate side. The reduction in local EMF may be linearized and rewritten as an effective local ohmic resistance by first expanding out the local EMF (Eq. 1) using the Nernst equation:

$$\varepsilon_n = I_j \frac{RT}{F} \ln \frac{\gamma_c C_{r,n}}{\gamma_d C_{d,n}}$$

where $T$ is the temperature, and $\gamma_c$ and $\gamma_d$ are the activity coefficients of the concentrate and diluate (as modeled by Pitzer [30]). Assuming the concentration difference between the membrane and bulk $|\Delta C_n|$ is the same in either channel, the local EMF may be rewritten as:

$$\varepsilon_n = I_j \frac{RT}{F} \ln \frac{\gamma_c C_{r,n}}{\gamma_d C_{d,n}} (1 - |\Delta C_n|/C_{r,n})$$

Equation B.2 may be separated and Taylor expanded, and if $|\Delta C_n|/C_{d,n}$ is small (which generally holds), may be simplified into the sum of a regular Nernst potential and a concentration polarization potential loss term:

$$\varepsilon_n = I_j \frac{RT}{F} \ln \frac{\gamma_c C_{r,n}}{\gamma_d C_{d,n}} \Delta C_n \left( \frac{1}{C_{r,n}} + \frac{1}{C_{d,n}} \right)$$

where the concentration change $\Delta C_n$ is given by Eq. 2. Substituting Eq. 2 into Eq. B.3, reveals a concentration polarization potential loss which is linear with local current density $j_{d,n}$. The effective concentration polarization surface resistance may therefore be defined as in Eq. 5:

$$\overline{R_{CP,n}} = \frac{2 I_j hRT (T_{sw} - T_{cw})}{F^2 S \eta_{d,n}} \left( \frac{1}{C_{r,n}} + \frac{1}{C_{d,n}} \right)$$

Appendix C. Plots of the required blending mass fraction $X_{blend}$

We model front-end blending in the following way, considering a salt and mass balance:

$$C_r \phi_i + X_{blend} C_{d,i} \phi_{sw} = C_{d,i} \phi_{d,i}$$

$$\rho_i \phi_i + X_{blend} \rho_{d,i} \phi_{sw} = \rho_{d,i} \phi_{d,i}$$

where $C_r$ is the river water concentration, $\phi_i$ is the river water flow rate, $C_{d,i}$ is the concentrate concentration at the stack inlet, $\phi_{sw}$ is the seawater flow rate, $C_{d,i}$ is the diluate concentration at the stack inlet, $\phi_{d,i}$ is the diluate flow rate at the stack inlet, $\rho_{c,i}$ is the concentrate density at the stack inlet, and $\rho_{d,i}$ is the diluate density at the stack inlet.

Figure C.5 shows the required mass fraction $X_{blend}$ of seawater to be blended with river water in blending to the optimal inlet diluate salinity. In the Rhone River case with a feed velocity of 1 cm/s, $X_{blend}$ is 0.02.

The blending fractions decrease nearly linearly with river water salinity as the river salinity approaches the optimal diluate salinity. Blending fractions increase with velocity as the optimal inlet diluate salinity also increases with velocity.
We model back-end blending with diluate recirculation in the following way, considering a salt and mass balance:

\[ C_i \phi_i + X_{\text{blend}} C_{d,N} \phi_{d,N} = C_{d,i} \phi_{d,i} \]  \hspace{1cm} (C.3)

\[ \rho_i \phi_i + X_{\text{blend}} \rho_{d,N} \phi_{d,N} = \rho_{d,i} \phi_{d,i} \]  \hspace{1cm} (C.4)

where \( C_{d,N} \) is the diluate concentration at the stack exit, \( \phi_{d,N} \) is the diluate flow rate at the stack exit, and \( \rho_{d,N} \) is the diluate density at the stack exit.

Figure C.6 shows the required mass fraction \( X_{\text{blend}} \) of the exiting diluate stream to be blended with river water in blending to the optimal inlet diluate salinity. In the Rhone River case with a feed velocity of 1 cm/s, \( X_{\text{blend}} \) is 0.26.

**Appendix D. Summary of the input model parameters**

A summary of the model parameters and equations is provided in Table D.2

Table D.2: Membrane, solution, channel and flow, as well as costing parameters and properties used in the analysis

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<th>Symbol</th>
<th>Value</th>
<th>Ref.</th>
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<td>[20]</td>
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<td>[31]</td>
</tr>
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<tr>
<td>( r_{CEM} )</td>
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<tr>
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</table>

\( ^a \) Represents the average of the measured head losses by Post et al. [1] in the summer, winter, and spring

\( ^b \) Includes estimated operating (chemical) costs associated with pretreatment