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Impact of salt retention on true batch reverse osmosis energy consumption: experiments and model validation

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Abstract: In theory, the batch reverse osmosis (RO) process achieves the lowest practical energy consumption by varying pressure over time. However, few batch RO systems have been built and operated. We have designed, built, and operated the first “true” batch RO prototype using a flexible bladder. The flexible bladder serves as the high-pressure variable-volume tank that is inherent to true batch RO designs (as opposed to batch RO with energy recovery devices). We experimentally validated a model of batch RO energy consumption (≤ 2.7% difference) by measuring the hydraulic work of the high pressure and circulation pumps. We find that batch RO energy consumption will be greater than expected mostly due to salt retention, a problem neglected by most previous studies. However, despite operating at elevated salinity and flux conditions, batch RO can still save energy relative to single-stage and multi-stage continuous systems. For a seawater desalination plant (35 g/kg intake, 50% recovery, 15 L m\textsuperscript{−2} h\textsuperscript{−1}), our newly-validated model predicts that batch RO would save 11% energy compared to a single-stage continuous RO plant. Our work demonstrates that batch RO is an energy-efficient process with the potential to reduce the cost of water desalination.

Keywords: desalination, reverse osmosis, batch reverse osmosis, salt retention, energy efficiency

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Nomenclature

Acronyms

BWRO  Brackish water reverse osmosis
CP    Concentration polarization factor
ERD   Energy recovery device
HP    High-pressure pump
RO    Reverse osmosis
SEC   Specific energy consumption kWh m$^{-3}$
SWRO  Seawater reverse osmosis

Greek symbols

$\pi$ Osmotic pressure bar
$\Theta$ Dimensionless feed salinity elevation

Roman symbols

$\bar{J}$ Average permeate flux L m$^{-2}$ h$^{-1}$
$A$ Membrane water permeability L m$^{-2}$ h$^{-1}$ bar$^{-1}$
$A_m$ Membrane area m$^2$
$C_F$ Flushing effectiveness coefficient
$J$ Permeate flux L m$^{-2}$ h$^{-1}$
$P$ Hydraulic pressure bar
$Q$ Flow rate m$^3$ s$^{-1}$
$RR$ Recovery ratio (by volume)
$t$ Time s
$t^*$ Dimensionless flush time
$t_{tr}$ Dimensionless transition flush time
$V_p$ Permeate volume m$^3$
$w$ Salt mass fraction g kg$^{-1}$
$W_{least}$ Least work of separation kWh

Subscripts

b Brine
circ Circulation pump
cyc Batch cycle
exp Experiment
f Feed
fl Flush phase
pl Plant conditions
plug Plug flow conditions
1. Motivation

An estimated four billion people face extreme water scarcity for at least one month of the year [1]. In the United States, 40 out of 50 state water managers expected to see water shortages from 2013 to 2024 [2]. As such, it is vitally important to find cheaper and more sustainable methods of water recovery. Although reverse osmosis (RO) desalination was relatively expensive when it was first introduced in the 1970s, its energy consumption has reduced dramatically due to improvements in membrane permeability, pump efficiency, and energy recovery [3]. However, the energy consumption of RO plants is still significant, leading researchers to explore different system configurations in an effort to reduce costs. Batch RO is one such configuration, but its practical limitations have not yet been fully explored. This paper details experimental findings of a true batch RO prototype and revisits the comparison of batch and conventional RO systems’ energetic performances.

1.1. Energy consumption in reverse osmosis processes

In the conventional RO process (continuous RO), a pressurized and salty feed stream flows through the RO system continuously (Fig. 1A). The RO membrane separates the feed into two streams: pure permeate and concentrated brine, which constantly flow out of the system. The feed passes only once through the system (which may have multiple stages), and the permeate is recovered at the desired rate.

In the batch RO process (Fig. 1B), the feed/retentate stream is recirculated through the system multiple times before the desired amount of permeate is recovered. In this time-variant process, the feed pressure can vary over time as the average system concentration increases.

A. Continuous RO

B. Batch RO

Several authors [4, 5, 6] frame batch RO as the ideal desalination process where feed pressure precisely matches the osmotic pressure. The specific energy consumption of this ideal batch RO process, $SEC_{batch, ideal}$,
is equivalent to the least work of separation (normalized by product volume) for a given feed osmotic pressure \( \pi_f \) and recovery ratio \( RR \) [4]:

\[
SEC_{\text{batch,ideal}} = \frac{W_{\text{least}}}{V_p} = \frac{\pi_f}{RR} \ln \left[ \frac{1}{1 - RR} \right]
\]

where \( W_{\text{least}} \) is the least work of separation for the specified desalination process and \( V_p \) is the volume of permeate obtained from the process. The equation above assumes a linear relationship between osmotic pressure and concentration, a uniform concentration throughout the system, and zero losses due to mass transfer resistance, mixing, friction, or component inefficiency. Actual batch RO systems will consume more energy than this due to practical realities such as spatial variations in concentration and a finite net driving pressure [4].

Energy consumption in cross-flow RO processes is nicely illustrated by a pressure-recovery diagram, as introduced by Liu et al. [4]. In this simplified model, the total area under the feed pressure curve corresponds to the energy consumed during each process, because instantaneous recovery ratio is proportional to the permeate volume. The batch RO process requires much less energy than continuous RO at the same recovery, as shown in Figure 2.

In a continuous RO system, the system osmotic pressure profile is steady in time and all variations occur in space. The applied feed pressure must overcome the brine osmotic pressure, which is larger than the feed osmotic pressure. A continuous RO system is unbalanced: the permeate flux at the front of the system is multiple times greater than at the back of the system. A more balanced system could produce the same amount of fresh water using less energy, such as in multi-stage RO systems [7, 8]. However, no process can consume less energy than the thermodynamic least work of separation, represented by the area under the osmotic pressure curve.

The batch RO process achieves the lowest practical energy consumption by varying feed pressure over time. The high-pressure pump can apply a low pressure at the beginning of the batch RO process and gradually increase the pressure in accordance with the salinity of the system. This process is balanced - the permeate flux is kept constant in time. In the ideal batch RO process the permeate flux would also be uniform over space, but this is not achievable in practice. Still, significant energy may be saved if the osmotic pressure (and thus flux) varies minimally along the length of a short membrane module.

1.2. Batch RO: background

A conceptual drawing of the batch RO process illustrates the essential characteristics of a batch RO system (Fig. 3). A constant net driving pressure is applied to produce a constant permeate flux. A high-pressure, variable-volume tank is required to accommodate the shrinking volume of salt water. A stirrer reduces the effects of concentration polarization.

Figure 2: Pressure-recovery diagram for continuous and batch RO at 50% recovery. The batch RO process achieves the lowest practical energy consumption by varying feed pressure to follow the osmotic pressure curve. Each process produces the same amount of water. The area under the feed pressure lines represent the majority of work done by the high-pressure pump (this diagram does not show the work required to overcome concentration polarization or frictional losses in either process).
Net Driving Pressure = $\Delta P - \Delta \pi$
(constant)

Piston
RO membrane
Salt water
Stirrer
Permeate flux
(constant)

High-pressure, Variable-volume tank

Figure 3: This drawing illustrates the essential characteristics of a true batch RO system.

Circulation pump

A circulation pump that produces crossflow serves as the stirrer in practical batch RO systems [5, 9]. Without the circulation pump, higher pressures would be required to drive the permeate flow due to salt accumulation at the membrane surface (i.e., concentration polarization). This addition is a significant difference from continuous RO systems - the circulation pump in a batch RO system presents an additional degree of freedom when it comes to system operation but also requires additional energy [10, 11], cutting into the potential energy savings of batch systems.

The use of a circulation pump in batch RO allows for flexibility in operating at different crossflow velocities and therefore different levels of concentration polarization [9]. In continuous RO, the maximum and minimum crossflow velocities are determined by the initial feed flow and the overall recovery ratio. In a batch RO system, the maximum crossflow velocity is determined by the circulation pump flow rate, which can be adjusted independent of the initial feed mass and applied pressure. The minimum crossflow velocity occurs at the end of the membrane module, but should be close to the maximum crossflow velocity in batch RO systems since the per-pass recovery ratio is relatively low.

True batch RO

The key challenge to implementing batch RO is the decreasing volume of pressurized saline water that remains in the system as permeate is produced (Fig. 3). Prior to the start of each cycle, a batch of feed is introduced into the system. During the batch RO process, permeate leaves the system through the membrane and the retentate circulates in a loop around the system. This retentate decreases in volume throughout the process since no new feed enters the system during permeate production. The retentate must stay pressurized to produce permeate, so a true batch RO system (also called “ideal” or “fully-pressurized” batch RO [10, 11]) requires a feed tank that can operate at high pressure while changing in volume.

One proposed batch RO design avoids the need for a variable-volume tank by using an energy recovery device (ERD) to allow the feed water to be stored in a tank at atmospheric pressure [10, 11]. The ERD is used to depressurize the retentate and recycle that energy to pressurize feed water before it enters the membrane module. To our knowledge, batch RO with ERDs has not yet been implemented. However, theoretical studies have shown that batch RO with ERDs consumes more energy than true batch RO [10, 11]. Significant energy is lost as the feed repeatedly cycles through the ERDs, which are not perfectly efficient [12].

One way to implement true batch RO is with a rigid piston, which divides a pressure vessel into two compartments [13]. One compartment is filled with the feed that is to be treated during the batch RO process, while the other compartment starts out empty. A pump introduces make-up fluid (e.g., water) into the empty compartment in order to trigger permeate production. The make-up fluid accommodates the shrinking feed volume and does not interact directly with the membrane since it is separated from the feed by the piston. Since the pressure vessel is already completely filled, the pressure in both compartments will rise very rapidly due to the near incompressibility of water. Once the pressure exceeds the feed osmotic pressure, permeate will leave through the membrane, the feed volume will decrease, and the piston will move accordingly. In this manner, the flowrate of make-up fluid is approximately equal to the permeate flowrate.
Another proposed true batch design uses a flexible bladder to divide the pressure vessel into two compartments [14]. In this work, we have built and operated the first true batch RO prototype using a flexible bladder. The bladder design was preferable to the piston design because the bladder was easier to fabricate and less susceptible to leaks.

**Batch RO cycle: phases of operation**

The operation of a true batch RO system using a bladder is illustrated in Figure 4. During the permeate production phase (A1, A2), feed circulates in a loop around the system. At the beginning of the batch cycle (A1) the circulation loop is at its maximum volume and the bladder at its minimum volume. The permeate production phase ends (A2) when the desired amount of permeate is produced (typically but not necessarily when the bladder is full). The combined volume of the circulation loop and the bladder remains the same throughout the permeate production phase. Make-up fluid is pumped by the high-pressure pump into the bladder and, as the bladder expands, permeate exits the circulation loop through the membrane at a similar rate. The permeate flow rate thus approximately matches the high-pressure pump’s flow rate.1

Following permeate production, the flush and recharge phases reset the system. At the end of permeate production (Figure 4A2) the circulation loop is filled with a concentrated brine solution. Brine is ejected from the system during the flush phase (Figure 4B). The circulation loop is broken (by opening and closing valves, not shown here) and new feed is introduced into the system. The incoming feed pushes brine out of the system and the system salinity reapproaches the feed salinity.

After the flush phase, the valves are opened or closed so that there is no outlet from the circulation loop. Instead, the valve blocking the bladder outlet is opened. During the recharge phase (Figure 4C) new feed is pumped into the system and the bladder is simultaneously emptied of make-up fluid. Once all make-up fluid has left the bladder, the recharge phase is complete and another batch cycle can restart with the permeate production phase.

### 1.3. Previous work

Two studies [10, 11] developed various models of RO processes and found that true batch RO consumes less energy than other RO processes. Closed-circuit RO (or semi-batch RO) energy consumption is higher due to continuous mixing (i.e., entropy generation) between fresh feed and brine during the permeate production phase. Batch RO with ERDs consumes more energy than true batch RO due to constant throttling of feed through the imperfect ERDs. These batch RO models included a circulation pump. The models mentioned above have not been validated with experimental results until now.

While numerous studies have used numerical models to investigate the potential benefits of batch RO systems [10, 11, 15], few systems have been built and operated. Davies et al. built and operated the first energy-efficient true batch RO system with a rigid piston [13]. They reported energy consumption lower than the minimum energy consumption of a single-stage continuous RO process (i.e., $SEC_{SSRO,ideal} = \pi f / (1 - RR)$) at relatively high recoveries. These measurements included the hydraulic work of the high-pressure pump but not the circulation pump.

### 1.4. Outcomes of present work

In this study, we designed, built, and operated the first true batch RO system using a flexible bladder. We validated models of batch RO energy consumption by measuring the hydraulic work of both the high-pressure pump and the circulation pump. The experimental measurements agree well with the model (≤ 2.7% difference) after accounting for concentration polarization.

Although batch RO consumes less energy for water separation than continuous RO, inherent inefficiencies are associated with batch processing. In a batch system, no permeate is produced during the flush and recharge phases, so the average flux is lower than its operating flux (the flux during permeate production) [16]. As reported previously [5, 17, 18], we find that batch RO systems will operate at a salinity higher than the intake salinity due to salt retention. Previous studies have compared batch RO and continuous RO energy consumption without considering these specific inefficiencies [10, 11]. Despite these inefficiencies, we find that batch RO can still save energy compared to continuous RO, although savings are lower than previously expected.

In Section 2, we share experimental data and observations from the operation of our true batch RO prototype. In Section 3, we validate the energy predictions from an existing model, which we modified to account for concentration polarization and the recharge phase. In Section 4, we present a revised estimate of the energy savings.
Figure 4: The batch RO cycle is composed of three phases, shown here as implemented in our true batch system with a bladder. During the permeate production phase (A1,A2), the circulation loop volume decreases but remains under pressure due to the expanding bladder. The flush and recharge phases (B,C) reset the system salinity and empty the bladder so that the batch cycle can be repeated. The make-up fluid is collected and reused. Dotted arrows indicate inactive flowpaths.
achievable by batch RO plants, accounting for salt retention. In Section 5, we discuss the results of our work and their implications on future directions of batch RO research. The appendices contain additional details on our experiments and analysis.

2. True batch RO prototype

In this section we share details about the true batch RO prototype and its successful operation. We built the bench-scale prototype (shown in Figure 5) using off-the-shelf parts except for the custom-molded silicone bladder, which has operated for over 100 consecutive cycles with no issues.

2.1. Overview

The bladder and RO membrane are both housed inside a pressure vessel. As explained in Section 1, the bladder starts out empty at the beginning of the batch RO cycle. During the permeate production phase, a make-up fluid (water, in this case) is pumped into the bladder. A roughly equal amount of water leaves the pressure vessel as permeate through the RO membrane. We provide more details on the batch RO prototype in Appendix A.

We collected pressure and flow data in order to measure the energy consumption and permeate production of the batch RO system. The pressure and flow data over the course of a single batch cycle are shown in Figure 6. The permeate production phase takes place over the first twelve minutes of the cycle. Once the high-pressure pump is turned on, the feed pressure takes about a minute to build up (see Appendix B). The permeate flux rises along with the feed pressure until it reaches a steady-state value. The feed pressure rises slowly over the course of the permeate production phase in order to drive a steady permeate flux (the system salinity increases gradually as permeate is produced). Once the desired amount of permeate is produced, the variable frequency drive (VFD) that drives the high-pressure pump is stopped and valves are opened in order to flush brine out of the system.

The beginning of the flush phase is marked by the rapid fall in feed pressure: the system depressurizes once valves are opened. This depressurization does not result in any significant change of internal energy (or temperature or entropy), because the liquid is essentially incompressible. During the flush phase, the circulation pump brings new feed water into the system and brine is ejected from the system. Once the desired amount of brine is flushed out of the system, valves open or close in order to start the recharge phase. The beginning of the recharge phase is marked by a rise in the circulation pressure and a fall in the circulation flow. The circulation pump is bringing new feed into the system but must push the make-up fluid out of the bladder. The added resistance to flow is due to the bladder. When the bladder is nearly empty, the circulation pressure rises rapidly and the circulation flow comes to a halt. The recharge phase is complete when there is no more circulation flow. At that point, valves...
Figure 6: Pressure and flow data over the course of a single batch RO cycle. The feed pressure gradually rises over the course of the permeate production phase in order to drive a constant permeate flow. The circulation pressure and flow are relatively constant throughout the entire batch RO cycle. The circulation pressure is greater during the recharge phase since the bladder is being emptied. The system recovery ratio for this batch cycle was 49.5%.

open or close to return back to the permeate production phase. The high-pressure pump turns back on and another batch cycle commences.

There are several practical concerns that arise during the flush and recharge phases, and some of these may be more problematic at higher salinities, as in seawater reverse osmosis (SWRO). Permeate quality can be worse in batch RO compared to continuous RO due to salt passage across the membrane during the flush and recharge phases. Permeate may also be lost due to osmotic backwash during the flush and recharge phases so the plant recovery ratio may be lower than the system recovery ratio.

2.2. Salt passage

Davies et al. observed that permeate quality was initially very poor at the beginning of the batch cycle but improved drastically after the “bad” permeate was flushed out [13]. We have also observed a similar phenomenon in our system. When the batch RO system is unpressurized (as when the system is off or during the flush and recharge phases), there is no water flux but salt continues to pass through the membrane from the feed channel to the permeate channel. Salt passage is driven by a concentration gradient across the membrane, independent of any pressure gradient. Thus, the permeate is relatively salty at the beginning of the next permeate production phase.

As shown in Table 1 permeate quality is worst when the permeate production phase is short (i.e. at low recovery ratios). All other factors held constant, if the permeate production phase is longer (i.e. at higher recovery ratios)
Table 1: Permeate quality is worse at the beginning of the permeate production phase due to salt passage before the system is turned on or during the flush and recharge phases. As shown in the upper portion of this table, permeate quality improves at higher recoveries (i.e. longer permeate production phases) since the salty permeate gets diluted with more fresh permeate. Permeate quality is worse for the first batch RO cycle but improves on subsequent cycles due to the shorter time between cycles.

| Cycle number | Feed salinity [g NaCl/kg] | Recovery ratio [%] | Brine salinity [g NaCl/kg] | Permeate salinity [%] | Permeate salinity [g NaCl/kg] | Salt rejection [%] |
|--------------|---------------------------|-------------------|-----------------------------|----------------------|-----------------------------|------------------|
| 1            | 2.0                       | 29                | 2.8                         | 72                   | 0.31                        | 87.0             |
| 1            | 2.0                       | 48                | 3.8                         | 81                   | 0.19                        | 93.4             |
| 1            | 2.0                       | 52                | 4.1                         | 84                   | 0.14                        | 95.5             |
| 2-5          | 2.0                       | 52                | 4.11                        | 81-86                | 0.08                        | 97.51            |

1 The actual values for these quantities are expected to be higher than shown here due to salt retention between batch cycles. The salt rejection given here is taken as a lower bound on the actual performance. For more on salt retention, see Section 4.4.

The permeate quality improves since the additional fresh permeate can dilute the salty permeate. This problem may be mitigated in a batch RO plant which operates at high recovery ratios and continuously throughout the day.

We also observe permeate quality improving after the initial system start-up (lower portion of Table 1). When the system is left off there is a relatively long time (hours) for salt to pass through the membrane. Once the system is running in successive cycles, there is only a shorter time (~2 minutes) for salt to pass through the membrane during the flush and recharge phases. This phenomenon is not unique to batch RO; any RO membrane system will be subject to impaired start-up performance [19].

2.3. Osmotic backwash

We have observed osmotic backwash during the flush and recharge phases: permeate re-enters the membrane module through the permeate tube once the system is depressurized. Although the batch RO system consumed energy to produce this permeate, we were not able to actually collect the permeate. Osmotic backwash could lead to entrainment of air into the system, which may result in membrane contamination, dry-out, or fouling [19]. Osmotic backwash may be more problematic at higher salinities since the rate of backwash depends on the salinity difference across the membrane. We do not attempt to quantify the effects of osmotic backwash in the present study.

3. Experimental validation of batch RO model

Previous studies have modeled the energy consumption of batch RO, but there has been no comparison to experimental data. We compared experimental measurements from our true batch RO prototype to a numerical model of batch RO. Our measurements agree with the model at the various feed salinities, recovery ratios, and fluxes tested. The high-pressure pump work is slightly overestimated in all cases. We believe that the newly-validated model can be used to predict batch RO performance under realistic operating conditions.

3.1. Experimental results

We calculated the hydraulic work (the product of flow rate and pressure rise, Q∆P) done by the high-pressure and circulation pumps using pressure and flow rate measurements. We assumed that the flow through the high-pressure pump is constant throughout the permeate production phase and equal to the steady-state permeate flow. We used these measurements to validate the energy predictions from an existing model [10], which we modified to account for concentration polarization and the recharge phase. The model predictions for specific energy consumption (SEC) agreed with the experimental measurements (largest error: -2.7%). In this section, we define the error as:

$$
\text{Error} = \frac{\text{Model} - \text{Experiment}}{\text{Experiment}} \times 100
$$

We show the overall results from fifteen tests in Figure 7. We ran a single batch RO cycle for different combinations of feed salinity and flux at various recovery ratios. The model predictions for SEC agree with the experimental measurements, with the largest error (-2.7%) occurring at low recoveries where the cycle times are relatively short. The predicted SEC for low salinity and high flux (w_f = 2 g/kg, J_{sys} = 20 L m^{-2} h^{-1}) levels out at
The results agree well at higher recoveries, where we expect batch RO systems to be most beneficial due to higher variation in osmotic pressure.

Figure 7: The model predictions for SEC agree with experimentally measured SEC (largest error: -2.7%) at various feed salinities, fluxes, and recovery ratios. This model accounts for concentration polarization. Note that the y-axis does not start from zero. Error bars indicate 95% confidence intervals for the measured data [20, 21]. Dotted blue lines indicate upper and lower bounds on predicted energy consumption based on 95% confidence intervals of the measured membrane permeability.

We present the measured and predicted SEC in Table 2. The model consistently underestimates circulation pump work and consistently overestimates high-pressure pump work, so those errors partially offset each other. The model underestimates the circulation pump work due to non-zero start-up time (see Appendix B). The model may overestimate the high-pressure pump work because it does not account for the pressure rise during start-up. At operating conditions higher than 10 bar (as in seawater RO), we expect the model to consistently overestimate the overall SEC, since the high-pressure pump work will be much greater than the circulation pump work. The SEC figures listed here represent the hydraulic work and therefore do not reflect pump efficiencies, which are presumed to be low at the bench-scale compared to commercial-scale pumps.

Our experiments were limited to operating pressures below 10 bar and recovery ratios below 55%. The parameters used in the model validation are shown in Table 3. The osmotic pressure and density of aqueous sodium chloride solutions were calculated using a MATLAB implementation of the Pitzer equations [22, 23]. See Appendix A for more experimental details.

Concentration polarization

We adjusted the original model from Warsinger et al. [10] to account for concentration polarization (CP). The original model (without CP) consistently underestimated the high-pressure pump work (largest error: -3.35%). After adjusting the model to account for CP the predicted SEC rose in all cases and the model consistently overestimated the high-pressure pump work (largest error: 1.53%). This difference is as expected since higher pressures are required to overcome the elevated osmotic pressure difference at the membrane surface relative to the bulk.

The concentration at the membrane surface is calculated according to the film theory model of concentration polarization [24, 25]. The mass transfer coefficient is obtained via the Sherwood number, which we calculated using correlations from an experimental study on spacer-filled channels [26, 27]. The adjusted model is included in the accompanying data repository [28].
Table 2: Experimental measurements and corresponding model predictions. We list the total specific energy consumption (Total) in addition to the contributions from the high-pressure pump (HP) and circulation pump (C). Some numbers do not add up precisely due to rounding.

| Feed salinity [g NaCl/kg] | Permeate flux [L m\(^{-2}\) h\(^{-1}\)] | Recovery ratio [%] | SEC\(_{\text{model}}\) [kW h m\(^{-3}\)] | SEC\(_{\text{exp}}\) [kW h m\(^{-3}\)] | Error [%] | Percent of SEC\(_{\text{Total}}\) [%] |
|--------------------------|-------------------------------|------------------|-------------------------|-------------------------|--------|-------------------------|
| 2.0                      | 20                            | 28.7             | HP 0.192                | 0.192                   | 0.87   | 79                      |
|                          |                               |                  | C 0.048                 | 0.052                   | −6.98  | 21                      |
|                          |                               |                  | Total 0.242             | 0.244                   | −0.80  | -                       |
| 38.3                     | HP 0.197                      | 0.197            | 0.81                     | 80                      |
|                          | C 0.046                       | 0.049            | −6.28                   | 20                      |
|                          | Total 0.244                   | 0.245            | −0.60                   | -                       |
| 43.1                     | HP 0.200                      | 0.200            | 0.35                     | 81                      |
|                          | C 0.048                       | 0.048            | −6.05                   | 19                      |
|                          | Total 0.245                   | 0.248            | −0.88                   | -                       |
| 49.8                     | HP 0.204                      | 0.204            | 0.50                     | 81                      |
|                          | C 0.044                       | 0.047            | −7.41                   | 18                      |
|                          | Total 0.249                   | 0.251            | −0.99                   | -                       |
| 3.5                      | 10                            | 28.6             | HP 0.168                | 0.168                   | 0.18   | 65                      |
|                          | C 0.082                       | 0.089            | −8.20                   | 35                      |
|                          | Total 0.251                   | 0.258            | −2.66                   | -                       |
| 39.5                     | HP 0.177                      | 0.177            | 0.17                     | 68                      |
|                          | C 0.079                       | 0.082            | −5.84                   | 32                      |
|                          | Total 0.256                   | 0.259            | −1.21                   | -                       |
| 49.4                     | HP 0.185                      | 0.185            | 0.92                     | 70                      |
|                          | C 0.078                       | 0.080            | −3.03                   | 30                      |
|                          | Total 0.263                   | 0.265            | −0.44                   | -                       |
| 53.4                     | HP 0.188                      | 0.188            | 0.62                     | 70                      |
|                          | C 0.077                       | 0.081            | −4.16                   | 30                      |
|                          | Total 0.267                   | 0.269            | −0.57                   | -                       |
| 3.5                      | 15                            | 29.7             | HP 0.202                | 0.202                   | 0.28   | 75                      |
|                          | C 0.062                       | 0.066            | −6.78                   | 25                      |
|                          | Total 0.264                   | 0.268            | −1.54                   | -                       |
| 39.6                     | HP 0.206                      | 0.206            | 0.01                     | 77                      |
|                          | C 0.059                       | 0.062            | −4.49                   | 23                      |
|                          | Total 0.268                   | 0.268            | 0.24                    | -                       |
| 49.5                     | HP 0.217                      | 0.217            | 0.68                     | 79                      |
|                          | C 0.058                       | 0.059            | −2.59                   | 21                      |
|                          | Total 0.276                   | 0.276            | 0.17                    | -                       |
| 52.3                     | HP 0.221                      | 0.221            | 0.96                     | 79                      |
|                          | C 0.057                       | 0.059            | −3.18                   | 21                      |
|                          | Total 0.280                   | 0.280            | −0.19                   | -                       |
| 5.0                      | 10                            | 29.7             | HP 0.210                | 0.208                   | 1.33   | 70                      |
|                          | C 0.083                       | 0.088            | −5.48                   | 30                      |
|                          | Total 0.293                   | 0.296            | −0.69                   | -                       |
| 39.6                     | HP 0.221                      | 0.218            | 1.16                     | 72                      |
|                          | C 0.080                       | 0.085            | −5.63                   | 28                      |
|                          | Total 0.301                   | 0.304            | −0.75                   | -                       |
| 44.5                     | HP 0.227                      | 0.223            | 1.53                     | 73                      |
|                          | C 0.080                       | 0.084            | −4.73                   | 27                      |
|                          | Total 0.307                   | 0.307            | 0.17                    | -                       |
Table 3: Parameters used in validation of the batch RO model. We used a 2.5 in. (6.4 cm) spiral wound membrane element (Hydranautics ESPA-2514) in these experiments.

| Parameter                          | Note       | Value | Units |
|------------------------------------|------------|-------|-------|
| **Operational parameters**         |            |       |       |
| Intake feed salinity               |            | 2-5   | g NaCl/kg |
| Recovery ratios                    |            | 29-53 | %     |
| Operating flux                     |            | 10-20 | L m\(^{-2}\) h\(^{-1}\) |
| Initial feed channel velocity      |            | 0.06  | m/s   |
| Concentration polarization factor  | calculated | 1.07-1.14 | - |
| Maximum feed pressure              |            | 10    | bar   |
| Circulation loop pressure drop     |            | 0.1   | bar   |
| Circulation pump flowrate          |            | 2     | L/min |
| Permeate pressure                  |            | 0.09  | bar   |
| **Batch RO model inputs**          |            |       |       |
| Membrane element area              |            | 0.47  | m\(^2\) |
| Membrane water permeability        |            | 4.05-4.25 | L m\(^{-2}\) h\(^{-1}\) bar\(^{-1}\) |
| Batch RO system volume             |            | 2.8   | L     |
| High-pressure pump efficiency      |            | 1     | -     |
| Circulation pump efficiency        |            | 1     | -     |

Membrane permeability

We calculated the membrane permeability, \(A\), for each series of tests (with the same feed salinity and flux) according to the following equation:

\[
A = \frac{J}{P_f - CP\pi_f}
\]  

where \(J\) is the instantaneous permeate flux, \(P_f\) is the feed pressure, \(CP\) is the concentration polarization factor, and \(\pi_f\) is the osmotic pressure of the feed (as calculated according to the average system concentration). The calculated membrane water permeability was higher (4.25 ± 0.11 L m\(^{-2}\) h\(^{-1}\) bar\(^{-1}\)) for one set of tests \((w_f = 2 \text{ g/kg}, J_{sys} = 20 \text{ L m}^{-2} \text{ h}^{-1})\) compared to the other three sets of tests \((4.07 ± 0.24, 4.06 ± 0.14, \text{ and } 4.05 ± 0.21 \text{ L m}^{-2} \text{ h}^{-1} \text{ bar}^{-1})\).

The reason for this 5% difference is unclear to us. Permeate flux could be a factor, but further investigation is needed. We also considered membrane compaction as a potential cause. However, there is no clear pattern in the calculated membrane permeability between successive tests.

Standard membrane compaction tests involve steady-state operation at test conditions (i.e., pressure and flux) with deionized water for hours at a time. Those tests may not be applicable to the batch RO process, because the membrane is regularly unpressurized during the flush and recharge phases. In lieu of membrane compaction tests, we operated the batch RO system for 28 hours prior to the model validation tests, using similar conditions (feed water, flux, and recovery). We used the same membrane element for all tests. There were no fouling or scaling agents in our feed water, so we did not clean the membrane between tests. As future studies delve into fouling or scaling in batch RO, it will be important to develop a membrane compaction test appropriate to the batch RO process.

We used permeability data from the beginning of the permeate production phase (first twenty seconds after reaching the nominal permeate flux) since we expect the salinity of feed in the membrane module to be close to the average system salinity at that point, as opposed to towards the end. It was also important to avoid using data during the pressure and flux fluctuations that occurred several minutes into some tests (see Appendix B).

4. Energy comparison: batch RO vs. continuous RO

The batch RO process has been previously shown to consume less energy than conventional continuous RO [10, 11]. These studies slightly overestimate the potential energy savings of a batch RO system by assuming that both RO systems operate at the same conditions (e.g., flux, feed salinity). That approach does not take into account inefficiencies inherent to batch processing: intermittent permeate production and elevated feed salinity. These inefficiencies are unavoidable in the real-life context of running a batch RO plant.

We have compared the energy consumption of a batch RO plant to a continuous RO plant while adjusting for batch processing inefficiencies. Despite operating at elevated feed salinity and flux, a batch RO plant will still save energy compared to a corresponding continuous RO plant. These energy savings are less than previously calculated, primarily due to the elevated feed salinity.
In this section, we establish a new framework for comparing the energy consumption of batch RO plants to continuous RO plants. We present a revised estimate of the energy savings achievable by batch RO plants. Finally, we provide the expressions used to account for batch processing inefficiencies.

4.1. System conditions vs. plant conditions

When considering the batch RO process in a real-world context, it is important to draw a distinction between the batch system’s operating conditions and the desalination plant’s overall conditions. Specifically, the system flux and system recovery ratio are defined with respect to the permeate production phase, whereas the plant flux and plant recovery ratio are time-averaged values across the duration of the entire batch cycle. The system feed salinity refers to the average salinity at the beginning of the permeate production phase, whereas the plant feed salinity refers to the salinity of the feed water at the plant intake.

Previous comparisons between batch RO and continuous RO kept system conditions fixed [10, 11]. This approach is flawed because, unlike continuous RO, batch RO system conditions are different from plant conditions. A batch RO system’s energy consumption depends on system conditions, but the desalination plant’s revenue and costs are more related to plant conditions. This difference is most easily understood by looking at a batch RO system’s operating flux versus the plant’s overall flux (average permeate production rate per unit membrane area).

As shown in Figure 8, a batch RO system might operate at a flux of 14.9 L m$^{-2}$ h$^{-1}$ during the permeate production phase, but the overall flux of the desalination plant (averaged over an entire batch cycle) is only 12.5 L m$^{-2}$ h$^{-1}$. The system’s operating flux is needed to calculate the energy consumption of a batch RO system, since the operating flux determines the feed pressures during the permeate production phase. On the other hand, the plant’s overall flux corresponds to the desalination plant’s revenues from selling fresh water since it accounts for the downtime in permeate production, which a continuous RO plant would not be subject to. A batch RO system must therefore operate at a higher flux than a corresponding continuous RO system in order to match the permeate production and revenue of the continuous RO plant (assuming equal membrane area).

Elevated feed salinity due to salt retention

A batch RO system will operate at a feed salinity higher than the plant’s intake feed salinity. This elevated feed salinity occurs due to incomplete flushing of brine from the system. This phenomenon, which takes place during the flush phase, is illustrated in Figure 9. At the beginning of the flush phase, the system is full of brine. In the ideal (and unrealistic) case, plug flow occurs and the incoming feed would perfectly displace the brine. At the end of the flush phase, all brine would be ejected and the system salinity would match the feed salinity [5, 17].

In reality, mixing occurs between the incoming feed and the outgoing brine. At the end of the flush phase, a mixture of feed and brine remains in the system. The initial feed salinity at the beginning of the next batch cycle is greater than the incoming feed salinity because salt is retained in the system between successive batch cycles. This substantially increases the energy consumption of the batch RO system.

In a continuous RO plant, the initial feed salinity seen by the RO module is roughly equal to the intake feed salinity. However, in a batch RO plant the initial feed salinity at the beginning of a batch cycle will be higher than the intake feed salinity. A proper comparison between batch RO and continuous RO plants must take feed salinity elevation into account.

Comparing batch RO to continuous RO

When comparing batch RO to continuous RO plant, our aim is to keep revenue and costs constant between the two plants, as much as possible. In each comparison, we have set the total feed flow rate, total permeate flow rate, and total membrane area equal between each desalination plant. In doing so, we presume that revenue, pretreatment costs, brine disposal costs, and membrane costs will be similar for both plants. This analysis does
not directly address the additional capital costs associated with batch or multi-stage RO configurations. We have separated those costs from the other costs and focus on the energetic benefits of the batch system. The remaining question (which we do not attempt to answer here) is whether the energy savings achieved by a batch system will outweigh those additional capital costs.

We match each plant’s overall conditions because those will correspond to the total feed and permeate flow rates. However, we must also identify system operating conditions because they will affect the energy requirements of the batch RO process. Using the expressions described in Section 4.4, we calculate the system conditions of the batch RO system for a given set of plant conditions. Then we can use the validated model (see Section 3) to predict the energy consumption of the batch RO system. We compare this to the energy consumption of a continuous RO system with the same plant conditions (i.e., feed salinity, flux, and recovery ratio).

For example, consider a continuous RO desalination plant with an intake feed salinity of 35 g/kg, a recovery ratio of 50% and an average permeate flux of 15 L m$^{-2}$ h$^{-1}$ as shown in Figure 10. The system’s operating conditions are similar to the plant conditions. Prior comparisons between batch RO and continuous RO made the system conditions the same [13, 10, 11]. A batch RO plant operating at the same system conditions as the continuous plant would save significant energy compared to a continuous RO plant (1.72 vs. 2.21 kWh/m$^3$, or a 22% savings). However, the plant’s permeate flux would only be 12.3 L m$^{-2}$ h$^{-1}$ and the plant’s intake feed salinity would have to be lower (32 g/kg) due to salt retention. This comparison is unreasonable.

In our comparison, we match the batch RO plant conditions to the plant conditions of the continuous RO plant. We then calculate the corresponding system conditions. In this example, the batch RO system operates at a higher feed salinity (38 vs. 35 g/kg) and flux (19 vs. 15 L m$^{-2}$ h$^{-1}$) due to the inefficiencies described above. Despite operating at elevated conditions, this batch RO plant would still save considerable energy compared to the continuous RO plant (1.97 vs. 2.21 kWh/m$^3$, or an 11% savings). The savings are less than previously expected.

This comparison is based on predictions from the numerical model we validated in Section 3. The results presented in this section reflect operating conditions not covered in the model validation, because operation of the batch prototype was limited to pressures below 10 bar and recoveries below 55%. As discussed in Section 3.1 we are confident that the model can be used to predict batch RO energy consumption at higher recoveries and salinities, but our results might ideally supported by experimental validation of the model at higher pressure and recoveries.

4.2. Batch RO energy savings

In this section, we compare the energy consumptions of batch RO and continuous RO under various conditions. As described above, we keep the plant conditions fixed between the two processes.

Figure 11 shows the energy consumption of various seawater RO processes. A batch RO plant would consume more energy than previously thought as a result of operating at elevated feed salinity and flux. At low recovery ratios, batch RO energy consumption is even higher than a single-stage continuous RO plant. All systems are modeled as single-pass RO. Batch RO begins to save energy compared to a single-stage plant at recoveries above 27%. Today’s seawater RO plants typically operate at recovery ratios between 40-50% [3, 29], although there are reports of two-stage RO plants operating at up to 60% [30, 31].
Figure 10: In continuous RO, system conditions are similar to plant conditions as opposed to batch RO. Previous comparisons between batch RO and continuous RO predicted significant energy savings but kept system conditions fixed between the two processes (an unreasonable comparison). In our comparison we match the plant conditions and find that batch RO will still save energy despite operating at an elevated flux and feed salinity.
Figure 11: Energy consumption of various seawater RO processes over a range of recovery ratios. A batch RO plant would consume more energy than previously thought \([10, 11]\) due to operating at elevated feed salinity. Seawater is approximated by 35 g/kg aqueous sodium chloride solution. The overall plant flux is 15 L m\(^{-2}\) h\(^{-1}\), the membrane water permeability is 3 L m\(^{-2}\) h\(^{-1}\) bar\(^{-1}\), and the dimensionless flush time is unity (see Equation 5).

We modeled the energy consumption of continuous two-stage RO plants based on two classes of ERDs. Centrifugal ERDs (Pelton turbine and turbochargers) can reach transfer efficiencies of 80% and have been used in two-stage plants \([30]\). Batch RO consumes less energy than two-stage RO with centrifugal ERDs at all recoveries. Isobaric ERDs are highly efficient (95%) but to our knowledge have not yet been incorporated into two-stage RO plants. A hypothetical two-stage RO system with isobaric ERDs would be more energy efficient, consuming a bit less energy than batch RO at recoveries below 63%.

Figure 12 shows the energy consumption of various brackish RO processes. Similarly to the seawater case, we find that batch RO energy consumption becomes more competitive with staged RO plants at higher recovery ratios where the osmotic pressure varies more throughout the process. Batch RO energy consumption matches that of a two-stage RO plant at a recovery of 85%. Batch RO and three-stage RO consume similar amounts of energy at an overall recovery ratio of 92%. We modeled the multi-stage brackish RO systems with centrifugal ERDs (\(\eta = 80\%\)). There is a relatively little energy remaining in the retentate of the multi-stage brackish RO systems, so the energy consumption does not improve much with higher efficiency ERDs (not shown here). Some three-stage RO plants may not utilize energy recovery at all due to the low energy requirements of brackish water desalination.

Table 4 shows the parameters used to generate Figures 11 and 12. In the staged continuous RO configurations, each stage has the same average flux as the overall plant flux. We used our experimentally validated batch RO model (Section 3) and a previously validated continuous RO model \([8]\) to calculate the energy consumption of batch and continuous RO systems. Both models account for concentration polarization and assume perfect salt rejection. Results from prior work \([12]\) and our model indicate that fewer elements per pressure vessel are desirable in true batch RO systems. Today’s single-stage continuous RO plants typically have 6-8 elements per pressure vessel \([3]\). In the multi-stage continuous RO systems, we place slightly more elements in the front stage(s) to optimize energy consumption \([8]\).

Sensitivity analysis

Batch RO energy consumption is sensitive to the values of system parameters, shown in Table 4. In Table 5, we show the effect of changing several parameters on batch RO energy consumption, single-stage continuous RO energy consumption, and the energy savings (or penalty) achieved by a batch RO plant. In this sensitivity analysis, we vary only one parameter at a time. The values of all other parameters are set to the base case.

Since there is no energy recovery in a true batch RO system,\(^2\) the ERD efficiency does not affect batch RO energy consumption. In a couple unfavorable cases, batch RO consumes more energy than single-stage continuous

\(^2\)In a continuous RO process, the entire feed stream is pressurized so typically the energy remaining in the pressurized brine stream is recovered by ERDs. In a true batch RO process only the permeate volume passes through the pump, so there is no pressurized fluid leaving the system and therefore no need for energy recovery.
Figure 12: Energy consumption of various brackish water RO processes over a range of recovery ratios. Batch RO consumes more energy than a single-stage continuous RO plant at recoveries below 74% due to operating at elevated feed salinity, which prior work did not account for [10, 11]. Brackish water is approximated by 5 g/kg aqueous sodium chloride solution. The overall plant flux is 25 L m$^{-2}$ h$^{-1}$, the membrane water permeability is 5 L m$^{-2}$ h$^{-1}$ bar$^{-1}$, and the dimensionless flush time is unity (see Equation 5).

RO (3 elements per pressure vessel, $A = 1$ L m$^{-2}$ h$^{-1}$ bar$^{-1}$). It is preferable for true batch RO systems to have fewer elements per pressure vessel in order to minimize the spatial variation of osmotic pressure [12]. The energetic benefits of batch RO are greater with higher permeability membranes because feed pressures can closely approach the osmotic pressure curve (and energy consumption approaches the least work); for lower permeability membranes, the benefit of osmotic pressure matching is outweighed by the effect of salt retention.

Single-stage continuous RO systems that use isobaric ERDs may also be susceptible to feed salinity elevation, because the brine and feed streams come into direct contact within the ERD [32]. There is limited data on mixing in today’s ERDs, but earlier sources report volumetric mixing of 3-6% [32, 33, 34], resulting in a feed salinity elevation of 1.5-3% in a reverse osmosis system operating at 50% recovery. In the case of 6% volumetric mixing, the energy savings of batch RO would increase to 14% (Table 5).

Double-acting batch RO

Other authors have proposed a double-acting batch RO system to compensate for the downtime in a batch RO cycle [35, 11]. A double-acting system adds a second feed tank to allow the membrane to produce permeate during the recharge phase. A double-acting system does not produce permeate continuously because the system is still depressurized during the flush phase.

Table 6 shows the energy consumption of various batch RO systems and a continuous RO system for reference. The double-acting batch RO system achieves a 2% energy savings compared to the standard batch RO system, despite operating at a lower flux (17.1 vs 19.2 L m$^{-2}$ h$^{-1}$). Meanwhile, prior comparisons assumed that a batch system would operate at 15 L m$^{-2}$ h$^{-1}$ with an energy consumption 12% lower than our findings [10, 11]. The double-acting system is not able to meaningfully lower the energy consumption of batch RO because it must also operate at an elevated feed salinity relative to prior comparisons. Feed salinity elevation is responsible for a majority of the increase in energy consumption, as we explain next.

Effects of feed salinity and flux on energy consumption

Here we isolate the effects of feed salinity and flux on the energy consumption of batch RO systems. In Table 7, we show the energy consumption of various hypothetical batch RO systems. The system recovery ratio is 50% in each case. The top row is our baseline: the energy consumption of a batch RO system with an system feed salinity of 35 g NaCl/kg and an average system flux of 15 L m$^{-2}$ h$^{-1}$. This baseline is representative of previous comparisons, when authors assumed that a batch RO system would operate at the same conditions as a continuous RO plant.

The following rows show the effects of elevated flux only, elevated feed salinity only, and both elevated flux and feed salinity on the energy consumption of batch RO systems. A batch RO system operating at elevated flux (19.2 vs 15.0 L m$^{-2}$ h$^{-1}$) but at the baseline feed salinity consumes only 4.7% more energy than the baseline. On
Table 4: Parameters used in energetic analysis of seawater and brackish reverse osmosis systems. Seawater and brackish water are approximated by aqueous sodium chloride solution.

| Parameter                        | Note       | Value | Units          |
|----------------------------------|------------|-------|----------------|
| **Batch & continuous RO**        |            |       |                |
| Membrane element area            |            | 37    | m²             |
| Total elements in system         |            | 14    |                |
| Pump efficiency                  |            | 0.8   |                |
| Intake feed salinity             | Seawater   | 35    | g/kg           |
| Membrane water permeability      |            | 3     | L m⁻² h⁻¹ bar⁻¹|
| Overall plant flux               |            | 15    | L m⁻² h⁻¹      |
| Intake feed salinity             | Brackish water | 5   | g/kg          |
| Membrane water permeability      |            | 5     | L m⁻² h⁻¹ bar⁻¹|
| Overall plant flux               |            | 25    | L m⁻² h⁻¹      |
| **Batch RO**                     |            |       |                |
| Elements per pressure vessel (PV)| Seawater   | 2     |                |
|                                   | Brackish water | 1 |                |
| Circulation loop pressure drop   |            | 0.1   | bar            |
| Initial feed channel velocity    |            | 0.2   | m/s            |
| Dimensionless flush time         |            | 1     |                |
| **Continuous RO**                |            |       |                |
| ERD efficiency                   | Single-stage | 0.95 |                |
|                                   | Two-stage   | 0.8, 0.95 |                |
|                                   | Three-stage | 0.8   |                |
| Elements per PV, by stage        | Single-stage | 7    |                |
|                                   | Two-stage   | 4-3   |                |
|                                   | Three-stage | 3-3-2 |                |
| Pressure vessels per stage       | Single-stage | 2    |                |
|                                   | Two-stage   | 2-2   |                |
|                                   | Three-stage | 2-2-1 |                |

the other hand, a batch RO system operating at elevated feed salinity (38 vs 35 g/kg) but at the baseline flux (15.0 L m⁻² h⁻¹) consumes 9.3% more energy than the baseline case. In this case, the elevated feed salinity accounts for two thirds of the total increase in batch RO energy consumption due to operating at both elevated feed salinity and flux.

4.3. Implications

We have demonstrated that a batch RO plant can save energy relative to a single-stage continuous RO plant, despite operating at elevated feed salinity and flux. We sought to keep plant revenue and operating costs (e.g., membranes, pretreatment, brine disposal) fixed in our comparison by matching plant conditions. However, a batch RO plant may cost more than a single-stage continuous RO plant due to additional capital expenditures (CAPEX), such as bladders, valves, and additional pressure vessels. While a batch RO plant may consume less energy than a staged continuous RO plant, we have not evaluated the levelized cost of water (LCOW) and cannot say which plant has the lower LCOW. Staged RO plants also require additional CAPEX (pumps, ERDs).

Batch RO systems should be carefully designed to minimize the amount of brine that remains in a system from cycle to cycle. Overflushing (operating the flush phase for a longer duration) is one way to decrease the operating feed salinity, but at the expense of throwing away pretreated feed and decreasing the overall plant recovery ratio. A double-acting batch RO system also provides only marginal energy savings relative to a single-acting batch RO system since it would still operate at elevated feed salinities.

Our comparison assumes a constant pump efficiency throughout the batch cycle. In reality, the pump efficiency may vary throughout the batch cycle since a batch RO system must operate over a wide range of pressures, whereas the pump in a continuous RO system operates at a relatively constant pressure.

4.4. Expressions for system and plant conditions

In this section we describe the expressions used to calculate system conditions based off of plant conditions. The system flux and recovery ratio are relatively straightforward derivations. The feed salinity elevation may be explained theoretically by Taylor dispersion through the flat channels in the membrane module. Experimental measurements support our calculations.
Table 5: Batch RO energy consumption varies significantly depending on the elements per pressure vessel and membrane water permeability. \(\Delta SEC\) is relative to the base case. Energy savings = \((SEC_{SSRO} - SEC_{batch})/SEC_{SSRO}\). In all cases the plant intake feed salinity is 35 g NaCl/kg, the overall recovery ratio is 50%, and the plant flux is 15 L m\(^{-2}\) h\(^{-1}\). L: low case. B: base case. H: high case.

| Parameter                          | Case | Value | \(\Delta SEC_{batch}\) [%] | \(\Delta SEC_{SSRO}\) [%] | Energy savings [%] |
|------------------------------------|------|-------|--------------------------|--------------------------|-------------------|
| Isobaric ERD efficiency            | L    | 92%   | -                        | +2.7                     | 13.4              |
|                                    | B    | 95%   | -                        | -                        | 11.0              |
|                                    | H    | 98%   | -                        | -2.7                     | 8.5               |
| Elements per PV (batch RO only)    | L    | 1 element | -2.2                   | -                        | 13.0              |
|                                    | B    | 2      | -                        | -                        | 11.0              |
|                                    | H    | 4      | +15.6                    | -                        | (2.9)             |
| Membrane water permeability        | L    | 1 L m\(^{-2}\) h\(^{-1}\) bar\(^{-1}\) | +22.6                    | +6.3                     | (2.6)             |
|                                    | B    | 3      | -                        | -                        | 11.0              |
|                                    | H    | 5      | -4.5                     | -0.2                     | 14.9              |
| Isobaric ERD volumetric mixing     | B    | 0%     | -                        | -                        | 11.0              |
|                                    |      | 3%     | -                        | +1.6                     | 12.5              |
|                                    | H    | 6%     | -                        | +3.4                     | 14.0              |

Table 6: A double-acting batch RO system has similar energy consumption to a single-acting batch RO system, despite operating at lower flux. The double-acting system is still subject to feed salinity elevation. The recovery ratio is 50% and the membrane water permeability is 3 L m\(^{-2}\) h\(^{-1}\) bar\(^{-1}\).

| System flux [L m\(^{-2}\) h\(^{-1}\)] | Plant flux [L m\(^{-2}\) h\(^{-1}\)] | System feed salinity [g NaCl/kg] | Plant feed salinity [g NaCl/kg] | SEC [kWh/m\(^3\)] |
|--------------------------------------|--------------------------------------|---------------------------------|---------------------------------|-------------------|
| Continuous RO                        | 15.0                                 | 35                              | 35                              | 2.21              |
| Batch RO                             | 19.2                                 | 38                              | 35                              | 1.97              |
| Double-acting batch RO               | 17.1                                 | 38                              | 35                              | 1.93              |
| Batch RO - prior work                | 15.0                                 | 12.3                            | 35                              | 1.72              |

Salt retention

When a batch RO system is starting up, the feed salinity at the beginning of the first batch cycle will match the plant feed salinity. However, during the flush phase the incoming feed and outgoing brine will mix (Taylor dispersion)[5]. The feed salinity (averaged over the system volume) at the beginning of the second batch cycle will be greater than the first cycle. After several batch cycles the system feed salinity reaches a steady-state value.

It is useful to introduce a dimensionless feed salinity elevation \(\Theta\):

\[
\Theta = \frac{w_{f,sys} - w_{f,pl}}{w_{b,pl} - w_{f,pl}}
\]

where \(w_{f,sys}\) is the system feed salinity, \(w_{f,pl}\) is the plant feed salinity, and \(w_{b,pl}\) is the initial brine salinity. The numerator of this equation is simply the feed salinity elevation of the batch system. The difference in the initial brine salinity and the plant feed salinity characterizes the mixing that occurs during the flush phase so we normalize by that quantity.

We also introduce the dimensionless flush time \(t^*\):

\[
t^* = \frac{t_f}{V_b/Q_{circ}} = \frac{t_f}{t_{f,plug}}
\]

where \(t_f\) is the duration of the flush phase, \(V_b\) is the volume of brine in the system (membrane module, pressure vessel, and piping) at the end of the permeate production phase, and \(Q_{circ}\) is the flow rate of the circulation pump during the flush phase. \(t_{f,plug}\) would be the flush time required to reject all of the brine (but no feed) from the system under ideal plug flow conditions.

We calculated the dimensionless feed salinity elevation \(\Theta\) as a function of the dimensionless flush time \(t^*\) assuming ideal plug flow and Taylor dispersion through a flat channel [36, 37, 5]. At steady-state, the salt entering the system during the flush and recharge phases matches the salt exiting the system during the flush phase. The salt
Table 7: A batch RO plant will consume more energy than previously thought primarily due to feed salinity elevation. Operating at elevated flux alone would result in a relatively small increase in energy consumed.

|                      | System flux [L m\(^{-2}\) h\(^{-1}\)] | System feed salinity [g NaCl/kg] | SEC [kWh/m\(^3\)] |
|----------------------|----------------------------------------|----------------------------------|-------------------|
| Batch RO - prior work | 15.0                                   | 35                               | 1.72              |
| Batch RO - elevated flux only | 19.2                                   | 35                               | 1.80              |
| Batch RO - elevated feed salinity only | 15.0                                   | 38                               | 1.88              |
| Batch RO - elevated feed salinity and elevated flux | 19.2                                   | 38                               | 1.97              |

Figure 13: Feed salinity elevation decreases as the flush time increases. The measured feed salinity elevation is close to the elevation calculated according to Taylor dispersion in a flat channel. The plug flow curve is shown here for reference but is not achievable in reality. Error bars indicate 95% confidence intervals.

We also measured the feed salinity elevation of our batch RO prototype. We ran successive batch cycles in a row while keeping all operating parameters constant until the system reached a steady-state (see Appendix A).

Our calculations and measurements are shown in Figure 13. The measured curve lies close to the calculated curve for \( t^* \leq 1 \). We expect most batch systems to operate in this region to avoid throwing out pretreated feed. At larger values of \( t^* \), the discrepancy between the measurements and the calculations increase. We believe this is due to residual brine in piping dead legs, where there is no flow during the flush phase: no matter how long we flush the system, that brine will always remain in the system for the next batch cycle. Dead legs during the flush phase account for 2% of the total system volume.

Note that the plug flow curve and Taylor dispersion curve coincide at relatively low values of \( t^* \). When \( t^* \) is less than the dimensionless transition flush time \( t^*_{tr} \), the flush phase does not last long enough for any of the incoming feed to make it out of the system. If the feed/brine mixture does not leave the system, the elevated feed salinity ends up being the same as in the plug flow case. Assuming a laminar velocity profile in a flat channel (an approximation of flow through the membrane module) \( t^*_{tr} = 2/3 \) since the mean velocity is two thirds the maximum velocity (at the center of the pipe) [5]. We observe a transition flush time close to the expected value from theory: \( t^*_{tr} \approx 0.71 \) for our system.

The measured curve was obtained by running batch cycles with a plant feed salinity of 2 g/kg, as indicated by the circular data points. We expect this curve to predict the elevation for any plant feed salinity since the mixing is purely convective. In order to verify the predictive ability of the measured curve, we ran additional tests at feed salinities of 3 and 4 g/kg (square data points) but could not go much higher due to limits on the operating pressure.
The results from these tests lie close to the measured curve. In our system, the Péclet number is large enough (80,000) that the effects of diffusion are negligible [36, 37]. We expect other batch RO systems to also operate at high Péclet numbers during the flush phase to minimize the flush time.

Our batch RO prototype has a maximum recovery ratio of 53%, but other batch RO systems may operate at higher recoveries. Figure 13 is only applicable at a recovery ratio of 52% (all tests were run at that recovery). We combine the calculated curve for ideal plug flow with the measured curve to form a general piecewise function for feed salinity elevation:

$$w_{f,sys} = \begin{cases} w_{f,p} [1 + RR_{sys} (t^* - 1)] & \text{for } t^* \leq t^*_f \\ w_{f,p} [1 + RR_{sys} (t^*_f - 1) e^{-C_p (t^* - t^*_f)}] & \text{for } t^* > t^*_f \end{cases}$$

where $RR_{sys}$ is the system recovery ratio and $C_p$ is the system’s flushing effectiveness coefficient. For $t^* \leq t^*_f$, the measured feed salinity coincides with the plug flow curve. For $t^* > t^*_f$, the feed salinity elevation curve departs from the ideal plug flow curve. We chose to fit the data to an exponential function; fitting the data to a power function yields similar results. Feed salinity elevation is at higher recovery ratios (all other factors held equal) because the brine salinity increases with recovery ratio. We measured $C_p = 2.57$ and $t^*_f = 0.71$ for our true batch RO prototype.

We can form a simplified (but not general) expression for the feed salinity elevation of a batch RO system that has similar system characteristics (i.e., $C_p$ and $t^*_f$) as our prototype and operates at $t^* = 1$:

$$w_{f,sys} = w_{f,p} [1 + 0.19RR_{pl}]$$

**Flux**

In order to achieve a desired plant flux $J_{pl}$, a batch RO system must operate at an increased system flux $J_{sys}$ to compensate for the downtime in permeate production. The system flux is related to the plant flux by the ratio of the batch cycle time $t_{cyc}$ to the permeate production time $t_{pp}$:

$$J_{sys} t_{pp} = J_{pl} t_{cyc} \Rightarrow J_{sys} = J_{pl} \frac{t_{cyc}}{t_{pp}}$$

The system flux is always higher than the plant flux because the batch cycle time is greater than the permeate production time. The system flux can be reduced by minimizing the flush and recharge times $t_f$ and $t_r$:

$$J_{sys} = J_{pl} \frac{t_{cyc}}{t_{cyc} - (t_f + t_r)} = J_{pl} \frac{1}{1 - \frac{J_{pl} A_m}{V_p} F V_p + V_p \frac{Q_{circ}}{V_p}} \frac{V_p}{t_{cyc}} \frac{t_f + t_r}{t_f + t_r}$$

where $Q_{circ}$ is the volumetric flow rate of the circulation pump during the flush and recharge phases. Maximizing the circulation pump flow rate during the reset phases is the most effective way to decrease system flux. Operating at a reduced value of dimensionless flush time $t^*_f$ would also reduce the flush time but at the penalty of increasing the feed salinity as described above. $V_p$ is the volume of permeate produced and $V_b$ is the volume of brine remaining at the end of the permeate production phase. These volumes are presumed to be constrained by other factors along with the membrane area $A_m$.

A double-acting batch RO system would reduce, but not eliminate, the downtime of a batch cycle [35] by enabling permeate production during the recharge phase. As described in the literature, a double-acting system would not produce permeate during the flush phase. The relationship between system flux and plant flux for double-acting batch RO is found as follows:

$$J_{sys} t_{pp} = J_{pl} (t_{pp} + t_f) \Rightarrow J_{sys} = J_{pl} \frac{t_{pp} + t_f}{t_{pp}}$$

$$J_{sys} = J_{pl} \frac{t_{pp} + t_f}{t_{pp} + t_f - t_f} = J_{pl} \frac{1}{1 - \frac{J_{pl} A_m}{V_p} F V_p}$$

A double-acting system reduces the required system flux to achieve a desired plant flux. However, double-acting systems still operate at elevated feed salinities, which increases the required work input (Table 7).
Recovery ratio

During the permeate production phase, a batch RO system recovers a volume of permeate \( V_p \) from the initial system volume \( V_{sys} \). The expression for a batch RO system recovery ratio \( RR_{sys} \) is:

\[
RR_{sys} = \frac{V_p}{V_{sys}} = \frac{V_p}{V_b + V_p}
\]  

(11)

where \( V_b \) is the volume of brine remaining in the circulation loop at the end of the permeate production phase. The system recovery ratio is relevant to calculating the energy consumption of a batch system. However, it does not necessarily reflect the amount of feed water that is taken in by the desalination plant. The plant recovery ratio \( RR_{pl} \) accounts for this:

\[
RR_{pl} = \frac{V_p}{V_{f,fl} + V_{f,xe}}
\]  

(12)

where \( V_{f,fl} \) and \( V_{f,xe} \) indicate the volume of feed that enters the system during the flush and recharge phases, respectively. The plant recovery ratio should be used when calculating pretreatment and brine disposal costs, since it is the ratio of permeate produced by the batch system to the feed water taken in by the desalination plant. \( V_{f,fl} \) and \( V_{f,xe} \) correspond to \( t^*V_b \) and \( V_p \),\(^3\) so we can relate the system recovery ratio to the plant recovery ratio:

\[
RR_{pl} = \frac{V_p}{t^*V_b + V_p} = \frac{RR_{sys}}{t^*(1 - RR_{sys}) + RR_{sys}}
\]  

(13)

where \( t^* \) is the dimensionless flush time, introduced above. Since the system volume is equal to the sum of the brine and permeate volumes, the system recovery ratio is the same as the plant recovery ratio when \( t^* \) is unity. When \( t^* > 1 \), the plant recovery ratio is smaller than the system recovery ratio: feed water is thrown out of the plant without being desalted.

While not reflected in the equations above, readers should note that the plant recovery ratio will also be affected by osmotic backwash that occurs during the reset phases (described in Section 2).

5. Discussion

In this study we designed, built, and tested the first true batch RO system using a flexible bladder and used it to validate an improved model of batch RO energy consumption. We have demonstrated the successful operation of batch RO with a bladder, operating for over a hundred consecutive cycles with no bladder-related issues.

We have found that permeate quality towards the beginning of the permeate production phase is worse in batch RO due to salt passage during the reset phases. However, the apparent salt rejection in a batch RO plant may be comparable to a continuous RO plant as long as they both have the same overall plant flux. High-quality permeate water may be needed for specific industrial applications or to comply with potable water regulations. Further studies should investigate the ability of batch RO to meet those requirements.

Due to osmotic backwash during the reset phases, some amount of permeate reenters the feed channel and will not be recovered. This will slightly decrease the overall plant recovery relative to the system recovery, but may also mitigate organic fouling and/or biofouling [38]. Osmotic backwash is a diffusion-based phenomena and may play a larger role as feed salinities increase beyond those tested (max feed salinity = 5 g/kg). Future work should model and measure the magnitude of osmotic backwash at higher salinities, such as in seawater reverse osmosis (SWRO).

We measured the energy consumption of our batch RO prototype via flowrate and pressure measurements in order to validate a model of batch RO energy consumption. We adjusted a previous model to account for the effects of concentration polarization. The adjusted model agrees well with the experimental measurements (≤ 2.7% difference) for the variety of feed salinities (2-5 g/kg), recovery ratios (29-53%), and system fluxes (10-20 L m\(^{-2}\) h\(^{-1}\)) tested. We believe that the model may be used for applications at higher salinities (as in SWRO). A clear next step is to confirm the validity of the model in the SWRO range by building a batch RO system capable of operating at higher pressures than our prototype. Our work only measured the pump’s hydraulic work and did not account for pump efficiencies. Future work should consider that pump efficiency may vary over the range of operating pressures in a batch cycle.

We revised previous comparisons between the energy consumption of batch RO and continuous RO to account for practical inefficiencies associated with batch RO. The energy consumption is greater than previously expected,

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\(^3\)This assumes that airspace in the system is negligible. See Appendix B.
mostly due to salt retention. A batch RO system will operate at a feed salinity that is higher than the plant’s intake salinity (e.g., 38 g/kg rather than 35 g/kg at 50% recovery) due to incomplete flushing of the brine between cycles. This means that batch RO energy consumption will generally be higher than previously expected. Although not quantified in this study, we note that other time-variant RO processes (semi-batch or closed circuit RO [18] and pulse-flow RO) will also be susceptible to salt retention. Future studies on batch RO and other time-variant processes must take into account the feed salinity elevation when estimating energy consumption.

In all of our comparisons, the dimensionless flush time \( t^* \) is unity. The amount of feed salinity elevation depends on the duration of the flush phase, but we expect that plant operators would prefer to avoid significant overflushing or underflushing. Overflushing, or operating the flush phase for longer durations \( (t^* > 1) \), would lower the system feed salinity but would decrease the overall plant recovery ratio since pretreated feed is being thrown out without being desalted. Underflushing \( (t^* < 1) \) would increase the overall plant recovery ratio but at the expense of increasing the system feed salinity even more.

Designers of batch RO systems should take care in designing flow paths to minimize the amount of brine that is retained in the valves and piping during the flush phase (i.e., dead legs). In fact, Swaminathan et al. [12] concluded that tank and piping volume should be minimized in batch systems to minimize spatial variation of osmotic pressure. Dead legs during the flush phase make up 2% of the system volume in our prototype. Salt retention may be even higher in batch systems with higher proportions of dead volume.

A double-acting batch RO system has been proposed and would operate at a lower flux than a single-acting batch RO system. However, it would provide marginal energetic benefits over a single-acting batch RO system since it still operates at an elevated feed salinity. The additional energy savings of a double-acting batch RO system may or may not justify the additional costs and complexity of such a system. To our knowledge, nobody has implemented a double-acting batch RO system. There is more work to be done on investigating the design challenges and costs associated with such a system.

Batch RO systems may have additional benefits besides improved energy consumption. In a true batch RO system, it is desirable to have a fewer elements per pressures vessel in order to minimize the spatial variation of osmotic pressure. The maximum flux seen at the front end of the lead element may be smaller than is seen in continuous RO systems with more elements per pressures vessel. This could mitigate fouling of the membranes. The osmotic backwash described above may also help to decrease membrane fouling. Finally, some authors have hypothesized that salinity cycling in batch RO (and other time-variant RO processes) prevents scaling [10, 11]. However, Lee et al. [39] found that semi-batch RO was more prone to gypsum scaling because higher levels of supersaturation were reached during the permeate production phase. Future work must be done to investigate scaling in time-variant RO processes.

What is the outlook for batch RO? Similar to staged RO or other time-variant RO processes, batch RO will save more energy compared to single-stage continuous RO at higher recoveries, where there is a large variation in osmotic pressure. However, we find that batch RO will save less than previously expected (our findings: 11% savings for 50% recovery of seawater at 15 L m\(^{-2}\) h\(^{-1}\)). Batch RO will also save energy compared to today’s multi-stage RO plants due to its inherent energy recovery. However, energy consumption is not the full picture. There are additional costs (CAPEX) associated with batch RO and staged RO plants. Future work should consider the full array of potential costs and benefits associated with alternative RO configurations to determine which process is the most cost-effective.

6. Summary and conclusions

We have built and successfully operated the first true batch RO system using a flexible bladder. Our work has three main conclusions:

- We have validated a model of true batch RO energy consumption with experimental measurements. We believe this model may be used to predict batch RO energy consumption.

- A batch RO system will operate at a feed salinity that is higher than the plant’s intake salinity (e.g., 38 g/kg rather than 35 g/kg at 50% recovery) due to salt retention between cycles.

- True batch RO is an energy-efficient process with the potential to reduce the cost of water desalination.

Future work should further investigate the potential benefits and drawbacks of batch RO systems. Is batch RO a cheaper alternative to single-stage or multi-stage continuous RO? A full cost analysis, accounting for CAPEX and other factors, will be needed to answer that question.
Data statement

The raw experimental data and the data processing code are provided in the accompanying data repository [28].

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Declarations of interest

Declarations of interest: none.
Appendix A  Experimental details

The pressure vessel (AMI PV2540SSAU-316) houses both the spiral wound membrane element (Hydranautics ESPA-2514) and the make-up bladder. We sealed off one end of the membrane module’s permeate tube with a metal cap. The bladder was custom molded and is made of a duplication silicone (Zhermack Elite Double 22). During initial testing, we ruptured some bladders by filling them past capacity. This can be avoided by carefully keeping track of the volume of make-up fluid in the bladder at all times.

The high-pressure pump is a positive displacement pump (Hydra-Cell F-20) driven by an electric motor (Marathon 56T17F5322). The flow rate of the high-pressure pump was controlled with a variable frequency drive (Automation Direct GS1). The circulation pump was a hot water circulation pump with a brushless motor (Yosoo DC 12V) and was rated for operation up to 10 bar, which limited the system pressure. Five valves were used to change flow paths during each of the phases (Magnatrol 18A52-W and 18AR52-W, Asco 8210G030 and 8210G087). Valves were opened and closed through solid state relays controlled by a LabVIEW Virtual Instrument.

We measured feed pressure (taken to be net pressure at the membrane) with a pressure transmitter (Wika A-10). We measured the pressure difference across the circulation pump with an digital differential pressure gauge (Omega DPG409-015DWU). We measured permeate and circulation flow rates with two flow meters (permeate: McMillan G-111, circulation: McMillan 104). Pressure and flow measurements were collected with a data acquisition unit (Omega OM-USB-1608GX) and processed by the same LabVIEW Virtual Instrument used for valve control. The relative locations of all instrumentation is shown in Figure 14.

The feed water was formulated in the lab using deionized water (Process and Water, Type II) and lab-grade sodium chloride (Sigma-Aldrich ACS reagent, ≥ 99.0% purity). We measured feed, brine, and permeate conductivities with a conductivity probe (Hach CDC40101 IntelliCAL) and converted those values to concentrations via linear interpolation of experimental data [40, 41].

We believe that the dead volumes (shown in Figure 14) during flush contribute partially to salt retention. These dead volumes account for 2% of the total system volume. The dead volume by valve 4 could be reduced by replacing valves 3-5 with a three way valve or manifold. Another dead volume is next to the front end of the bladder; we believe the flow is reduced at this location because it is occluded from the feed inlet by the bladder’s nozzle.

We designed our batch RO system to avoid entrainment of air into the system. We attached a length of tubing to the permeate outlet of the pressure vessel and secured the tubing to a vertical post, running it up about one meter above the pressure vessel. At the top of the post we bent the tubing into an arch so that the tubing outlet lay directly over a collection tub. During the permeate production phase the water level in the tubing would rise until it reached the apex of the arch and could drip into a collection tub. When the system depressurizes at the beginning of a flush phase, we visually observe the water level fall from the apex, rapidly at first but then at a steadier rate. During the recharge phase the water level would fall more slowly than during the flush phase. We made the tubing long enough such that the water level would still be above the pressure vessel at the end of the recharge phase (to avoid air entrainment).
Experimental procedure

For the model validation measurements (Fig. 7) it was important that the initial average system salinity at the beginning of the batch phase matched the nominal feed salinity (e.g., 2 g/kg). Prior to each test, we mixed up feed solution to the appropriate conductivity. We flushed the system of brine from the previous cycle by introducing the newly-mixed feed. Throughout this step we switched between batch mode and flush mode in order to flush brine remaining in dead legs. We measured the outlet stream’s conductivity until it was within 2% of the feed solution’s conductivity, and then emptied the bladder. At this point we switched to batch mode and began the test, running one complete batch cycle at the nominal recovery ratio. Each data point in Figure 7 is a single-sample observation. We ran five auxiliary tests at the same initial feed salinity, flux, and recovery ratio in order to assess the repeatability of our energy measurements and to calculate the 95% confidence intervals [20, 21].

In order to quantify the feed salinity elevation of our batch RO prototype (Fig. 13), we measured the initial average system salinity at the beginning of the batch phase once the system reached a steady state. We ran consecutive batch cycles, keeping relevant operating conditions (feed salinity, recovery ratio, and dimensionless flush time) constant. The first batch cycle of each test started out at the nominal feed salinity (e.g., 2 g/kg), and increased in successive cycles due to salt retention. After each batch phase we measured the average conductivity of the rejected brine, which increased in successive cycles along with the initial system salinity. We ended the test once there were three consecutive cycles where the brine conductivity varied by less than 2%. In our tests, this happened within 5-10 cycles.

At the end of the each test, we measured the volume average system salinity. First, we allowed the feed in the system to recirculate for three passes so that the solution would be well-mixed. Next, we switched to the flush phase and collected five successive samples of the reject stream (80 mL each). We measured the conductivity of each sample. The conductivities of the second and third samples varied by less than 1%; we took the average of these measurements to be the average system salinity. The conductivities of the fourth and fifth samples were consistently lower than the first and second samples by up to ∼3%; we attribute this to osmotic backwash. The conductivity of the first sample was much higher than the rest of the samples. It was closer to the brine salinity at the end of the final batch phase because some of that solution was collected from the dead leg with low (or no) flow during the permeate production phase that immediately preceded the system outlet at valve 5 (Figure 14).

Each data point in Figure 13 represents a single-sample observation. We took eight auxiliary measurements in order to assess the repeatability of the measurement method described above and to calculate the 95% confidence intervals. We were able to take these measurements in concert with some of our steady-state tests because we expected the average system salinity at the end of the first batch phase to be the same in each of those tests, which started at the same initial feed salinity (2 g/kg) and operated at the same recovery ratio (52%). At the end of the first batch phase we turned off the high-pressure pump, but did not immediately switch to the flush phase. We allowed the remaining brine to recirculate for three passes and then measured five samples of the reject stream, as above. Again, we used the measurements of the second and third samples as proxies for the average system salinity. In this case, the first sample’s conductivity was much lower than the rest of the samples (closer to the initial feed salinity at the beginning of the batch phase).

The raw experimental data and the data processing code are provided in the accompanying data repository [28].

Appendix B  Residual air

Despite our best efforts at de-airing the system, we believe that there was still residual air in the system throughout our tests. We attribute the system start-up time (∼1 minute) and the pressure and flux fluctuations to residual air. The system pressure does not respond instantly to changes in flow, as we would expect if there were no air in the system.

System start-up

If all fluid contents in the batch RO system were perfectly incompressible (i.e., only filled with water) and the system components were perfectly rigid, then permeate production would start instantly once the high pressure pump is turned on, as shown in Figure 15 (model flux curve). In reality, we observe the pressure increase gradually, and there is a non-zero start-up time before the system flux is achieved. This shows that residual air remained in the system. Air is compressible, so when the high pressure pump introduces water into the empty bladder, the pressure does not rise as rapidly as it would if the system were only filled with water. In our experiments, $t_{50}$,

\[t_{50}\]

We decreased the operating flux towards the end of the batch phase in later cycles in order to keep the pressure under 10 bar.
the time that it takes for the permeate flux to reach 50% of the system flux, has ranged from 20-44 seconds. The circulation pump operates throughout this start-up time, so it consumes more energy than the model predicts.

We think that the start-up time can be reduced by eliminating dead space in the system. However, a gradual pressure build-up may be desirable in order to reduce mechanical stress on the membrane elements [19].

![Graph showing model and actual flux](image)

Figure 15: The model underestimates the circulation pump work since the batch cycle last longer than predicted. The model assumes permeate production starts instantly at the desired flux. In an actual system there is a non-zero start-up time, likely due to residual air in the system. Here, it takes 29 seconds for the permeate flux to reach 50% of the system flux.

**Pressure and flux fluctuations**

In some tests, we observed pressure and flux fluctuations occur several minutes into the permeate production phase. We show some of these fluctuations in Figure 16. At point 1 the pressure departs from its gradual ascent and dips. This is followed by a sharp increase in the permeate flux, which peaks about twelve seconds later just before the pressure hits a local minimum at point 2, ten seconds afterwards. The pressure rises to another peak at point 3 (34 seconds after point 2) as the permeate flux falls to a local minimum and then sharply increases. The permeate flux hits another peak just before the feed pressure hits the last obvious dip at point 4 (23 seconds after point 2).

These fluctuations might be explained physically by the spatial variation in feed concentration. The dip in pressure at point 1 would occur when relatively less salty feed enters the membrane module. The drop in osmotic pressure would cause the net driving pressure and permeate flux to increase. The lag in behavior between the permeate flux and the feed pressure suggests that there is air in the system. Assuming the high pressure pump flow is constant throughout the permeate production phase, this increase in permeate flux would be accompanied by a drop in pressure as the residual air in the system is allowed to expand (due to the net outflow of water). As the pressure drops (to a minimum at point 2), so does the permeate flux (to a minimum between points 2 and 3). As less water leaves the system, the residual air would compress and pressure rises again (to a peak at point 3) due to the net inflow of water.

These fluctuations do not appear in every single test. When they do appear, they do not start at the same time. Sometimes multiple fluctuations appear in a row, and other times only one fluctuation is apparent.

**Appendix C  Feed salinity derivations**

Here we present the derivation of the expressions for feed salinity elevation presented in Section 4.4. We have calculated these expressions for both ideal plug flow conditions (unrealistic) and realistic conditions (Taylor dispersion). In both cases we are interested in the steady-state feed salinity of the batch RO system. When a batch RO system is first started, the system feed salinity will initially be the same as the plant feed salinity. After the first batch cycle, the system feed salinity will rise a bit due to incomplete flushing of brine. The system feed salinity will eventually reach a steady-state value. We are interested in this steady-state feed salinity since we assume that batch RO desalination plants will operate for many cycles consecutively.
For both of these derivations, we are interested in the volume average system feed salinity at the beginning of the permeate production phase. This value will reach a steady-state value when the amount of salt that leaves the system is equal to the amount of salt that enters the system between each cycle:

\[ m_{\text{salt,in,fl}} + m_{\text{salt,in,re}} = m_{\text{salt,out,fl}} \]  \hspace{1cm} (14)

where \( m_{\text{salt,in,fl}} \) and \( m_{\text{salt,in,re}} \) are the masses of salt that enter the system during the flush and recharge phases, respectively and \( m_{\text{salt,out,fl}} \) is the mass of salt that exits the system during the flush phase. Zero salt flux into the permeate is assumed.

Ideal plug flow

Under ideal plug flow conditions Equation 14 is readily expressed in terms of the incoming and outgoing concentrations and mass flows:

\[ w_{f,pl}[m_{\text{sol,in,fl}} + m_{\text{sol,in,re}}] = w_{b,sys}[m_{\text{sol,out,fl}}] \]  \hspace{1cm} (15)

where \( m_{\text{sol,in,fl}} \), \( m_{\text{sol,in,re}} \), and \( m_{\text{sol,out,fl}} \) are the masses of solution that enter and exit the system in the appropriate phases. \( w_{f,pl} \) is the plant intake feed salinity and \( w_{b,sys} \) is the system brine salinity when operating under steady-state conditions. Neglecting the effects of osmotic backwash, we assume the mass of feedwater entering the system is equal to the mass of brine exiting the system during the flush phase. We rearrange this equation and put it in terms of RO operating parameters:

\[ w_{b,sys} = w_{f,pl}\left[1 + \frac{m_{\text{sol,in,re}}}{m_{\text{sol,out,fl}}}\right] = w_{f,pl}\left[1 + \frac{RR}{r^*(1 - RR)}\right] \]  \hspace{1cm} (16)

where \( RR \) is the system operating recovery ratio and \( r^* \) is the non-dimensional flush time as defined in Section 5.

We are interested in the system feed salinity rather than the brine salinity. We make a simplifying assumption that salt does not enter the permeate channel to express the system brine salinity (from Eq. 16) in terms of the system feed salinity:

\[ w_{f,sys} = w_{b,sys}(1 - RR) = w_{f,pl}\left[1 - RR + \frac{RR}{r^*}\right] \]  \hspace{1cm} (17)
where $w_{f,sys}$ is the system operating feed salinity (the quantity we are interested in). It is useful to express the feed salinity elevation as a dimensionless quantity, so we substitute the equation above into Equation 4:

$$\Theta = (1 - RR) \left( \frac{1}{t^*} - 1 \right)$$  (18)

where $\Theta$ is the dimensionless feed salinity elevation. As expected, the feed salinity elevation is zero when $t^*$ is unity. As described in Section 4.4, Equation 18 is valid for system that are operating with $t^*$ less than the transition flush time $t^*_tr$.

**Realistic conditions - Taylor dispersion**

Under realistic conditions, there is convective mixing between the incoming feed and the outgoing brine. Calculating the mass of salt that leaves the system during the flush phase is not as simple as in the case of plug flow. The concentration at the system outlet varies over time so we must integrate over the duration of the flush phase $t_f$:

$$w_{f,pl} \left[ m_{sol,in,fl} + m_{sol,in,te} \right] = \int_{t=0}^{t=t_f} w_{exit}(t) \dot{m}_{sol,out,fl} \, dt$$  (19)

where $w_{exit}(t)$ is the concentration of the feed/brine mixture at the exit of the system at any time $t$. $\dot{m}_{sol,out,fl}$ is the mass flow rate of solution exiting the system and is assumed constant throughout the flush phase. At this point we recognize that it takes time for the incoming feed to travel to the system exit. Towards the beginning of the flush phase the concentration at the system exit will simply be the brine salinity:

$$w_{f,pl} \left[ m_{sol,in,fl} + m_{sol,in,te} \right] = w_{b,sys} \dot{m}_{sol,out,fl} t_{tr} + \dot{m}_{sol,out,fl} \int_{t=0}^{t=t_f} w_{exit}(t) \, dt$$  (20)

where $t_{tr}$ is the (dimensional) transition flush time, which depends on the flow velocity profile throughout the system. We assume that the system concentration is uniform ($w_{b,sys}$) at the beginning of the flush phase in order to simplify the analysis.

At this point we assume that the mixing is purely convective, as is valid at high Peclet numbers [36, 37, 5]. In purely convective mixing, the concentration profile in time and space simply depends on the flow velocity profile. We can write an expression for the outlet concentration as long as we keep track of the leading edge of the interface between the incoming feed and outgoing brine. The concentration at the system exit is:

$$w_{exit}(t) = w_{b,sys} \left[ 1 - \frac{r(t)}{a} \right] + w_{f,pl} \left[ \frac{r(t)}{a} \right]$$  (21)

where $a$ is the width of the entire channel and $r(t)$ is the width of the channel that is occupied by feedwater. We rearrange the terms in this equation:

$$\frac{w_{exit}(t) - w_{f,pl}}{w_{b,sys} - w_{f,pl}} = 1 - \frac{r(t)}{a}$$  (22)

We model the flow as laminar flow through a flat channel, which approximates the flow through the membrane channel. Qiu and Davies found that it is reasonable to neglect the spacer’s effect on dispersion [5]. The expression for $r(t)$ is known [36, 5]:

$$\frac{r(t)}{a} = \sqrt{1 - \frac{2}{3t^*}} \text{ for } t^* \geq t^*_c$$  (23)

where $t^*$ is the dimensionless flush time and $t^*_c$ is the dimensionless flush transition time. This expression shows us that $t^*_c = 2/3$ for laminar flow through a flat channel, which matches our experimental measurements. We evaluated the integral in Equation 20 numerically with MATLAB’s `integral` function. We also calculated results for laminar flow through a circular pipe (as in the circular tubing) but our experimental data agreed much better with the flat channel expression.
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