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Cost and energy requirements of hybrid RO and ED brine concentration systems for salt production

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Abstract

A new concept to concentrate seawater up to 200 g/kg for producing vacuum salt using a reverse osmosis (RO) system hybridized with an electrodialysis (ED) system is presented. The RO system operates up to pressures of 120 bar and concentrates seawater up to 120 g/kg with the ED system concentrating RO brine to 200 g/kg. A parametric analysis to minimize the specific cost of brine concentration was conducted. Parameters varied were: the degree of RO-ED hybridization, ED current density, electricity prices and water prices. Optimal hybrid RO-ED designs reduced brine concentration costs by 33-70\% over standalone ED systems, with revenue generated from water co-production further subsidizing costs by 1-6\%. Optimizing ED current density reduced costs the most. Including a crystallizer, the total reduction in production cost over a standalone ED-crystallizer system was 19-55\%, with the production cost for a typical case being $111/\text{tonne-salt}$. The proposed RO-ED-crystallizer (REC) systems were found to be techno-economically feasible in Cyprus, Japan, Kuwait, Saudi Arabia, and the USA. At a road transportation distance of 735 km, REC based seawater vacuum salt was competitive with conventional vacuum salt. REC systems may open up the potential of small-scale decentralized salt production.

Keywords: reverse osmosis, electrodialysis, hybrid, salt production, brine concentration, seawater

1. Introduction

Globally, each year, more than 280 million tonnes of salt\textsuperscript{[1]} is produced from solar evaporation of seawater, from conventional rock mining and from the solution mining of saline brines\textsuperscript{[2]}. Of this, 39\% is used by the chloralkali industry for the production of chlorine and sodium hydroxide, 22\% for human consumption, 22\% by the soda ash industry, 9\% for de-icing of roads and 9\% for other uses\textsuperscript{[2]}. Salt is
important both as a crucial nutritional source and as a crucial raw material for global chemical production. Given the strategic importance of salt for the world, producers globally are looking at ways to make salt production more sustainable and reduce production costs. Salt production costs vary significantly with the method of production, with the lowest production costs being for “solar salt” (around $5-10/tonne of salt) and the highest production costs being for “vacuum salt” (around $30-50/tonne-salt for conventional vacuum salt produced from saturated brines). In “solar salt” production, seawater is collected in evaporation ponds and freely available solar insolation is used to evaporate off water leaving behind salt. Solar salt production requires significant land areas and is used only in locations where land prices are low. In conventional “vacuum salt” production, saturated solutions of brine (S = 260 g/kg) extracted from solution mines are sent to thermal or electrically driven evaporators and crystallizers (referred in this paper hence simply as “crystallizer”) to evaporate off water and produce salt. Vacuum salt is more expensive than solar salt production primarily because of the higher capital costs in equipment needed and the energy costs since in the latter solar energy is used for free. Despite the differing production costs, the global salt market can support both methods of salt production partly because of high transportation costs. Transporting 1 tonne of any commodity per 100 km costs around $10.34/tonne-100 km by truck, 2.53 $/tonne-100 km by rail, and 1.14 $/tonne-100 km by barge. Thus at a certain distance and between certain cities, vacuum salt becomes a more competitive option.

![Figure 1: Simplified flow diagram of a conventional salt production plant with an ED system and crystallizer](image)

One of the most expensive types of vacuum salt is vacuum salt produced from seawater through the use of brine concentration and crystallizer systems. In such systems seawater (S = 35 g/kg) is first concentrated to near saturation levels (S = 180-260 g/kg) using a separate brine concentration system before being sent to a crystallizer. Separate brine concentration systems are needed because crystallizers are designed to be cost effective only when using nearly saturated brine. Significant energy is needed to concentrate seawater to near saturation levels with the required volume reduction being around 90% of the initial volume. When low cost land
is not available for the construction of solar evaporation ponds, separate brine concentration systems such as Mechanical Vapor Compression (MVC) [7–11] or Electrodialysis (ED) [12–18] needs to be used. To the best of our knowledge, the only commercial vacuum salt production systems that produce salt from seawater through a combination of brine concentration systems and crystallizers are plants in Japan, Korea and Kuwait [14, 17, 19] that use ED to first concentrate seawater from 35 g/kg to around 180-200 g/kg after which brine is sent to a crystallizer where salt is produced for either human consumption [19] or for chlor-alkali production [14]. A flow diagram of such a typical ED based salt production plant is shown in Fig. 1.

ED is a versatile electric driven membrane based desalination technology, that was first conceptually conceived for demineralizing sugar syrup in 1890 [20], and developed for saline water desalination in the 1940s and 1950’s [20, 21]. ED systems primarily consist of several pairs of anion exchange membranes (AEM), and cation exchange membranes (CEM) placed in between a cathode and anode (shown later in Fig. 5). Feed water is fed into channels created between CEMs and AEMs. When a voltage is applied, cations move to the cathode and anions move to the anode; however, the AEMs prevent the movement of cations and CEMs prevent the movement of anions resulting in one channel becoming concentrated in ions while the adjacent channel is depleted of ions. Consequently, these two adjacent channels are referred to as the ‘concentrate’ and ‘diluate’ channels. Depending on the type of AEMs and CEMs used, ED can be designed for several applications including city-scale brackish water desalination [17, 21], village-scale water treatment [22], seawater brine concentration [14–19], denitrification of water for municipal water supply [21], demineralization of wine, whey and sugar [23], in-home water treatment [24–26] and wastewater treatment [17, 27]. For salt production from seawater, AEMs and CEMs that are additionally selective to monovalent ions are used. The resulting monovalent selective ED (MSED) systems concentrates sodium chloride preferentially over other ions in seawater leading to the production of brines of salinity 180 g/kg to 200 g/kg that are rich in sodium chloride. MSED systems for concentrating brine for salt production have been in commercial operation for more than 50 years in Japan [19].

In this paper, our objective is the production of salt from seawater and our analysis is restricted to MSED systems. For the convenience of readers, henceforth in this paper, we will be referring to MSED systems simply as ED systems. The flow diagram of a conventional salt production plant using ED is shown in Fig. 1. Seawater feed, typically at 35 g/kg, first flows in to both the diluate and concentrate channels of an ED stack. In the concentrate channel, seawater is typically concentrated from 35 g/kg to 177-200 g/kg with the diluate discharged back in to the sea at a salinity much less than seawater. Typically the salinity change happens along the length of a single membrane with the voltage kept constant along the length [14]. Such a design using a single voltage along the length of a stack is referred to in the literature as a “single electric stage” design [28].
While ED based brine concentration systems for salt production have been used in Japan for around 50 years, such systems have not been deployed widely outside of Japan and South Korea partly due to the high costs involved in concentrating seawater. For reference, Miyake et al. [29] had reported a production cost of 100 $/tonne-salt for an ED-crystallizer system with 65% of the costs being capital costs.

Another membrane based desalination technology that has become very popular now is reverse osmosis (RO). RO is a pressure driven membrane based process that separates saline water into pure product water and a saltier brine stream [30]. Today, RO has become the most widely adopted seawater desalination technology [31]. Seawater RO is also now the most energy efficient desalination technology [9, 32] for seawater desalination. Thermodynamic analysis has also shown conceptually that RO can also be the most energy efficient brine concentration technology for concentrating brine from 150 to 260 g/kg [33]. Conventional seawater RO operates at a pressure of 50-70 bar recovering 40-50% of seawater feed as pure product water with the RO brine discharged back in to the sea at around a salinity of 60-70 g/kg. RO systems for concentrating brine to 260 g/kg do not currently exist in operation due to the high operating pressures required for brine concentration (379.2 bar at NaCl saturation [33]). However, in the past 5 years, advances in RO membrane technology have enabled the development of higher pressure RO systems that can operate up to pressures of 120 bar [34] corresponding to a RO brine salinity of 120 g/kg. These higher pressure RO systems are new and have not yet been adopted widely by the seawater desalination industry.

Some key research questions are: can the cost of seawater brine concentration be reduced below that of current ED costs? If so, by how much can the cost be reduced? Can hybridizing ED with RO reduce costs? And what are the economic implications of reducing brine concentration costs? Can a new salt production industry be created out of small-scale salt production plants that produce vacuum salt through brine concentration and crystallizer systems? Our concept, explained in Section 2, seeks to answers these questions and may inspire further research and development work in this area.

2. Proposed concept: RO-ED hybrid brine concentration systems used with crystallizers for salt production

In this paper, we evaluate potential cost reduction for ED based seawater brine concentration by using ED systems that are hybridized with reverse osmosis (RO). Figure 2a shows an illustration of our proposed RO-ED brine concentration system. Figure 2b shows the detailed flow diagram of the RO, ED and crystallizer systems. Seawater feed (35 g/kg) first flows into an RO system where pure product water and desalination brine are produced. The brine produced by the RO system (60-120 g/kg salinity) is then concentrated in the concentrate channel of the ED system to a salinity of 200 g/kg. To make full use of the RO brine for salt production, all of the RO brine is sent to the concentrate channel for further concentration, with only seawater feed at 35 g/kg used as input for the diluate channel. Simulations developed here were later used
Figure 2: (a) Salt production plant with Reverse Osmosis, Electrodialysis and crystallizer sub-systems. (b) Flow diagram of plant with Reverse Osmosis, Electrodialysis and crystallizer sub-systems.
to confirm that RO-ED hybrid system costs were also minimized when all of the RO brine was sent to the concentrate channel. [35] The concentrate from the ED system is then sent to the crystallizer where it is separated into pure salt and water streams. Because ions are transferred from the diluate stream to the concentrate stream, the ED system discharges a stream more dilute than seawater back into the sea.

2.1. Why hybridize ED with RO for brine concentration?

For increasing the economic feasibility of ED based brine concentration, we considered two paths: reduce system costs and increase revenue. From Fig. ?? we can see that the capital cost of ED contributes the most to the overall cost of salt production. From first principles, ED capital costs scale directly with membrane area and the amount of salt removed by the ED system. Thus, to reduce ED capital costs, the ED membrane area deployed for the transfer of salt must be reduced. Holding the total salt production capacity constant, ED membrane area can be reduced by:

(a) shifting the salt removal load from ED to a more efficient and cost-effective technology

(b) increasing the salt removed per square meter of ED membrane

Both these approaches are evaluated in this paper.

For reducing ED capital costs by shifting some of the salt removal load from ED to an alternative technology, the best alternative today is arguably seawater RO and its variants. Hybridizing RO and ED systems for brackish water desalination (S < 3 g/kg) has already been shown to be lower in cost than standalone ED systems [36]. Similarly, hybrid RO-ED systems have been conceived to treat high salinity produced water from a salinity of 120 g/kg, producing pure water and ED concentrate at 167 g/kg and RO brine at 70 g/kg [37]. Previously, Tanaka et al. [38] have also shown that using RO brine salinity at 88 g/kg as feed in to a ED brine concentration system reduced ED costs by 20 % when compared with using seawater feed. However, Tanaka et al. [38] did not report on the combined cost of the RO-ED system or its economic feasibility and did not analyze the effect of varying the operating pressure of RO.

For reducing ED capital costs by increasing the salt removed per m² of membrane, a direct approach is to increase the operating current density. However, increasing the current density also increases the energy consumed and the operating cost. Tanka et al. [38] had reported that a current density of 300 A/m² was optimal for ED stacks using both seawater feed and desalination feed. However, the optimal current density for ED feed salinities between 70 g/kg and 120 g/kg have not been reported and the dependence of optimal current density on electricity prices has not been discussed.

For increasing revenue over standalone ED brine concentration systems, a simple approach is by producing pure water from the RO sub-system of the RO-ED concept and selling the water. Water prices vary significantly around the world, with water tariffs ranging from 0 to 7.54 $/m³ with a mean global water tariff.
of 1.21 \$/m³\[39\]. Furthermore, salt market prices also vary significantly in the world. To the best of our knowledge, the literature has not addressed how salt and water prices affect the design of RO-ED systems.

Several questions remain unanswered. If the primary application is salt production, does hybridizing ED with RO reduce costs and energy consumption for salt production when compared with standalone ED systems concentrating seawater? Does extending the use of RO to higher pressures and salinities (120 bar and 120 g/kg) increase the economic feasibility of RO-ED systems for seawater brine concentration? How does energy consumption of RO-ED systems vary with increasing hybridization (i.e., increasing the proportion of the salt removal load taken by the RO system)? At what current densities should ED in RO-ED systems be operated for reducing costs? How does the system performance vary with water and electricity prices? Given the varying salt prices around the world, in what locations and scenarios would RO-ED based brine concentration be commercially feasible for salt production? In this paper, we present an answer to these questions through a techno-economic and parametric analysis of RO-ED systems for concentrating seawater from 35 g/kg to 200 g/kg. Parameters studied include: RO brine salinity (i.e., an indicator of amount of hybridization), ED current density, electricity prices, water prices and salt prices. We also discuss the implications of the proposed RO-ED concept for salt production and highlight areas for future research work.

### 3. Methodology

In this section we discuss the techno-economic models used to simulate the RO, ED and crystallizer systems.

#### 3.1. Plant configuration

A standard seawater salinity of \( S_{sw} = 35 \text{ g/kg} \)\[40\] is used as a baseline for this study. Annual salt production is desired to be roughly 100,000 tonnes/year. For simulation purposes, the seawater feed flow rate to the RO system was held constant at 50 m³/hour, corresponding to an annual salt production capacity of 100,000 to 900,000 tonnes/year depending on the degree of RO-ED hybridization. For a standalone ED system, the concentrate inlet flow rate was taken to be 50 m³/hour.

#### 3.2. Reverse osmosis process model

The reverse osmosis model is an adaption of the model described by Thiel et al.\[33\] with performance parameters adapted from Mistry et al.\[32\]. Seawater feed enters the RO system at a pressure of 1 bar. A circulation pump (CP) increases the feed pressure by \( \Delta P_{CP} = 1 \text{ bar} \). The feed stream from the circulation pump, \( \dot{m}_{f,RO} \), is then split in two streams, a feed stream of mass flow rate \( \dot{m}_{p,RO} \) goes to a high pressure pump (HP), and a feed stream having a mass flow rate equal to the brine flow rate \( \dot{m}_{b,RO} \) goes to a pressure exchanger. The HP then pressurizes part of the feed to a pressure of \( P_{HP,RO} \), to overcome the osmotic pressure difference and the pressure losses in the
RO module. The pressure exchanger meanwhile pressurizes the other feed stream to $P_{\text{recover,RO}}$ pressure, which is then passed to a booster pump to attain the pressure $P_{\text{HP,RO}}$. The two feed streams then combine and enter the RO module. The RO module separates the feed into a concentrated brine stream and a pure product stream at a recovery ratio $RR_{\text{RO}}$:

$$RR_{\text{RO}} = \frac{m_{\text{p,RO}}}{m_{\text{f,RO}}}$$

(1)

For a given RO brine salinity ($S_{\text{b,RO}}$), the osmotic pressure at the end of the unit is:

$$\Pi_{\text{RO,b}} = SW_{\text{Osm,Press}}(S_{\text{b,RO}}, t = 25^\circ C)$$

(2)

Here, the osmotic pressure of seawater was determined using seawater osmotic pressure correlations developed by Sharqawy et al. [41] and Nayar et al. [42]. Equation 3 gives the pressure corresponding to feed outlet from the circulation pump:

$$P_{\text{CP,RO}} = P_{\text{atm}} + \Delta P_{\text{CP}}$$

(3)

while, the pressure from the high pressure pump is given by:

$$P_{\text{HP,RO}} = \Pi_{\text{RO,b}} + \Delta P_{\text{pinch,RO}} + \Delta P_{\text{loss,RO}}$$

(4)

where a pinch pressure ($\Delta P_{\text{pinch,RO}}$) of 10 bar and pressure loss ($\Delta P_{\text{loss,RO}}$) of 2 bar is assumed in the RO module. From Mistry et al. [32], the pressure recovered from the pressure exchanger was:

$$P_{\text{recover,RO}} = P_{\text{CP,RO}} + \eta_{\text{PX}} \frac{\rho_{\text{f,RO}}}{\rho_{\text{b,RO}}} (P_{\text{HP,RO}} - \Delta P_{\text{loss,RO}} - P_{\text{atm}})$$

(5)
Efficiencies of the RO pumps ($\eta_{p,RO}$) and of the pressure exchanger ($\eta_{PX,RO}$) were taken to be 0.85 and 0.96 respectively. From these pressures, the work of the pumps could be calculated:

$$W_{CP,RO} = \dot{V}_{f,RO} \Delta P_{CP} \frac{1}{\eta_{p,RO}}$$

(6)

$$W_{HP,RO} = \dot{V}_{f,RO} \frac{\dot{m}_{p,RO}}{\dot{m}_{f,RO}} \Delta P_{HP} \frac{1}{\eta_{p,RO}}$$

(7)

$$W_{BP,RO} = \dot{V}_{f,RO} \frac{\dot{m}_{b,RO}}{\dot{m}_{f,RO}} \Delta P_{BP} \frac{1}{\eta_{p,RO}}$$

(8)

Hence, the total work input to the system is:

$$W_{RO} = W_{CP,RO} + W_{HP,RO} + W_{BP,RO}$$

(9)

The specific energy consumption ($\text{kWh}_e/\text{m}^3$-RO-product) is:

$$E_{\text{total, RO}} = \frac{W_{RO}}{1000[W/kW] \dot{V}_{p,RO,hr}}$$

(10)

where $\dot{V}_{p,RO,hr}$ is the product water produced per hour. The specific energy consumption ($\text{kWh}_e$/tonne-salt) of the RO system normalized to salt produced is:

$$SpE_{RO} = \frac{W_{RO}}{\dot{m}_{salt} \times 3600[s/hr]}$$

(11)

where $\dot{m}_{salt}$ is the salt produced per second. The RO energy costs were obtained from the electricity cost ($\text{Cost}_{\text{elec}}$), and in our analysis an electricity cost of US $0.10/kWh was assumed:

$$OpEx_{RO, energy, hour} = \frac{W_{RO}}{1000[W/kW]} \text{Cost}_{\text{elec}}$$

(12)

Accounting for a capacity factor, $Cap_{\text{fac}} = 0.9$, the annual RO energy cost is:

$$OpEx_{RO, energy, yr} = OpEx_{RO, energy, hour} \times 8760[\text{hr/year}] \times Cap_{\text{fac}}$$

(13)

Cost model for conventional seawater RO

DesalData has extensive data for the specific capital cost of real-world seawater RO desalination plants. The capital cost estimator tool from DesalData accounts for a variety of factors such as the size of the plant, the type of pretreatment, the salinity of the seawater, etc. For the purposes of this paper, we are interested
in obtaining the specific capital cost of RO in the 500-2500 m$^3$/day capacity range. We observed a strong dependence of the specific cost of RO on the capacity of the RO plant. To obtain an accurate estimate of the capital cost of RO, capacity ranges from 250-250,000 m$^3$/day were entered into DesalData’s capital cost estimator tool, with the input parameters being conventional operating conditions and “standard” pretreatment. Figure 4 shows the variation of the specific capital cost of RO with the capacity of the RO system. The specific capital cost of RO was found to vary logarithmically with product water capacity as:

$$\text{SpCapEx}_{\text{RO},p,\text{day}} = 3619 - 201.3 \times \ln (V_{\text{RO},p,\text{day}})$$ (14)

with an $R^2$ value of 0.999996 and absolute percentage deviation from data of 0.06 %.

Thus, the total capital cost for the RO system was:

$$\text{CapEx}_{\text{RO}} = \text{SpCapEx}_{\text{RO},p,\text{day}} \times \dot{V}_{p,\text{RO,day}}$$ (15)

The capacity factor ($\text{Capfactor} = 0.9$), a project life ($t_{\text{life}} = 20$ years) and the rate of return ($r_{\text{return}} = 7\%$) was assumed to calculate the annuity factor of the RO plant:

$$\text{Annuity factor} = \frac{1}{r_{\text{return}}} \left( 1 + \frac{1}{1+r_{\text{return}}} \right)^{t_{\text{life}}}$$ (16)

Thus, the total RO capital cost after annualization was:

$$\text{CapEx}_{\text{RO, yr}} = \frac{\text{CapEx}_{\text{RO}}}{\text{Annuity factor}}$$ (17)

From the RO process model, we had previously calculated the annual RO energy costs. However, operat-
ing expenses also include membrane replacement costs, maintenance costs and labor costs. From the capital

cost data from DesalData [43], we know that RO membranes only contribute to 6.5 % of the total capital
cost of an RO system:

\[
\text{CapEx}_{\text{RO, mem.}} = 0.065 \times \text{CapEx}_{\text{RO}}
\] (18)

Seawater RO membranes have been reported to last 3-7 years [44–47] with an increase in membrane life
observed in recent years with membrane development. The most recent seawater RO plant data we obtained,
from 2013 [47], reported an observed life of almost 7 years even when the plants only used conventional
pretreatment consisting of media filters. For our calculations, we conservatively chose a membrane life of
5 years. Annualizing membrane replacement cost every 5 years we have the annual membrane replacement
cost:

\[
\text{OpEx}_{\text{RO, mem., yr}} = \frac{\text{CapEx}_{\text{RO, mem.}}}{\text{Annuity factor}} \times [(1 + r_{\text{return}})^{-5} +
(1 + r_{\text{return}})^{-10} + (1 + r_{\text{return}})^{-15}]
\] (19)

which translated financially to an annual membrane element replacement rate of 15 %. DesalData [43]
reported that the other operating costs of other maintenance work, chemicals and labor amounted to be
approximately constant for seawater RO systems at 0.03 $/m^3, 0.07 $/m^3 and 0.08 $/m^3 respectively per
amount of RO product water produced. Together these other operating costs amounted to 0.18 $/m^3 of RO
product water. This translated to an annual operating cost contribution of:

\[
\text{OpEx}_{\text{RO, maint., chem., labor, yr}} = 0.18 \frac{\text{$/m^3}}{} \times \dot{V}_{p, \text{RO, yr}}
\] (20)

where \(\dot{V}_{p, \text{RO, yr}}\) is the annual amount of water produced by the RO system in m³/year. Thus, the total
operating cost of RO was:

\[
\text{OpEx}_{\text{RO, yr}} = \text{OpEx}_{\text{RO, energy, yr}} + \text{OpEx}_{\text{RO, mem., yr}} + \text{OpEx}_{\text{RO, maint., chem., labor, yr}}
\] (21)

which brings us to the specific capital cost of RO water normalized to the annual salt production as:

\[
\text{SpCost}_{\text{RO}} = \frac{\text{CapEx}_{\text{RO, yr}} + \text{OpEx}_{\text{RO, yr}}}{\dot{m}_{\text{salt, yr}}}
\] (22)

were, \(\dot{m}_{\text{salt, yr}}\), is the annual amount of salt produced from the RO-ED-crystallizer based salt production
plant in tonnes/year.

To validate our RO cost model, we compared it against water price predictions from Desaldata for the
same input conditions. For the production capacity ranges 250-2500 m$^3$/day of product, our model deviated from DesalData’s estimates by only 4.3-5.5 %. The deviation arose from minor differences from DesalData in calculating energy consumption and the assumed membrane life.

3.3. High pressure (60-120 bar) RO stage process model

Analysis of seawater RO plant operation data from the literature [48] showed that, on average, seawater RO plants around the world operated at a recovery ratio of 42 % with an operating pressure near 60 bar. Practically, extending RO operation beyond 60 bar to 120 bar pressure involves the addition of a high pressure RO (HPRO) stage designed around high pressure RO membranes [34]. In this paper, the reference “RO” refers to the sub-system that includes a conventional SWRO stage and, a HPRO stage when the operating pressure exceeds 60 bar.

The energy required to operate the HPRO stage was calculated using the same process model described for RO in Section 3.2.

For RO operating pressures higher than 60 bar, no datasets on equipment costs were publicly available. However, by looking at the individual cost contributions to a conventional seawater RO plant and by using both engineering estimates and quotes from component manufacturers, we were able to estimate the cost of an HPRO stage (see Appendix A for details). Overall we calculated that for the same recovery ratio, the specific capital cost of a HPRO stage was only 1.09 times that of conventional seawater RO.

The main reason why the HPRO stage cost increase was minor despite the operating pressure being double is that only 29 % of the seawater RO capital cost came from pressure affected components. Components such as pretreatment and intake/outfall were not added to the HPRO stage since they were already accounted for in the preceding conventional seawater RO system. HPRO membrane replacement costs were three times higher due to the higher cost of HPRO membrane elements. Other maintenance, labor and chemical costs were assumed to be the same on a per m$^3$ feed basis as conventional seawater RO.

The costs and energy needs associated with the HPRO stage were added to the conventional SWRO stage and reported together in the results for the RO sub-system.

Scaling potential and need for antiscalants or further pretreatment for HPRO stage:

We further evaluated the scaling potential when concentrating seawater to 120 g/kg using a HPRO stage.

The composition of 120 g/kg brine leaving a HPRO stage was estimated by neglecting salt passage through the membrane and concentrating standard seawater composition [40]. We verified our predicted composition against that predicted by Dow’s ROSA [49] and WAVE [50] simulation software. They were on average within 5 % of each other for all major ions. The scaling potential was evaluated by calculating the saturation indices for possible scale forming salts by running the HPRO brine composition against the Pitzer electrolyte models [51] in the software PHREEQC [52, 53]. There was a clear need for antiscalants due to the potential for both carbonate and sulfate scaling. We worked with an antiscalant company [54] that had considerable
operational experience with seawater desalination to determine the correct antiscalant and dosage needed to prevent scale formation. The divalent ions in the 120 g/kg HPRO brine was determined to be well within the operational limits of basic antiscalants. The associated chemical cost had been accounted for from the specific chemical costs from DesalData. We ascertained that if the right antiscalants were used there was no operational need from a scaling point of view, to carry out additional pretreatment such as using nanofiltration (NF) to enable the addition of a HPRO stage. However, the addition of NF pretreatment maybe considered by salt production plant designers for other reasons such as to produce salt of even higher purity (> 99.8 %), to reduce scaling and maintenance costs in the crystallizer and to increase the plant capacity factor. For the benefit of readers interested in including an NF system, we have reported an NF process model and the costs for NF pretreatment in Appendix B.

3.4. Electrodialysis process model

| Symbol | Value | Reference |
|--------|-------|-----------|
| System inputs | | |
| $S_{c, i, ED}$ | $S_{b, RO}$ | - |
| $S_{c, o, ED}$ | 200 g/kg | - |
| $S_{d, i, ED}$ | $S_{sw} = 35$ g/kg | - |
| Solution Properties | | |
| $D$ | $1.61 \times 10^{-9}$ m$^2$/s | 55 |
| $l_{cu}$ | 0.5 | 55 |
| $v$ | $8.9 \times 10^{-7}$ m$^2$/s | 55 |
| Flow properties/geometry | | |
| $h$ | 0.5 mm | - |
| Membrane parameters | | |
| $\sigma$, Mem$_{eff}$ | $0.64 \pm 0.03$ | 55 |
| $\bar{r}_m$ | $3.5 \times 10^{-4} \pm 1 \times 10^{-4}$ $\Omega$ m$^2$ | 55 |
| $T_{cp}$ | Eq. 30 | 55 |
| $T_{w}$ | Eq. 31 | 55 |
| $L_{cp}$ | Eq. 32 | 55 |
| $L_{w}$ | Eq. 33 | 55 |
| Stack Parameters | | |
| $V_{el}$ | 2.1 V | 55 |

The ED system was modeled by adapting a model developed by McGovern et al. 55, which was itself based on a transport model developed originally by Fidaleo and Moresi 56. In this work, we adapted the model by McGovern et al. 55 for the concentration of brine coming from seawater RO (see Fig. 2b), keeping both the concentrate and diluate channels fully continuous, with the salinities of both channels varying along the length of the ED stack. The salinities at the inlet ($S_{c, i}$) and outlets ($S_{c, o}$) of the concentrate stream are known, with the former being the salinity of the brine leaving the RO system and the latter being 200
g/kg. The value of 200 g/kg for the ED concentrate outlet was chosen to reflect the maximum salinity seen in industrial ED systems [57]. Seawater was directly fed in to the diluate channel of the ED system. All the brine leaving the RO system was assumed to flow into the inlets of the ED concentrate channels. An ED stack can be considered as a collection of $N_{cp}$ identical parallel cell-pairs which consist of an anion exchange membrane, a cation exchange membrane, a diluate channel, and a concentrate channel. Currently operational industrial ED stacks are ‘single stage’ designs [57] where the change in salinity on the concentrate side from 35 g/kg to 200 g/kg happens across a single ED cell-pair. We have assumed the same ‘single stage’ operation for our ED system. For capturing the effect of transport variation along the length of an ED cell-pair, the full length of a single ED cell-pair was discretized in to $N$ computational cells such that each cell sees the same change in salinity on the concentrate side:

$$S_{c,k+1} = S_{c,k} + \frac{S_{c,o} - S_{c,i}}{N-1}$$  \hspace{1cm} (23)

where $S_{c,i}$, $S_{c,o}$ are the salinities at the inlet and outlet of the concentrate stream and $k = 1$ to $N-1$. For

---

Figure 5: Diagram showing an ED stack. The ED plant is assumed to be a single stack with $N_{cp}$ identical parallel cell-pairs.
ED simulations, $N = 50$ was used as it gave grid independent accurate results while keeping the simulations computationally fast. The transfer of salt and water in an ED computation cell is shown in Fig. 6.

The total molar flow rate of salt ($\dot{N}_{s,c,j+1}$) and water ($\dot{N}_{w,c,j+1}$) in the concentrate channels are obtained from the known mass flow rate flowing in to the concentrate channel and the salinity along the concentrate channel:

$$\dot{N}_{s,c,j} = \frac{n_{c,j} S_{c,j}}{1000 \times MW_s}$$

(24)

The net salt flux ($J_{s,j}$) and net water flux ($J_{w,j}$) from the diluate channel in to the concentrate channel can then be related to the total concentrate molar flow rates as:

$$\dot{N}_{s,c,j+1} - \dot{N}_{s,c,j} = A_{cp, tot,j} J_{s,j}$$

(25)

$$\dot{N}_{w,c,j+1} - \dot{N}_{w,c,j} = A_{cp, tot,j} J_{w,j}$$

(26)

where $A_{cp, tot,j}$ is the total cell-pair area discretized to each computational cell.

From data given by an ED manufacturer[58], we set the effective cell-pair area for each membrane pair ($A_{cp}$) to be 0.395 m$^2$. Thus, we can determine the number of cell pairs as:

$$N_{cp} = \frac{\sum_{j=1}^{N} A_{cp, tot,j}}{A_{cp}}$$

(27)

The transport model from McGovern et al. was then applied to each individual cell to obtain the net salt ($J_{s,j}$) and water flux ($J_{w,j}$) going from the diluate to the concentrate in each cell:

$$J_{s,j} = T_{s,j} \cdot \frac{ij}{F} - L_{s,j} \cdot (C_{c,m,j} - C_{d,m,j})$$

(28)
\[ J_{W,j} = T_{w,j} \cdot \frac{i_j}{F} + L_{w,j} \cdot (\pi_{c,m,j} - \pi_{d,m,j}) \]  

(29)

where, \( T_{s,j} \) and \( T_{w,j} \) are the salt and water transport numbers for the ED membrane and \( L_{s,j} \) and \( L_{w,j} \) are the salt and water permeabilities of the membrane. \( C_{c,m,j} \) and \( C_{d,m,j} \) are the molar concentrations of salt in the concentrate and diluate at the surface of the membrane. \( \pi_{c,m,j} \) and \( \pi_{d,m,j} \) are the osmotic pressures of the diluate and concentrate at the surface of the membrane.

McGovern et al. [55] experimentally determined the transport and permeability numbers for a NEOSEPTA AMX and CMX membranes at high salinities. McGovern et al. had expressed the transport and permeability values as simple correlations which are reproduced below:

\[ T_{s}^{cp} = -4 \times 10^{-6} S_d^2 + 4 \times 10^{-5} S_d + 0.96 \pm 0.04 \]  

(30)

\[ T_{w}^{cp} = -4 \times 10^{-5} S_c^2 - 1.9 \times 10^{-2} S_d + 11.2 \pm 0.6 \]  

(31)

\[ L_{s}^{cp} = \min(2 \times 10^{-12} S_d^2 - 3 \times 10^{-10} S_d + 6 \times 10^{-8}, \]  

\[ 2 \times 10^{-12} S_c^2 - 3 \times 10^{-10} S_c + 6 \times 10^{-8}) \pm 6 \times 10^{-9}[m/s] \]  

(32)

\[ L_{w}^{cp} = 5S_c^{-0.416} \pm 2 \times 10^{-5}[mol/m^2 s bar] \]  

(33)

The molar flow rate of salt and water in the diluate in \( 'j+1^{th} \) computational cell is calculated from:

\[ \dot{N}_{s, d, j+1} - \dot{N}_{s, d, j} = -A_{cp, tot, j}J_{s, j} \]  

(34)

\[ \dot{N}_{w, d, j+1} - \dot{N}_{w, d, j} = -A_{cp, tot, j}J_{w, j} \]  

(35)

The salt and water flux across the ED membrane is driven by the voltage applied across an ED cell-pair \( (V_{cp}) \). This voltage is related to the current density by the following expression from McGovern et al. [55]:

\[ V_{cp} = \left( \bar{r}_{am,j} + \bar{r}_{cm,j} + \frac{h_d}{\sigma \Lambda_d C_{d,j}} + \frac{h_c}{\sigma \Lambda_c C_{c,j}} \right) i_j + \frac{\bar{r}_{cm,j}}{N_{cp}} i_j + \frac{2h_r}{\sigma k_r N_{cp, j}} i_j + E_{am,j} + E_{cm,j} \]  

(36)

The expression can be best understood by referring to Fig. 7 which reproduces the equivalent circuit.
diagram of an ED cell-pair from McGovern et al. [55]. Briefly, from left to right, the terms in Eq. 36 represent the resistances of the anion and cation exchange membranes, the diluate and concentrate channel resistances, the resistance of an extra cation exchange membrane at the end of a stack, the resistance of the rinse stream which may be present in stacks and the voltage drops across the anion and cation exchange membranes arising from concentration polarization. The detailed description of each term with the expressions for calculating concentration polarization are given in Ref. [55]. The voltage across the each cell pair can be added together with the voltage across the electrodes to get the total voltage across the ED stack:

\[ V_{\text{ED, stack}} = N_{\text{cp}} V_{\text{cp}} + V_{\text{el}} \]  \hspace{1cm} (37)

The stack power can then be calculated as:

\[ \dot{W}_{\text{ED, stack}} = \sum_{j=1}^{N} i_j A_{\text{cp, j}} (N_{\text{cp}} V_{\text{cp}} + V_{\text{el}}) \]  \hspace{1cm} (38)

The power required for pumping the concentrate and the diluate was calculated using the expressions for friction factor and pressure drop from McGovern et al. [36]:

\[ \dot{W}_{\text{ED, pump}} = \frac{\Delta P_{\text{dil}} \dot{V}_{\text{dil, in}}}{\eta_{\text{p, ED}}} + \frac{\Delta P_{\text{con}} \dot{V}_{\text{con, in}}}{\eta_{\text{p, ED}}} \]  \hspace{1cm} (39)

The total power required to run the ED sub-system was thus:

\[ \dot{W}_{\text{ED}} = \dot{W}_{\text{ED, stack}} + \dot{W}_{\text{ED, pump}} \]  \hspace{1cm} (40)

Normalizing the total power consumed by the ED sub-system to the salt produced, we can obtain the specific
energy consumption (kWh/tonne-salt) of the ED system:

\[ \text{SpE}_{\text{ED}} = \frac{W_{\text{ED}}}{m_{\text{salt}} \times 3600 [\text{hr}]} \] (41)

The hourly energy costs of the ED system can also be obtained as:

\[ \text{OpEx}_{\text{ED, energy, hour}} = \frac{W_{\text{ED}}}{1000 [\text{W/kW}]} \times \text{Cost}_{\text{elec}} \] (42)

Accounting for a capacity factor, \( \text{Cap}_{\text{fac}} = 0.9 \), the annual ED energy cost was obtained as:

\[ \text{OpEx}_{\text{ED, energy, yr}} = \text{OpEx}_{\text{ED, hour}} \times 8760 [\text{hr/year}] \times \text{Cap}_{\text{fac}} \] (43)

The capital costs for the ED system can be calculated from the total membrane area required.

\[ A_{\text{cp, total}} = N_{\text{cp}} \sum_{j=1}^{N} A_{\text{cp, j}} \] (44)

\[ A_{\text{mem, total}} = 2 \times \frac{A_{\text{cp, total}}}{\text{Mem}_{\text{eff}}} \] (45)

Here, the factor 2 accounts for the fact that each cell pair has two membranes and Mem_{eff} accounts for the fact that only a portion of the membrane area is used for transport due to the presence of spacers and gaskets. From conversations with industry \[57\] we know that the specific capital cost of a high salinity ED plant (SpCost_{ED, CapEx, m}) is approximately $ 600/m^2 of membrane. This leads to the total capital costs being:

\[ \text{CapEx}_{\text{ED, total}} = \text{SpCost}_{\text{ED, CapEx, m}} \times A_{\text{mem, total}} \] (46)

Thus the total ED capital cost after annualization is:

\[ \text{CapEx}_{\text{ED, yr}} = \frac{\text{CapEx}_{\text{ED, total}}}{\text{Annuity}_{\text{factor}}} \] (47)

From the ED process model, we had calculated the annual ED energy costs. To obtain the total operating costs, we need to account for costs for membrane replacement, maintenance and labor. From the industry \[57\], we know the membranes cost (SpCost_{ED, mem., m}) to be $ 222/m^2 of membrane. Thus the capital cost for one set of ED membranes for an ED system will be:

\[ \text{CapEx}_{\text{ED, mem.}} = \text{SpCost}_{\text{ED, mem., m}} \times A_{\text{mem, total}} \] (48)
Monovalent selective ED membranes have a reported life of 5-10 years. From our conversations with ED manufacturers and plant operators, we chose an average life of 7 years for our cost model. Annualizing this cost of replacing ED membranes every 7 years we have the annual membrane replacement cost:

\[
\text{OpEx}_{\text{ED, mem., yr}} = \frac{\text{CapEx}_{\text{ED, mem.}}}{\text{Annuity factor}} \times \left[ (1 + r_{\text{return}})^{-7} + (1 + r_{\text{return}})^{-14} \right]
\]  

(49)

ED brine concentration systems for salt production typically have one full-time operator. The average annual salary for a plant operator in the USA is $50,000 [59, 60]. This was assumed to be the annual labor cost from the ED system (\(\text{OpEx}_{\text{ED, labor, yr}}\)). Other maintenance costs and chemical costs were obtained by converting data from brackish water ED plants in the literature [59] by the method previously used by McGovern et al. [55]. Sajtar and Bagley had reported the specific cost of other maintenance and chemicals to be 0.0285 and 0.007 $/m³ of feed. Using the model from McGovern et al. [55] a brackish ED plant would need 0.39 m² of cell pair area per m³/day of feed capacity. Using a membrane effectiveness of 0.64 and accounting for 2 ED membranes per cell pair, this translates to 1.22 m² of total ED membrane area per m³/day of feed capacity. Assuming 365 days of operation, the maintenance and chemical costs can be converted from per $/m³ of feed to per m² of total ED membrane area per year. Thus, the specific annual cost of maintenance (\(\text{SpCost}_{\text{ED, maint., m}}\)) and chemicals (\(\text{SpCost}_{\text{ED, chem., m}}\)) was calculated to be 8.5 and 2.1 $/m²-year. We can then get the annual maintenance and chemicals costs for the ED system and the total annual ED operating expenses:

\[
\text{OpEx}_{\text{ED, maint., chem., yr}} = (\text{SpCost}_{\text{ED, maint., m}} + \text{SpCost}_{\text{ED, chem., m}}) \times A_{\text{mem, total}}
\]  

(50)

\[
\text{OpEx}_{\text{ED, yr}} = \text{OpEx}_{\text{ED, energy, yr}} + \text{OpEx}_{\text{ED, mem., yr}} + \text{OpEx}_{\text{ED, labor, yr}} + \text{OpEx}_{\text{ED, maint., chem., yr}}
\]  

(51)

which brings us to the specific cost of ED normalized to salt production as:

\[
\text{SpCost}_{\text{ED}} = \frac{\text{CapEx}_{\text{ED, yr}} + \text{OpEx}_{\text{ED, yr}}}{m_{\text{salt, yr}}}
\]  

(52)

Values of inputs and constants used in the ED model are given in Table [1].
3.4.1. Validation of Electrodialysis model

Laboratory scale validation:

The ED model presented here was validated against experiments carried out by McGovern et al. [55].

![Figure 8: Experimental data on the diluate and concentrate salinities corresponding to stages from [55]](image)

Figure 8: Experimental data on the diluate and concentrate salinities corresponding to stages from [55]

![Figure 9: ED model presented in this work validated against experimental data from [55] for (a) Specific process time and (b) Specific energy. Note that the specific process time is a proxy for ED membrane area](image)

Figure 9: ED model presented in this work validated against experimental data from [55] for (a) Specific process time and (b) Specific energy. Note that the specific process time is a proxy for ED membrane area

(Figure 8). Each stage shown in Fig. 8 represents a pairing of diluate and concentrate salinities ranging from 0 to 225 g/kg. McGovern et al. experimentally determined the specific process time, a proxy for the membrane area required to treat a given quantity of brine and the specific energy required for treatment. The model described in this work was used to simulate each stage test (Fig. 9a and Fig. 9b). As can be seen from these figures, the model described here closely matched the results from [55]. The average absolute percentage deviation between the model and the experimental data was 13% for specific process time and 11% for specific energy.
Industrial stack scale validation:

To further validate the ED model, we also compared the results predicted by our model against results from an industrial stack [57]. The variation of concentrate and diluate salinities along the length of a single ED membrane is shown along with the model inputs parameters and output area is shown in Fig. 10. The model took in input conditions from data given by the ED manufacturer, Asahi Glass and Chemicals Co., for an ED stack which had an effective cell pair area of 24 m\(^2\). The industrial stack data used as input conditions for the model were: concentrate salinity in the stack varying from 39 g/kg to 177 g/kg (corresponding to a concentration variation from 40 g/L to 200 g/L), an inlet concentrate flow rate of 59 L/hr, a total stack inlet flow rate of 1000 L/hr, a diluate to concentrate flow rate ratio at the outlet of the stack (Dil-con-ratio\(_o\)) of 16 and a current density of 250 A/m\(^2\). For a diluate to concentrate flow rate ratio at the stack inlet (Dil-con-ratio\(_i\)) of 80, the model matched the reported value for Dil-con-ratio\(_o\) of 16. At this condition, the model predicted the required cell pair area to be 22.7 m\(^2\) matching the actual cell pair area of the industrial stack to within 5.4%. We were also able to verify the predicted membrane area against a full installation of ED stacks. For producing 100,000 tonne/year of salt, our predicted membrane area matched that reported by the manufacturer [57] to within 11.8%.

Since the dataset from the manufacturer did not report energy consumption, the energy consumption predictions of the model was separately verified against industrial stack data from the literature [16, 61, 62]. Tanaka [16] had reported that an industrial ED stack operated at 266 A/m\(^2\) producing a concentrate at 174 g/L of NaCl (at 90% purity) had an energy consumption of 165 kWh/tonne. Simulating the same conditions, our model reported an energy consumption of 172 kWh/tonne, within 4.7%.

The ED process model was thus verified to be accurate against both laboratory scale and industrial scale data.
3.5. Crystallizer

The crystallizer component of the system separates out salt from high salinity brine. The crystallizer has the subcomponents: an evaporator, a de-watering unit, and a dryer to produce salt ready for shipment. The critical subcomponent, which consumes the most energy and is most expensive, is the evaporator. Evaporators for “vacuum salt production” are either steam driven multi-effect evaporators or electricity driven mechanical vapor recompression (MVR) systems [63]. Both types of evaporators require similar amounts of primary energy at around 450 kWh-thermal/tonne-salt [63]. For this paper, we assume an electric MVR evaporator for the crystallizer. From conversations with the industry [64], we obtained a generic capital cost figure of $15 million for a MVR based crystallizer system that can produce 100,000 tonnes of salt, and an average energy consumption number of 150 kWh/tonne of salt. These crystallizer systems typically take in nearly saturated brine at 250 g/kg, as opposed to 177-200 g/kg brine delivered by conventional industrial ED systems. Thus the capital cost and operating cost numbers need to be adjusted.

In a crystallizer, the feed gets separated into to salt, pure product water and a purge stream:

\[ \dot{m}_{f,Crys.} = \dot{m}_{p,Crys.} + \dot{m}_{salt} + \dot{m}_{purge} \]  

\[ \dot{m}_{f,Crys.} \frac{S_{f,Crys.}}{1000 \text{ g/kg}} = \dot{m}_{salt} + \dot{m}_{purge} \frac{S_{purge}}{1000 \text{ g/kg}} \]  

where,

\[ \dot{m}_{f,Crys.} = \dot{m}_{c,o,ED} \]  

For our calculations we assumed that the purge stream is saturated with a salinity of 250 g/kg:

\[ S_{purge} = 250 \text{ g/kg} \]
The ratio of the purge stream to the feed stream is given by the purge ratio (PR),

$$PR = \frac{\dot{m}_{\text{purge}}}{\dot{m}_{\text{t, Crys.}}}$$

which is determined by the composition of the feed and the required purity of the salt. For salt production from solution mining, the feed coming in to the crystallizer is typically saturated with sodium chloride and has about 8 g/L of impurities of potassium and sulfate ions. Lime and soda ash are used to reduce the levels of calcium and magnesium to levels on the order of mg/L leaving only potassium and some residual sulfates behind.

At typical evaporator operating temperatures, potassium chloride is more soluble than sodium chloride and the purge ratio is used to control the potassium concentration in the salt stream. A high purge ratio leads to a lower build-up of potassium and sulfates in the purge stream which directly leads to a more pure salt stream. For conventional solution mining based salt production, the purge ratio (PR_{250 g/kg}) is around 15 % for table salt production and 20-40 % for chloralkali grade vacuum salt [64], reflecting the degree of purity needed for each application. Table salt only needs to be at least 97 % pure [65] while chloralkali grade vacuum salt must be at least 99.9 % pure [66].

The purge ratio of the crystallizer in a ED-crystallizer or RO-ED-crystallizer is however different from the purge ratio of a crystallizer taking in solution mined brine as feed. The salinity of the ED concentrate (200 g/kg) is lower than that from solution mining (250 g/kg). Furthermore, the amount of potassium and sulfate in the ED concentrate is lower than that in solution mined brine. The composition of a typical ED concentrate stream was reported by an ED manufacturer to have 190 g/L of sodium chloride, 1.2 g/L of calcium and magnesium ions and 0.24 g/L of sulfate ions [58].

Since the monovalent selective ED membranes concentrate potassium in a similar way as sodium, from the amount of sodium we estimated the potassium in the ED concentrate. By our estimates, sodium chloride represented 93.2 % by weight of the salts in the ED concentrate. We then increased the concentration of the ED concentrate until the saturation limit of sodium chloride was reached and calculated the concentration of potassium and sulfate ions to be 4.8 g/L. This was 40 % lower than the impurity level in solution mined brine. Thus, the purge ratio for a crystallizer using ED concentrated seawater can be 40% lower than that in conventional solution mining. For our concept of using an RO-ED-crystallizer, the effective purge ratio was:

$$PR = PR_{250 g/kg} \times \frac{200 g/kg}{250 g/kg} \times \frac{4.8 g/L}{8 g/L} = 0.096$$

The approach above to calculate PR was validated by a crystallizer manufacturer [64]. We know the capital cost of a crystallizer taking in conventional solution mined brine as feed. What governs the capital cost is the evaporation rate of the crystallizer. The evaporation rate of a fixed crystallizer size is constant
regardless of the salinity of the feed entering the crystallizer.

\[
\dot{m}_{\text{evap, Crys.}} = \dot{m}_{\text{p, Crys.}} = \dot{m}_{\text{salt}} \left[ \frac{1000 \text{ g/kg}}{S_f, \text{Crys.}} \right] \times \frac{1 - \text{PR}}{1 - \text{PR} \frac{S_{\text{purge}}}{S_f, \text{Crys.}}} - 1 \text{ = constant} \tag{59}
\]

The amount of salt produced by the crystallizer varies with the salinity of the feed entering the crystallizer. This leads to the specific capital costs and operating costs increasing inversely with the salt production rate. The increase in specific cost or decrease in salt production for a crystallizer using a feed different from that in solution mining is captured by the crystallizer salinity scale-up factor (CSSF):

\[
CSSF = \frac{\dot{m}_{\text{salt, 250 g/kg}}}{\dot{m}_{\text{salt}}} = \frac{\left( \frac{1000 \text{ g/kg}}{S_f, \text{Crys.}} \right) \times \frac{1 - \text{PR}}{1 - \text{PR} \frac{S_{\text{purge}}}{S_f, \text{Crys.}}} - 1}{\left( \frac{1000 \text{ g/kg}}{250 \text{ g/kg}} \right) \times \frac{1 - \text{PR}250 \text{ g/kg}}{1 - \text{PR}250 \text{ g/kg} \frac{S_{\text{purge}}}{S_f, \text{Crys.}}} - 1} \tag{60}
\]

For feed salinity of 200 g/kg, PR_{250 g/kg} of 0.2 and PR of 0.096, CSSF was 1.38. Accounting for the plant capacity factor, the annual salt produced in tonnes is:

\[
\dot{m}_{\text{salt, yr}} = \dot{m}_{\text{salt}} \times \frac{86400 \text{ s/day} \times 365 \text{ day/yr}}{1000 \text{ kg/tonne}} \times \text{Capfac} \tag{61}
\]

Thus, the capital expense of the crystallizer is

\[
\text{CapEx}_{\text{Crys.}} = \frac{15,000,000}{100,000 \text{ tonne/yr}} \times \dot{m}_{\text{salt, yr}} \times CSSF \tag{62}
\]

and the specific capital cost of the crystallizer is:

\[
\text{SpCapEx}_{\text{Crys.}} = \frac{\text{CapEx}_{\text{Crys.}}}{\dot{m}_{\text{salt, yr}} \text{Annuityfactor}} \tag{63}
\]

The specific energy consumption of the crystallizer adjusted for the lower feed salinity is:

\[
\text{SpE}_{\text{Crys.}} = 150 \text{ kWh/tonne} \times CSSF \tag{64}
\]

leading to a specific energy cost for the crystallizer of:

\[
\text{SpOpEx}_{\text{Crys., energy}} = \text{SpE}_{\text{Crys.}} \times \text{Costelec} \tag{65}
\]

Crystallizer systems of the capacity we are considering typically have two full-time plant operators and a process chemist to ensure quality of the salt being produced [64]. The average salary for a plant operator in
the USA is $50,000 \text{ [60] while that of a process chemist is $65,000 [67]} bringing the annual labor cost of the crystallizer system (\text{OpEx}_{\text{Crys., labor, yr}}) to $165,000. The cost of replacing parts, maintenance and chemicals for the crystallizer system was obtained from the literature \text{ [68] from costs reported for a zero liquid discharge system for treating 1 million gallons per day of feed (3785 m}^3\text{/day). Normalizing these costs per unit feed, the specific annual cost of parts, maintenance and chemicals was 273, 164 and 73 $/m^3 of feed. Together, the specific annual cost of overall maintenance and chemicals for the crystallizer (\text{SpCost}_{\text{Crys., maint., chem. m}^3\text{/day}}) was 510 $ \text{-day/m}^3 of feed to the crystallizer, resulting in an annual cost of parts, maintenance and chemicals of:

\[
\text{SpOpEx}_{\text{Crys., tot.}} = \text{SpOpEx}_{\text{Crys., energy}} + \frac{\text{OpEx}_{\text{Crys., maint., chem. yr}} + \text{OpEx}_{\text{Crys., labor, yr}}}{\dot{m}_{\text{salt, yr}}} \times \dot{V}_{i, \text{Crys., m}^3\text{/day}} \times 365 \text{ day/yr} \times \text{Cap \_factor} \tag{66}
\]

Thus, the total specific operating cost of the crystallizer was:

\[
\text{SpCost}_{\text{Crys.}} = \text{SpCapEx}_{\text{Crys.}} + \text{SpOpEx}_{\text{Crys., tot.}} \tag{67}
\]

The cost predictions of the crystallizer model described above was independently verified with an ED manufacturer \text{ [57].}

3.6. Overall cost and energy required for RO-ED-crystallizer based salt production system

From the specific cost expressions for the RO system, the ED system and the crystallizer, we can obtain a total specific cost of the overall system:

\[
\text{SpCost}_{\text{REC}} = \text{SpCost}_{\text{RO}} + \text{SpCost}_{\text{ED}} + \text{SpCost}_{\text{Crys.}} \tag{69}
\]

It must be noted that specific cost of the overall system does not include the cost of packaging, iodizing and storage facilities. From a ED-crystallizer plant operator, we know that the purity of the salt produced is at least 99.8 %. The overall specific cost reported in this paper is for producing 99.8 % pure salt.

The water production from the RO system reduces the overall cost of production of salt. The revenue from the water produced by the RO system can be normalized to obtain a specific revenue in terms of salt
specific revenue per tonne of salt produced:

\[ SpRev_{REC} = V_{p,RO,yr} \times Price_{water} \] (70)

The overall specific energy required for the RO-ED-crystallizer was:

\[ SpE_{REC} = SpE_{RO} + SpE_{ED} + SpE_{Crys.} \] (71)

**Overall cost validation:**

The RO specific cost model was validated against DesalData. The ED-crystallizer specific cost model was validated against production cost figures reported in the literature [29]. Miyake et al. [29] had reported the production cost from a 250,000 tonne/year ED-crystallizer plant to be 100 $/tonne-salt. The operating current density was given to be 300 A/m² while the electricity price reported was 5 cents/kWh. Since project life and a rate of return was not reported, we assumed a 20 year project life and factored in the lower interest rates in Japan and assumed a rate of return of 5%. For these inputs, our model reported a production cost of 109 $/tonne-salt coming within 9% of the figure reported by Miyake et al. [29]. Thus, the overall RO-ED-crystallizer specific model was verified to be reasonably accurate.

### 3.7. Salt and brine purity

In this paper, the composition of feed seawater we used for our analysis was the reference composition of standard seawater from Millero et al. [40]. The proportion of sodium chloride in seawater by weight is 77.9%. Concentration of seawater by RO does not change the purity of brine significantly. However, the ED system with monovalent selective membranes increases the amount of sodium chloride in the brine during the concentration process. Based on public data reported by an ED manufacturer [58], a standard ED brine concentrator with monovalent selective membranes concentrates sodium chloride such that the final ED concentrate has 190.1 g/L of sodium chloride, 1.2 g/L of calcium ions, 1.2 g/L of magnesium ions and 0.24 g/L of sulfate ions. The corresponding proportion of sodium chloride in the ED concentrate was by our calculation was 93.2%. The crystallizer with its purge ratio further separates sodium chloride from other constituents in the brine which leaves through the purge stream. Once the salt is dried, the purity of the dry salt obtained from an ED-crystallizer is at least 99.8% pure. The salt produced by the RO-ED-crystallizer is also expected to have the same degree of purity. It must be noted that the purity of the salt produced is not sufficient for direct use in an electrolyzer for chloralkali production. Further purification of the produced wet salt is needed for use in chloralkali production.
4. Parametric design of hybrid RO-ED systems for brine concentration for salt production

This section discusses the effects of hybridizing RO with ED, ED current density on RO-ED design, electricity prices and water prices.

4.1. Hybridizing RO with ED reduced costs and energy required for brine concentration for salt production

![Bar chart showing specific cost of RO-ED systems with increasing RO hybridization]

Figure 12: Specific cost of RO-ED systems decreases with increasing RO hybridization (i.e., with increasing ED concentrate inlet salinity).

Hybridizing RO with ED all the way up to 120 g/kg RO brine salinity (i.e., “fully hybridizing”) while operating the ED system at conventional operating conditions (described later in the paragraph) reduced the specific cost of brine concentrations by 7 %. Figure 12 shows the specific cost of the RO-ED brine concentration system varying with RO hybridization. Hybridization with RO is represented on the $x$-axis by the salinity of the concentrate inlet to the ED system (salinity of brine leaving RO). The leftmost column corresponding to a 35 g/kg ED concentrate inlet salinity represented a standalone ED system with the other columns representing RO-ED hybrid systems. For each of the RO-ED configurations, the ED current density was fixed at 300 A/m$^2$, the cost of electricity was 0.1 $/\text{kWh}$, the Dil-con-ratio$_i$ selected was fixed at 80, and the concentrate salinity leaving the ED stack was fixed at 200 g/kg. The specific cost of concentrating brine from 35 g/kg to 200 g/kg using a standalone ED system was $89/\text{tonne-salt}$ while an RO-ED hybrid system with RO brine at 120 g/kg cost only $82/\text{tonne-salt}$ — a cost reduction of 7 %. In comparison, partial hybridizing of RO with ED by stopping at the limits of operation of conventional seawater RO (i.e., RO brine salinity of 60 g/kg) cost $87/\text{tonne-salt}$ — a cost reduction of only 2 %.

Hybridizing RO with ED all the way up to 120 g/kg RO brine salinity while operating the ED system at conventional operating conditions reduced the specific energy consumption for brine concentration by 13 %. Figure 13 shows the specific energy consumption of RO-ED brine concentration varying with RO hybridization. Just as in Fig. 12, the degree of hybridization is captured in the $x$-axis by the salinity of
the concentrate inlet to the ED system. The parameters held constant in each configuration in Fig. 13 are identical to the ones described in Fig. 12. The leftmost column corresponded to a standalone ED system. Each column shown in the Fig. 13 represents an RO-ED hybrid configuration, with the parameters held constant being the same as in Fig. 12. While a standalone ED system needed 219 kWh/tonne-salt to concentrate brine from 35 g/kg to 200 g/kg, a RO-ED hybrid system with RO brine at 120 g/kg consumed only 191 kWh/tonne-salt for the same brine concentration load. The reduction in specific energy consumption seen in RO-ED hybrid systems with increasing hybridization with RO highlights that RO is more energy efficient than ED at concentrating brine from 35 g/kg to 120 g/kg when the current density in the ED system is 300 A/m².

Figure 13: Specific energy consumption of RO-ED systems reducing with increasing RO hybridization represented by increasing ED concentrate inlet salinity.

Figure 14: Specific costs, of an RO-ED system with an RO brine salinity of 120 g/kg, broken down across ED and RO capital and operating costs.
**ED capital and membrane costs contributed the most to RO-ED specific costs**

The distribution of the specific cost of an RO-ED system with an RO brine salinity of 120 g/kg across the capital and operating costs of the RO-ED system is shown in Fig. 14. The individual cost contributions are labeled in the figure as well. ED capital costs contributed the most at $38/tonne-salt corresponding to 47% of the total cost followed by the operating cost which was $40/tonne-salt corresponding to 49% of the total cost. The operating costs consisted of $18/tonne-salt in energy costs, $14/tonne-salt in membrane replacement costs and $8/tonne-salt for other maintenance, chemical and labor expenses. If the ED membrane replacement costs are combined with the initial ED capital costs, they together amount to $53/tonne-salt or 64% of the total cost. RO as a whole contributed to only around $4/tonne-salt corresponding to 5% of the total cost.

**RO was more cost-effective than ED at concentrating seawater**

While RO contributed only 5% to the total specific cost of an RO-ED system with 120 g/kg RO brine salinity, the RO sub-system was responsible for 52% of the total salinity increase, concentrating seawater from 35 g/kg to 120 g/kg. However, since RO and ED systems work differently, the salinity change does not directly equate to salt transferred. In an RO system, strictly speaking no salt transfer occurs, only water transfer; in an ED system, both salt and water are transferred from the diluate to the concentrate. To effectively compare the brine concentration abilities of the two systems, we imagined replacing the RO system with an “ideal blackbox salt concentrator” that increases salinity while keeping the total mass flow rate constant. The “ideal blackbox salt concentrator” delivered the same output brine of salinity $S_{b,RO}$ and mass flow rate $\dot{m}_{b,RO}$ from seawater with the same feed salinity $S_{f,RO}$ and assumed unchanged mass flow rate $\dot{m}_{b,RO}$. Thus, the “effective salt transferred” by the “ideal blackbox salt concentrator” that replaced the RO system was:

$$\Delta \text{Salt}_{eff, RO} = \dot{m}_{b,RO} (S_{b,RO} - S_{f,RO})$$ (72)

The “effective salt transferred” by the ED system was the same as the actual salt transferred by it:

$$\Delta \text{Salt}_{eff, ED} = \Delta \text{Salt}_{ED} = \dot{m}_{c,o,ED} S_{c,o,ED} - \dot{m}_{c,i,ED} S_{c,i,ED}$$ (73)

Adding the effective salt transferred by the RO and ED systems we have the total effective salt transferred as:

$$\Delta \text{Salt}_{eff, RO-ED} = \dot{m}_{c,o,ED} S_{c,o,ED} - \dot{m}_{c,i,ED} S_{f,RO}$$ (74)

The percentage of the “effective salt transferred” in the RO and ED systems along with the percentage salinity change from each system is shown in Fig. 15. While RO contributed to 52% of the salinity change, it only contributed to 9% of the effective salt transferred in the brine concentration process. Even with the
RO system providing brine at 120 g/kg for the ED system to concentrate to 120 g/kg, 91 % of the salt in the ED concentrate outlet was transferred by the ED system from the ED diluate stream.

Figure 15: Percentage of salinity change and salt transferred in RO and ED sub-systems for a RO-ED system with RO brine at 120 g/kg salinity

To effectively compare the cost-effectiveness of RO and ED sub-systems, we normalized the specific cost of the RO and ED sub-systems to the effective salt transferred in each sub-system:

\[ \text{NormSpCost}_{RO} = \frac{\dot{m}_{salt}}{\Delta \text{Salt}_{eff, RO}} \times \text{SpCost}_{RO} \]  
\[ \text{NormSpCost}_{ED} = \frac{\dot{m}_{salt}}{\Delta \text{Salt}_{eff, ED}} \times \text{SpCost}_{ED} \]

Figure 16 shows the salt transfer normalized specific cost of the RO and ED sub-systems for an RO-ED system operating with a RO brine salinity of 120 g/kg. While the ED system had a normalized specific cost of $78/tonne-salt, the RO system had a normalized specific cost of $37/tonne-salt. This effectively meant that on a “effective salt transferred” basis, ED was twice as expensive as RO. To confirm this, we used our models and simulated the cost of producing a fixed amount of 120 g/kg brine from seawater, using first an RO system and then an ED system. Our simulations again showed that RO could produce 120 g/kg brine from seawater at around half the cost of an ED system producing the same amount of brine.

**RO was more energy efficient than conventional ED \((i = 300 \text{ A/m}^2)\) at concentrating seawater**

To effectively compare the effective energy use of the RO and ED sub-systems, we used the same approach used to compare cost-effectiveness. The normalized specific energy consumption of the RO and ED sub-systems was obtained by normalizing the specific energy consumption to the effective salt transferred:

\[ \text{NormSpE}_{RO} = \frac{\dot{m}_{salt}}{\Delta \text{Salt}_{eff, RO}} \times \text{SpE}_{RO} \]  
\[ \text{NormSpE}_{ED} = \frac{\dot{m}_{salt}}{\Delta \text{Salt}_{eff, ED}} \times \text{SpE}_{ED} \]
Figure 16: Normalized specific cost of RO and ED sub-systems, in an RO-ED system with a RO brine salinity of 120 g/kg. The specific cost was normalized per tonne of salt transferred in each sub-system.

\[
\text{NormSpE}_{\text{ED}} = \text{SpE}_{\text{ED}} \times \frac{\dot{m}_{\text{salt}}}{\Delta S_{\text{eff, ED}}} \tag{78}
\]

For an RO-ED system operating with a RO brine salinity of 120 g/kg and at a ED current density of 300 A/m², the ED system had a normalized specific energy consumption of 182 kWhₑ/tonne-salt while the RO system had a normalized specific energy consumption of only 85 kWhₑ/tonne-salt. Thus, on “effective salt transferred” basis, RO was slightly more than twice as energy efficient as conventional ED \((i = 300 \text{ A/m}^2)\) at concentrating seawater. This is why hybridizing ED with RO led to reductions in the specific energy consumption seen in Fig. 13.

4.2. Revenue from co-production of water increased cost-effectiveness of hybridizing RO with ED by 1-6 %

Figure 17: Variation of specific revenue from water production at different water prices

Unlike a standalone ED system which only produces 200 g/kg brine as the useful product, RO-ED based
Table 2: Production cost of water at various desalination plants across the globe

| Country, City        | Year of commissioning | Capacity (m³/day) | Cost ($/m³) | Reference |
|----------------------|-----------------------|-------------------|-------------|-----------|
| Singapore            | 2005                  | 136000            | 0.45        | 69        |
| Israel, Soreq        | 2013                  | 410000            | 0.52        | 70        |
| Israel, Ashkelon     | 2005                  | 296000            | 0.51        | 70        |
| USA, Tampa           | 2004                  | 86000             | 0.68        | 70        |
| Australia, Perth     | 2006                  | 126000            | 0.75        | 70        |
| Kuwait               | –                     | 1128              | 0.84        | 71        |
| Saudi Arabia, Yanbua | 2011                  | 127800            | 1.03        | 72        |
| India, Chennai       | 2010                  | 100000            | 1.03        | 73        |
| Japan, Fukuoka       | 2005                  | 50000             | 1.9         | 74        |

Brine concentration systems also produce pure water, a useful byproduct. Water prices and the cost of water production have a wide distribution. Research conducted by IBM [75] reported that the real cost of water production after factoring in water wastage, varied from 0.34 $/m³ in Manila, Philippines to 2.26 $/m³ in London, United Kingdom with a global weighted index cost of 1.31 $/m³. Zetland and Gasson [39] analyzed water tariffs from 308 cities around the world and reported that water tariffs ranged from 0 to 7.54 $/m³ with a global average water tariff of 1.21 $/m³. Table 2 further gives the cost of water produced at various desalination plants around the world. To account for the large spread of water costs and water prices around the world, we calculated the potential revenue that can be generated from co-production of water from an RO-ED brine concentration system at water prices ranging from 0.25 $/m³ to 2 $/m³. Figure 17 shows the water revenue from a RO-ED brine concentration system normalized to the quantity of salt produced, with the water revenue also represented as a percentage of standalone ED brine concentration costs. The RO-ED configuration simulated here had an RO brine salinity of 120 g/kg, with other parameters same as in Section 4.1. Thus, depending on the geographic location, co-production of water makes an RO-ED system more cost-effective than a standalone ED system by 1 % to 6 %. When the revenue from the RO system is coupled with the cost reduction of 7 % reported in Section 4.1 RO-ED hybrid systems can be more cost-effective than a standalone ED system by 8 % to 13 %.

4.3. Optimizing ED current density further reduced RO-ED brine concentration costs

From Fig. 14 we know that for a fully hybridized RO-ED system (i.e., RO brine salinity at 120 g/kg), the biggest contributor to the cost is the capital cost and the membrane replacement cost of the ED subsystem. From Fig. 14 it can be further seen that the energy cost was only 34 % of the capital and membrane replacement cost, implying that there was room for optimizing the trade-off between capital and energy costs. As discussed in Section 2 the capital cost can be reduced by increasing ED current density which increases the salt removed per m² membrane and reduces the membrane area required for a fixed salt production capacity.
Figure 18: Variation of specific cost of ED and RO-ED systems with current density

Figure 18 shows how ED current density affects the specific cost of a standalone ED system and two RO-ED systems, where one RO system operates like conventional seawater RO producing brine at 60 g/kg and the other RO system is fully hybridized operating at the highest pressure of around 120 bar corresponding to a RO brine salinity of 120 g/kg. The other parameters had the same values reported in Section 4.1. For reference, the electricity cost used for this result was $0.1/kWh. It can be seen that from a specific cost point of view, the optimal current density of operation is 600 A/m². For an RO-ED system with RO brine salinity of 120 g/kg, the specific cost reduced by 27% when the current density was increased from 300 A/m² to 600 A/m². From 300 A/m² to 600 A/m², such a sharp reduction in specific cost was seen because the reduction in membrane area reduced the ED capital and membrane replacement costs far more than the associated increase in energy costs. Beyond 600 A/m², the specific costs start increasing as the energy cost begins to dominate overall costs.

Figure 19: Variation of specific energy of ED and RO-ED systems with current density

Although the optimal current density from a specific cost point of view was 600 A/m², the current density
that minimized the specific energy consumption was around 300 A/m$^2$. This difference is highlighted in Fig. 19 which captures the variation of specific energy with ED current density for the systems shown in Fig. 18. Tanaka et al. [38] had previously reported that for both a standalone ED system and an RO-ED system where the RO system was a conventional seawater RO system with a brine outlet of around 60 g/kg, the current density that reduced the specific energy consumption was 300 A/m$^2$. While the RO-ED configurations in this paper used seawater as a diluate input and Tanaka et al. used RO brine as diluate input, our findings on optimal energy consumption match. We found that 300 A/m$^2$ is the ED current density that minimized specific energy consumption for both a standalone ED system and an RO-ED system with RO brine salinity of 60 g/kg (i.e. conventional seawater RO). For a fully hybridized RO-ED system with a RO brine salinity of 120 g/kg, the ED current density that minimized specific energy consumption was slightly less, at 280 A/m$^2$.

4.4. Effect of electricity prices on optimal ED current density

![Figure 20: Variation of specific cost with current density at different electricity prices](image)

The optimal ED current density that minimizes the specific cost of RO-ED sub-systems is also dependent on the electricity prices. In this paper, we have selected an electricity driven crystallizer which is also affected by electricity prices. Figure 20 shows the variation of specific cost for the RO-ED-crystallizer system with ED current density and electricity prices. The RO-ED configuration represented here is a fully hybridized RO-ED system with RO brine salinity of 120 g/kg. The lowest electricity price shown here of 1.6 US cents/kWh corresponds to the industrial electricity price in Kuwait, where electricity is subsidized. From Fig. 20, we can see that the optimal ED current density that minimizes the total specific cost decreases with increasing electricity price. For electricity prices of 1.6, 5, 10, 15 and 20 US cents/kWh, the optimal ED current densities correspondingly were 1350, 800, 600, 500 and 450 A/m$^2$. 

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5. Optimal RO-ED hybrid designs significantly reduced the cost of brine concentration and salt production

![Specific costs of RO-ED systems optimized for various global cases with percentage reduction in costs compared to standalone ED case.](image)

**Figure 21:** Specific costs of RO-ED systems optimized for various global cases with percentage reduction in costs compared to standalone ED case.

**Table 3:** Cost reducing steps and the associated specific cost reductions in RO-ED systems corresponding to cases in Fig. 21. The last column shows the cumulative percentage reduction from the baseline case caused with each cost reducing step.

| System                      | Cost reducing step                      | SpCost ($/tonne) | ΔSpCost ($/tonne) | ΔSpCost (%) | ΔSpCost total (%) |
|-----------------------------|----------------------------------------|------------------|-------------------|-------------|------------------|
| Standalone ED (global)      | Baseline case                          | $89              | $-                | 0%          | 0%               |
| Standalone ED (global)      | Optimizing i density                    | $66              | $23               | 26%         | 26%              |
| RO-ED (global)              | Fully hybridizing with RO to 120 g/kg   | $60              | $6                | 7%          | 32%              |
| RO-ED (5 c/kWh_e)           | Optim. for 5 c/kWh_e                    | $43              | $17               | 19%         | 52%              |
| RO-ED (Kuwait)              | Optim. for 1.6 c/kWh_e                  | $27              | $16               | 18%         | 70%              |
|                            | Including water revenue                 |                  |                   |             |                  |
| RO-ED (Kuwait)              | Water price at $ 0.5/m³                  | $25              | $1                | 2%          | 71%              |
| RO-ED (Kuwait)              | Water price at $ 1/m³                   | $24              | $3                | 3%          | 73%              |
| RO-ED (Kuwait)              | Water price at $ 1.5/m³                 | $23              | $4                | 5%          | 74%              |
| RO-ED (Kuwait)              | Water price at $ 2/m³                   | $21              | $5                | 6%          | 76%              |

When all the parameters discussed in Section 4 are considered together, fully hybridized and current density optimized RO-ED systems can cost 33-70% less than standalone ED brine concentrators for salt production application. When including the crystallizer, optimized RO-ED-crystallizers can cost 19-55% less than standalone ED-crystallizer based seawater salt production systems.

Figure 21 shows the specific costs of brine concentration for standalone ED and optimal RO-ED designs, along with the percentage reduction in costs of RO-ED systems from that of standalone ED systems. For all brine concentration systems shown, the feed was seawater at 35 g/kg and the final concentration was 200
g/kg. All the RO-ED systems shown are fully hybridized with the RO brine being at the highest possible salinity of 120 g/kg. The cases shown are: a global average case corresponding to an electricity price of 10 US cents/kWh, a low electricity price case of 5 US cents/kWh and the special case of Kuwait where the electricity price is 1.6 US cents/kWh. A standalone ED system designed for a global case, concentrated seawater at a cost of $89/tonne-salt, while a fully hybridized RO-ED system cost 33% less at $60/tonne-salt. In Kuwait, an optimal RO-ED system can take advantage of the subsidized electricity price to concentrate brine for just $27/tonne-salt. A conventional standalone ED system designed for higher electricity prices even with the lower electricity costs in Kuwait still had a cost of $70/tonne-salt (capital cost contribution was $43/tonne-salt). Thus, the optimal RO-ED system cost 70% less than a conventional standalone ED system operated at an electricity price of 10 US cents/kWh and 62% less than one operated at an electricity price of 1.6 US cents/kWh. Thus, hybridizing RO with ED up to 120 g/kg RO brine salinity and optimizing ED current density for electricity prices, may reduce the cost of brine concentration significantly.

Table 3 highlights the critical cost reducing steps and the associated specific cost reductions that led to the outcome shown in Fig. 21. The baseline case was a standalone ED system operated in a conventional case at a current density of 300 A/m² and electricity price of 10 US cents/kWh, with the specific cost of $89/tonne-salt. Optimizing the current density of a standalone ED system for an electricity price of 10 US cents/kWh reduced the specific cost of brine concentration 26%. Fully hybridizing an ED system with an RO system such that the RO brine salinity is 120 g/kg reduced specific costs another 7%. At lower electricity prices of 5 US cents/kWh and 1.6 US cents/kWh, optimizing ED current density and the reduced RO-ED operating expenses, led to specific cost reductions of 19% and 37% respectively. The cumulative effects of each cost reduction step on the specific cost of brine concentration is highlighted in the last column in Table 3. The revenue generated by water is also accounted for in the last rows of the table with water prices of 0.5 $/m³, 1 $/m³, 1.5 $/m³ and 2 $/m³, leading to the effective specific cost reducing further by 2%, 3%, 5%, 36%
and 6% respectively. At a fixed electricity price, the effective specific cost reductions in order of magnitude of contribution were from: optimizing ED current density (26%), fully hybridizing ED with RO up to a RO brine salinity of 120 g/kg (7%) and producing water (2-6%).

Figure 22 adds the specific cost of the crystallizer component to the brine concentration systems shown in Fig. 21 showing the total specific costs of an RO-ED-crystallizer system for various cases along with the percentage reduction in the specific costs when compared with a standalone ED-crystallizer configuration. With the crystallizer, a standalone ED system operated for a global average case with an electricity price of 10 US cents/kWh produced salt at a cost of $137/tonne-salt. A fully hybridized and current density optimized RO-ED system with a crystallizer produced salt at a cost of $111/tonne-salt, corresponding to a production cost decrease of 19%. For the special case of Kuwait, when the subsidized electricity prices are included in the system design and cost calculations, the total cost of producing salt using an optimized RO-ED-crystallizer was $61/tonne-salt. A conventional standalone ED-crystallizer when operated in Kuwait produced salt at a cost of $101/tonne-salt. Thus, for the special case of Kuwait, the optimized RO-ED-Crystallizer produced salt at a cost 55% lower than a conventional standalone ED-crystallizer system operated at 10 US cents/kWh and 40% lower than one operated at Kuwaiti electricity prices. Thus, hybridizing RO with ED up to 120 g/kg, RO brine salinity and optimizing ED current density for electricity prices, may reduce the cost of salt production significantly.

6. Global feasibility of RO-ED hybrid systems for concentrating seawater for salt production

| Country    | Purity % | Type of salt, Degree of processing       | Price ($/tonne) | Min. order, Packaging, Distrib. level | Ref. | Is RO-ED Feasible? |
|------------|----------|------------------------------------------|----------------|--------------------------------------|------|-------------------|
| China      | 99.2 %   | Vacuum salt (Domestic), refined           | 90             | 25 tonnes, 50kg bags, ex-works       | 76   | No                |
| Cyprus     | 99.5 %   | Solar salt (Domestic), refined            | 200            | n.a.                                 | 77   | Yes               |
| India      | 99.3 %   | Solar salt (Industrial), refined          | 80             | 28 tonnes, 50kg bags, ex-works       | 78   | Yes               |
| Kuwait     | 97-99 %  | Raw salt (mix of solar and vacuum)        | 100            | Not defined                          | 79   | Yes               |
| Saudi Arabia | 99.8 %  | Vacuum salt (Industrial), refined         | 190            | 25 tonnes, 1000kg bags, ex-works     | 80   | Yes               |
| USA        | n.a.     | Vacuum salt                               | 190            | avg. f.o.b. plant                    | 81   | Yes               |

Based on production costs discussed in Section 5 and market prices of salt listed in Table 4 obtained from salt manufacturers and national government agencies, we believe that RO-ED hybrid brine concentration systems for salt production are technically and economically feasible in parts of Japan, the Middle East,
Europe and the United States. The economic feasibility of each region depends on its unique circumstances that arise from the following conditions:

(a) High market demand for edible salt or industrial salt by-products such as chlorine and sodium hydroxide in the region

(b) High transportation costs from available resource to point of consumption (road and shipping costs)

(c) Few natural salt deposits

(d) Lack of land availability

(e) Strategic national security reasons

(f) Scarcity of drinking water

The presence of one or more of these indicators substantially increases the price for vacuum salt in the region. Figure 23 illustrates the different salt prices found throughout the world. To evaluate the economic feasibility of the proposed RO-ED-crystallizer concept in a country, we compared the production cost of an optimal RO-ED-crystallizer, $111/tonne-salt reported in Section 5, against the market price of salt. For Kuwait, we used a production cost of $61/tonne-salt accounting for the lower electricity price in Kuwait. We reiterate here that salts depending on their production method have different production costs with vacuum salt being significantly more expensive than solar salt, and this fact is reflected in their market prices as well. The market prices of vacuum salt are well above the salt production cost from the proposed RO-ED-crystallizer concept in Cyprus, Japan, Saudi Arabia and the USA. The RO-ED-crystallizer concept may likely be competitive in these countries. It must be also noted that solar salt prices are far lower than the production cost of the RO-ED-crystallizer concept.
In this section, we further discuss some interesting dynamics related to salt pricing and transportation costs, that highlights potential opportunities for the RO-ED-crystallizer concept and new technologies for producing salt from seawater. We further discuss the specific feasibility of the RO-ED-crystallizer concept for the USA, Japan and Kuwait.

6.1. High transportation costs can create spot market opportunities

Table 5: Average cost of transportation by road, rail and barge, with distance where RO-ED-crystallizer costs break-even with the production cost of solution mined [2] vacuum salt ($35/tonne-salt assumed) and solar salt ($10/tonne-salt assumed)

| Mode      | Transportation cost | Break-even distance (km) |
|-----------|---------------------|--------------------------|
|           | Ref.                | 500 km       | 1000 km      | 1500 km      | Vacuum salt | Solar salt |
| By Road   | 10.34               | $52          | $103         | $155         | 735         | 977         |
| By Rail   | 1.87                | $9           | $19          | $28          | 4064        | 5401        |
| By Ship   | 1.14                | $6           | $11          | $17          | 6667        | 8860        |

Table 6: Container shipping costs between ports and countries, with the costs translated into cost per tonne of salt. Routes and associated costs shown are not indicative of popular salt shipping routes and are shown only to convey general variation in shipping costs.

| Origin to Destination | Cost     | Ref. | Cost per tonne |
|-----------------------|----------|------|----------------|
| City to City          | Country to Country | ($/container) | ($) |
| China to USA          | $400     | [82] | $16           |
| India to USA          | $800     | [82] | $32           |
| New York to Haifa     | $1,478   | [83] | $59           |
| Miami to Haifa        | $1,757   | [83] | $70           |
| New York to Izmir     | $1,025   | [83] | $41           |
| Barcelona to Izmir    | $978     | [83] | $39           |
| New York to Shanghai  | $748     | [83] | $30           |
| Valencia to Shanghai  | $657     | [83] | $26           |

The economic competitiveness of a salt production method is influenced heavily by the cost of transporting salt between the manufacturing facility and the customer, especially the cost of transporting salt by road. Table 5 shows the the average cost for transporting a tonne by road, rail and barge. The average cost of transportation obtained from data published by the US Department of Transportation (US DOT) [5], is given in units of US cents per tonne per km, as well as in units of $/tonne for transporting across 500 km, 1000 km and 1500 km. We observed from market reports in the literature that there was a wide range of transportation costs especially for road transportation costs. However, market reports in the literature did not specify the methodology and sample size used to obtain the transportation figures, and made it difficult to obtain an accurate average estimate of the cost of transportation. The most reliable averaged
transportation data we found was from the US DOT. For reliability of our predictions, we restricted our transport cost estimation to this dataset. Transportation by road at 10.34 US cents/tonne-km, was the most expensive mode of transport, with transportation by rail and barge being 1.87 and 1.14 US cents/tonne-km respectively.

For understanding the impact of transportation costs on the feasibility of RO-ED-crystallizer systems for salt production, the production costs for conventional-solution-mining based vacuum salt and solar salt need to be discussed. Solar salt production costs vary between $5-10/tonne-salt. For our analysis work in this section, we have assumed a production cost of $10/tonne-salt. In conventional-solution-mining based vacuum salt, near saturated brine is sent to a conventional crystallizer. From Section 3.5 on crystallizer costs, we have calculated the production cost of conventional solution mining based vacuum salt to be $ 35/tonne-salt. From Section 5 we know that the cost of producing salt from seawater through an optimized RO-ED-crystallizer system for a typical global case is $111/tonne-salt. Given this information and the transport costs mentioned in Table 5, we calculated the break-even distance at which the seawater based RO-ED-crystallizer system we have proposed becomes cost competitive with current methods. The break-even distances are also shown in Table 5. It is quite clear that the seawater-based optimized RO-ED-crystallizer systems become truly competitive when salt is transported by road as compared to rail and barge, with these systems being competitive at a distance of 735 km from conventional solution mining and at a distance of 977 km from existing solar salt production locations.

Another aspect related to transportation costs is the cost of shipping salt between countries by sea, which is different from “barge” transport. Shipping costs between countries don’t scale directly with distance. The volume of trade between countries and whether containers return empty or full affects costs more than distance. The limited relationship with distance is clear from Table 6 which shows the cost of shipping 20 foot containers between select ports around the world. It must be noted here that the routes reported in Table 6 are not in anyway representative of popular salt shipment routes and are only shown to highlight that distance does not affect shipping costs as much as in other transportation modes. For example, despite the vastly different distances, shipping a container from New York, USA to Izmir, Turkey costs almost as much as shipping from Barcelona, Spain to Izmir, Turkey. Furthermore, shipping from New York, USA to Shanghai, China is still cheaper than the examples mentioned previously. For contextualizing the container costs, a full 20 foot container can hold 25 tonnes of salt [84]. The cost of shipping from one port to another in units of USD per tonne of salt is also given in Table 6. In general, the cost to ship salt between countries can contribute $10-80/tonne-salt to the market price of salt, depending on the ports considered.

From Tables 5 and 6 it can be seen that the difference in production costs between the salt produced from an optimized RO-ED-crystallizer system and solar salt and conventional solution mined vacuum salt, can be made up in specific circumstances when salt needs to be transported over long distances by road and
between specific ports where shipping costs are high.

**Small scale decentralized production of salt has economic potential**

Table 4 showed that the RO-ED-crystallizer concept can be feasible in certain countries based on market price of salt. From the discussion in the previous section, the RO-ED-crystallizer concept for producing salt in certain regions is high due to high salt transportation costs. What follows naturally from these observations is the potential opportunity that the RO-ED-crystallizer concept can create for the small-scale decentralized production of salt. Both RO and ED systems can be scaled down to installations that can produce 10,000-50,000 tonnes per year of salt close to customers without significant changes to the brine concentration costs discussed in Section 5. However, the specific cost of crystallizers do increase as production capacity reduces and determining the minimum practical capacity of a crystallizer is beyond the scope of this work. Nevertheless, we believe that small RO-ED-crystallizer salt production plants by the sea could generate significant amount of savings for large customers of salt as a result of significant savings on transportation and shipping costs.

### 6.2. Case studies

#### 6.2.1. USA

The average market price for vacuum salt in the USA is $190/tonne-salt \[81\] which is far above the production cost of $111/tonne-salt from the RO-ED-crystallizer concept. The lack of available natural salt supply in some parts of the USA along with the high transportation costs present an opportunity for RO-ED along the west coast and the southeast coast of the USA where there are few resources to produce vacuum salt. The dearth of resources results in tonnes of salt being trucked and sent via railroad. Transporting salt by road, the most common method, adds $52/tonne for 500 km of travel. The creation of a couple of 10,000-50,000 tonne plants could lower the purchasing price of salt in this part of the USA and change the market dynamic within those regions. This suggests a reasonable commercial potential for the RO-ED-crystallizer concept in the USA.

#### 6.2.2. Kuwait

Kuwait, like other countries in the Middle East, has abundant sources of hydrocarbons needed for production of plastics, petrochemicals, organic intermediates, etc., which require chlorine for organic synthesis. Vacuum salt is the raw material used for chlorine production, and thus Kuwait and other Middle Eastern countries have a high local demand for salt. Kuwait already has a standalone ED plant that concentrates brine for salt production for chloralkali production \[14\] and which is a good candidate for adopting the proposed RO-ED-crystallizer concept for producing salt from seawater. From our literature review, we were able to obtain an import price of “raw salt” of $100/tonne-salt \[79\]. This aggregated value includes both
high purity vacuum salt as well as less pure solar salt. We expect that the import price of high purity vacuum salt will be higher. From Section 5 when the low electricity prices in Kuwait are taken advantage of, the cost of producing salt using an optimized RO-ED-crystallizer was $61/tonne-salt. Thus, even at a market price of $100/tonne-salt, the RO-ED-crystallizer concept can be feasible in Kuwait. Another factor that enhances the feasibility of the RO-ED-crystallizer concept in Kuwait, and the Middle-East in general, is the capability to co-produce drinking water. Drinking water in the Middle East is obtained primarily from seawater desalination plants with the actual price of water being $1/m³ or more. From Section 4 co-production of water could generate a secondary revenue stream greater than $2.7/tonne-salt in Kuwait and the Middle East. Thus, we strongly believe that there is a reasonable commercial potential for the RO-ED-crystallizer concept in Kuwait and by extension many other parts of the Middle East.

7. Limitations and areas for future research

As we arrive at our conclusions, we felt it necessary to highlight some of the limitations of our work and areas for future research.

- The final salt purity produced from the RO-ED-crystallizer concept is expected to be only at least 99.8% pure and cannot be used directly in an electrolyzer for chloralkali production. Further purification of the salt is needed for use in an electrolyzer. This could be done either before the ED concentrate enters the crystallizer or on the salt produced. Future work can look into the economics of how this adds to cost and changes the price the salt could command.

- The ED transport for salts other than sodium chloride has not been characterized at high salinity. Thus, the composition of the ED concentrate would be best determined through experiments. Future research work should look into developing an ED model that can accurately predict the composition of ED concentrate.

- While we have shown that operating ED at higher current densities than 300 A/m² reduces costs, further development work needs to be carried out to properly design such a system with appropriate choices of channel height, stack length etc.

- Our analysis showed that hybridizing ED with RO up to RO brine salinities of 120 g/kg will reduce costs. The cost advantage in our opinion justifies a pilot trial of the proposed configuration. For pilot testing, achieving higher current densities in the ED system should be prioritized over hybridizing with RO.

- While the focus of our analysis was on concentrating seawater and producing salt, the RO-ED-crystallizer concept could also be used for producing salt from saline inland lakes. However, saline in-
land lakes have compositions very different from seawater. Hence, the feasibility of RO-ED-crystallizers for producing salt from saline inland lakes should be examined on a case to case basis.

8. Conclusions

Brine concentration

RO systems are more cost effective at concentrating seawater than ED systems, and hybridizing ED with RO reduced brine concentration costs. However, the ED systems should be fully hybridized with the emerging high pressure RO technology, up to operational limits of an RO brine salinity of 120 g/kg for significantly reducing costs (7 %). Partial hybridizing of RO and ED systems, with the RO brine salinity restricted to the operational limit of conventional seawater RO of 70 g/kg brine salinity, maybe considered only if water prices are sufficiently high as the cost reductions alone over standalone ED are small (2-3 %).

ED capital costs are the main contributor to the cost of brine concentration and increasing ED current density can reduce ED capital costs significantly with there being a cost optimal value for ED current density. Optimizing ED current density to electricity prices can reduce specific costs (∼ 20 %) more than fully hybridizing with RO alone (∼ 10 %).

Fully hybridizing RO with ED up to an RO brine salinity of 120 g/kg, optimizing ED current densities and customizing current density for local electricity prices reduced the brine concentration cost by 33-70 % when compared with standalone ED systems with the variation coming from varying electricity prices globally.

Salt production

Hybrid RO-ED brine concentration systems coupled with a crystallizer are technically and economically feasible for producing salt from seawater in parts of Japan, Kuwait, the Middle East, Europe and the USA.

The total production cost of salt from a RO-ED-crystallizer was lower that that from conventional standalone ED-crystallizer systems by 19-55 %. The production cost of salt from an optimal RO-ED-crystallizer system varied from $61-111/tonne-salt. For a global case corresponding to an electricity price of 10 US cents/kWh, the salt production cost from the proposed RO-ED-crystallizer was $111/tonne-salt — sufficiently below the market price of vacuum salt in Cyprus, Kuwait, Saudi Arabia, and the USA.

Co-production of water from RO-ED brine concentration systems can further subsidize salt costs 1-6 % improving the economic feasibility in these markets.

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Nomenclature

**Acronyms**

CapEx capital expense, $

Cap_{fac}$ capacity factor, -

Cost$_{elec}$ electricity cost, $/kWh$_e$

Cryst crystallizer

CSSF crystallizer salinity scale-up factor, -

ED electrodialysis

OpEx operating expense, $

Price$_{water}$ water price, $/m^3
RO reverse osmosis
SpCapEx specific capital expense, $/tonne-salt
SpCost specific cost, $/tonne-salt
SpE specific energy, kWh/tonne-salt
SpOpEx specific operating expense, $/tonne-salt

Roman Symbols

\( A \) area, m\(^2\)
\( C \) concentration, mol/m\(^3\)
\( D \) diffusion coefficient, m\(^2\)/s
\( E \) potential, V
\( E \) specific energy consumption, kWh/m\(^3\)
\( F \) Faraday’s constant, C/mol
\( h \) channel height, m
\( i \) current density, A/m\(^2\)
\( J \) molar flux, mol/m\(^2\)-s
\( L_s \) salt permeability, m\(^2\)/s
\( L_w \) permeability to water, mol/m\(^2\)-s-bar
\( \text{Mem}_{\text{eff}} \) Membrane effectiveness, -
\( \dot{m} \) mass flow rate, kg/s
\( \dot{m}_{\text{salt, yr}} \) salt mass flow rate, tonne/yr
\( \dot{N} \) molar flow rate, mol/s
\( P \) pressure, bar
\( \Pi \) osmotic pressure, bar
\( r \) rate of return on capital, -
\( \text{Re} \) Reynolds number, -
\( \bar{r} \) area resistance, \( \Omega \cdot m^2 \)
\( S \) salinity, kg salt/kg solution
\( Sc \) Schmidt number, -
\( Sh \) Sherwood number, -
\( t \) solution transport number, -
\( T_s \) membrane salt transport number, -
\( T_w \) membrane water transport number, -
\( \bar{T} \) integral ion transport number, -
\( v \) flow velocity, m/s
$V_{cp}$  cell pair voltage, V
$\dot{V}$  volume flow rate, m³/s
$w$  water
$\dot{W}$  Power, W

Greek Symbols

$\Delta$  difference
$\Lambda$  molar conductivity, S.m²/mol
$\mu$  chemical potential, J/mol
$\nu$  viscosity, -
$\pi$  osmotic pressure, bar
$\rho$  density, kg/m³
$\Sigma$  sensitivity, -
$\tau$  time, years

Subscripts

am  anion exchange membrane
c  concentrate
cm  cation exchange membrane
cp  cell pair
cu  counter ion
d  diluate
ED  electrodialysis
f  feed
m  at membrane surface
i  counting index, inlet
o  outlet
pump  pump
p  product
RO  reverse osmosis
s  salt
w  water
Appendix A. Cost model for high pressure reverse osmosis

In this section, we evaluate the capital costs of adding a HPRO stage to a seawater RO system producing 550 m$^3$/day of product water. The brine from the seawater RO system goes to the HPRO stage at a rate of 726 m$^3$/day where pure product water and 120 g/kg brine is produced. To obtain the cost of HPRO, we first evaluated the capital costs of a seawater RO system taking in the same amount of feed (726 m$^3$/day) producing 305 m$^3$/day at 42% recovery. For this purpose we used DesalData’s seawater RO capital cost estimator [43]. DesalData had collected the cost information from numerous plants and created an accurate cost estimator. The input data used for the calculator is shown in Table A.7.

Table A.7: Input variables used on DesalData [43] to obtain capital costs for a conventional seawater RO plant

| Plant feature          | Data                  |
|------------------------|-----------------------|
| Capacity               | 305 m$^3$/day         |
| Seawater salinity      | 30,000 mg/L           |
| Seawater min. temp.    | 15 °C                 |
| Seawater max. temp.    | 24 °C                 |
| Pretreatment           | Standard              |
| Second pass            | 0%                    |
| Remineralization       | No                    |
| Intake/Outfall         | Typical               |
| Permitting             | Typical               |
| Country                | Any                   |

Figure A.24: Breakdown of capital costs in a 305 m$^3$/day capacity seawater RO plant with the data sourced from Desaldata [43]. The components affected by pressure account for 29% of the capital costs.

Figure A.24 lists the different components contributing to the total capital costs of a seawater RO plant. Of all the components listed, only five terms are affected by pressure: piping, pumps, membranes, pressure vessels and energy recovery devices. This is highlighted in Fig. A.24. These components together contribute 29% to the overall capital costs for a seawater RO plant. Individually: piping, pumps, membranes, pressure vessels and energy recovery devices, each account for 14%, 7%, 7%, 1% and 1% of overall capital costs. The
Table A.8: Break down of capital cost of RO and HPRO per m³/day production capacity for a 305 m³/day system. RO capital costs obtained from Desaldata [43].

| CapEx contributor         | RO CapEx ($/day/m³) | HPRO-RO CapEx ratio | HPRO CapEx ($/day/m³ prod) |
|---------------------------|---------------------|----------------------|---------------------------|
| Equipment and materials   | $530                | 1                    | $530                      |
| Civil costs               | $387                | 1                    | $387                      |
| Installation services     | $221                | 1                    | $221                      |
| Intake / Outfall          | $166                | 0                    | $0                        |
| Pretreatment              | $173                | 0                    | $0                        |
| Design costs              | $249                | 1                    | $249                      |
| Legal and professional    | $16                 | 1                    | $16                       |
| Piping, high-grade alloy  | $334                | 1.24                 | $414                      |
| Pumps                     | $169                | 1.67                 | $282                      |
| Membranes                 | $161                | 3                    | $484                      |
| Pressure vessels          | $35                 | 1.67                 | $58                       |
| Energy recovery devices   | $27                 | 1.67                 | $45                       |
| Total                     | $2,468              |                      | $2,686                    |

The cost of operating seawater RO at 120 bar can be accurately estimated if the cost of these four components at 120 bar operation are estimated.

Table A.8 lists the specific capital cost of each component of a seawater RO plant, a conversion factor to convert the component cost from RO to HPRO and the final estimated cost of each component of a HPRO stage of the same size. The total specific capital cost of a 305 m³/day seawater RO plant was 2468 $/day/m³ of product water. A HPRO stage added to a seawater RO plant can be assumed to have the same specific costs of the RO plant for equipment and materials (i.e. refers to control systems and other peripheral equipment [48]), civil costs, installation services, design costs, and legal and professional costs. The costs associated with the intake/outfall and pre-treatment (i.e. refers to coagulation and media filters etc. to remove particulates) can be neglected for the HPRO stage since these components have already been accounted for in the RO plant. For the piping, pumps, membranes, pressure vessels and energy recovery devices, we estimated conversion factors of 1.24, 1.67, 3, 1.67 and 1.67. The justifications for these are described below.

**Pumps** We were able to find in the literature pump costs at the pilot scale of around 250 m³/day flow rates [85, 86]. At the pilot scale, the 120 bar pump [85] cost 1.67 times more than the 60 bar pump [86] for the same amount of liquid being pumped. We believe that at larger capacities and with economies of scale coming in with wider adoption, cost differences between a 60 bar and 120 bar pump will decrease significantly. We applied this 1.67 factor to the data from DesalData and estimated the cost of pumps for a 120 bar RO plant. This is shown in Table A.8.

**Membranes** From conversations with membrane suppliers, we found that the current price of the high
pressure RO membrane, DOW XUS180808 Reverse Osmosis Element [34], was 3 times that of conventional seawater RO membranes.

**Pressure vessels** The pressure vessels in an RO plant must surround the membranes to ensure that they do not burst when the high pressure feed enters them. The DOW XUS180808 element has an 8 inch diameter and needs a pressure vessel that can withstand the full 120 bar that the pumps will provide. The ASME (American Society of Mechanical Engineers) publishes an equation on pipe thickness relative to many factors when choosing piping for plants [37]. The equation is as follows:

\[
    t = \frac{P \times (d + 2c)}{2(SE - P(1 - Y))} \tag{A.1}
\]

where \( t \) is pipe thickness, \( P \) is the design pressure, \( d \) is the pipe diameter, \( c \) is the mechanical allowances including corrosion, \( S \) is the allowable strength, \( E \) is the quality factor, and \( Y \) is the material factor. The quality factor \( E \) is influenced by the type of pipe. For seamless pipes, \( E \) is 1. The Material Factor \( Y \) is influenced by the operating temperature of the pipes. Below 900°F, the factor is taken to be 0.4. For steel, the allowable stress is 20,000 psi. The mechanical allowance typically chosen to allow for threading and corrosion is 0.125 in.

Given this equation, and a pressure vessel internal diameter of 8 in, we can calculate that at 60 bar (870 psi), the thickness of material required is 0.309 in. or 7.85 mm. This means that the cross section area of the vessel was 8.07 in² or 52.08 cm². However, at 120 bar, the thickness of material required is 0.504 in or 12.80 mm. This leads to a cross section area of 13.45 in² or 87.80 cm². The cross-sectional area increased by 1.67 times. Assuming that the only cost change comes from requiring thicker material, the pressure vessels in an HPRO system will cost 1.67 times that in a conventional seawater RO system.

**Piping** Similar to pressure vessels, using equation (A.1) piping also needs an upgrade. The piping for the DOW XUS180808 element [34] must have an internal diameter of 1.125 in or 28.575 mm. This leads to a 60 bar (870 psi) requirement of 0.158 in or 3.955 mm thickness corresponding to a cross section area of 0.63 in² or 4.04 cm². At 120 bar (1740 psi), the thickness of material required is 0.188 in or 4.78 mm corresponding to a cross section area of 0.78 in² or 5.01 cm². Thus, the cross-sectional area and the volume of material required increased 1.24 times. Assuming that the material cost is the main price determinant, we can estimate the capital cost of piping in an HPRO system to be 1.24 times that in a conventional seawater RO system.

**Energy recovery devices** While we could not get direct quotes for how much an energy recovery device (ERD) operating at 120 bar could cost, we estimated that at scale, the cost increase would be similar to that for pumps. Thus conservatively we assumed that ERDs in a HPRO system would cost 1.67 times that in a conventional seawater RO system.

**Specific capital cost of of HPRO** The factors used estimated that the specific cost of a HPRO stage was
2686 $\text{-day/m}^3$ of product water at 42 % recovery or 1.09 times the cost of conventional seawater RO. It must be noted here that the HPRO skids itself (pressure vessels, piping, pumps, membranes and ERD) cost 1.77 times that of conventional seawater RO. However, the skids only amounted to 29 % of the total specific cost leading to only a 9 % increase in the stage specific cost:

$$\text{SpCapEx}_{\text{HPRO}, \text{p}, \text{day}} = 1.09 \times \text{SpCapEx}_{\text{RO}, \text{p}, \text{day}}$$ (A.2)

Thus, the total capital cost for the HPRO stage was:

$$\text{CapEx}_{\text{HPRO}} = \text{SpCapEx}_{\text{HPRO}, \text{p}, \text{day}} \times \dot{V}_{\text{HPRO}, \text{p}, \text{day}}$$ (A.3)

The HPRO stage capital expenses was then annualized to obtain:

$$\text{CapEx}_{\text{HPRO, yr}} = \frac{\text{CapEx}_{\text{HPRO}}}{\text{Annuity factor}}$$ (A.4)

HPRO membranes contributed to 18 % of the cost of the HPRO stage:

$$\text{CapEx}_{\text{HPRO, mem.}} = 0.18 \times \text{CapEx}_{\text{HPRO}}$$ (A.5)

Assuming a membrane life of 5 years and annualizing we have a HPRO membrane replacement cost of:

$$\text{OpEx}_{\text{HPRO, mem., yr}} = \frac{\text{CapEx}_{\text{HPRO, mem.}}}{\text{Annuity factor}} \times \left[ (1 + r_{\text{return}})^{-5} + (1 + r_{\text{return}})^{-10} + (1 + r_{\text{return}})^{-15} \right]$$ (A.6)

The maintenance, chemical and labor cost for seawater RO was 0.18 $/\text{m}^3$ of RO product water. Multiplying this with the annual product water flow from the HPRO stage we have the annual operating cost of maintenance, chemicals and labor:

$$\text{OpEx}_{\text{HPRO, maint., chem., labor, yr}} = 0.18 \times \dot{V}_{\text{p, HPRO, yr}}$$ (A.7)

The total annual operating expense for the HPRO stage was:

$$\text{OpEx}_{\text{HPRO, yr}} = \text{OpEx}_{\text{HPRO, energy, yr}} + \text{OpEx}_{\text{HPRO, mem., yr}} + \text{OpEx}_{\text{HPRO, maint., chem., labor, yr}}$$ (A.8)

where $\text{OpEx}_{\text{HPRO, energy, yr}}$ was obtained be applying the RO process model to the HPRO feed and brine salinities. Thus, the total specific capital cost of the HPRO stage normalized to the annual salt production
was:

\[
\text{SpCost}_{\text{HPRO}} = \frac{\text{CapEx}_{\text{HPRO},\text{yr}} + \text{OpEx}_{\text{HPRO},\text{yr}}}{\dot{m}_{\text{salt},\text{yr}}}
\]  \hspace{1cm} (A.9)

**Appendix B. Cost model for nanofiltration**

While an NF system may not be needed for the operation of a HPRO stage due, it may be implemented for other reasons such as increasing the purity of the salt and for reducing maintenance costs on the ED and crystallizer systems. For the benefit of salt production plant designers, we have provided a preliminary analysis of NF here.

To the best of our knowledge, there is some operational experience in using NF systems with seawater as feed [88]. We used Dow Chemical’s ROSA program [49] to evaluate the divalent rejections of Dow’s NF90 and NF270 membranes and determined that rejection was poor if RO brine was used. Seawater feed yielded much higher rejections with NF270-4040 showing the highest rejection of divalent ions. The NF system configuration and composition of the permeate was determined using Dow’s ROSA software [49]. A configuration using two stages with the first stage operated at 22.4 bar (\(\Delta P_{p,NF} = 21.4\) bar), with the first stage having roughly 3 times as many pressure vessels as the second stage, with each pressure vessel having 6 elements was found to operate at 70% recovery with seawater as feed. A booster pump was not used between stages. ROSA estimated that the reduction in calcium, magnesium and sulfate ions from the feed to the permeate was 56, 59 and 97% respectively. We worked with an antiscalant company [54] and ascertained the type of and dosage of antiscalant needed to prevent scale formation in the NF system on the concentrate side.

A pump efficiency of 0.85 was assumed and the work consumed by the NF pump was calculated as follows:

\[
\dot{W}_{NF} = \dot{V}_{f,NF} \Delta P_{p,NF} \frac{1}{\eta_{p,NF}}
\]  \hspace{1cm} (B.1)

The hourly energy costs of the NF system can be obtained as:

\[
\text{OpEx}_{NF, \text{energy, hour}} = \frac{\dot{W}_{NF}}{1000[W/kW]} \text{Cost}_{\text{elec}}
\]  \hspace{1cm} (B.2)

Accounting for a capacity factor, \(\text{Cap}_{\text{fac}} = 0.9\), the annual NF energy cost was obtained as:

\[
\text{OpEx}_{NF, \text{energy, yr}} = \text{OpEx}_{NF, \text{energy, hour}} \times 8760[\text{hr/year}] \times \text{Cap}_{\text{fac}}
\]  \hspace{1cm} (B.3)

The specific cost of maintenance, chemicals and labor of RO on a per feed basis was assumed for NF. Assuming an average RO recovery of 42%, this is 0.0756 $/m^3$-feed. Multiplying this with the annual feed
flow to the NF stage we have the annual operating cost of maintenance, chemicals and labor:

\[ \text{OpEx}_{\text{NF, maint., chem., labor, yr}} = 0.0756 \frac{\$}{m^3} \times \dot{V}_{\text{f, NF, yr}} \]  

The capital cost for the seawater NF system was sourced from an actual installation of a 2-stage NF system treating 4000 m\(^3\)/hour of seawater at a recovery ratio of 70 \% \[^{[89]}\]. The specific capital cost of the system complete with process equipment, controls, civil construction, membranes and installation costs was 948 \$-day/m\(^3\) of NF permeate \[^{[89]}\]. A seawater RO plant taking in the same amount of feed would be able to produce 40,320 m\(^3\)/day of product water. The specific capital cost of such a plant from Eq. 14 is 1484 \$-day/m\(^3\) of RO product \[^{[43]}\]. The specific capital cost of the NF system for the capacity we were interested in was obtained by scaling with Eq. 14 as:

\[ \text{SpCapEx}_{\text{NF,p,day}} = \text{SpCapEx}_{\text{RO,p,day}} \times \frac{948.9}{1484} \]  

Nanofiltration membranes accounted for only 5 \% of the specific capital costs. A membrane life of 5 years was assumed. Based on the same method previously used for RO, we obtained the annualized the capital costs and membrane replacement costs. The energy and maintenance cost was added to obtain an NF specific cost. For a NF system that treated 100 \% of the RO feed, for an optimized RO-ED-crystallizer operated at 600 A/m\(^2\) and 120 g/kg RO brine salinity, the specific cost of the NF system was 5.1 \$/tonne-salt. This could be considered as an upper bound in costs if NF was added to the RO-ED-crystallizer configuration.

Pilot tests would have to be conducted to ascertain and weight the benefits and cost of using NF. It is possible that the use of NF may lead to reasonable reductions in maintenance costs. Furthermore, NF pretreatment may not be needed on 100 \% of the RO feed depending on the final objective. By our estimates, even if only 55 \% of the RO feed was treated using NF, the saturation limit of calcium sulfate at 120 g/kg brine salinity could be avoided. The specific cost of such an NF system was only 2.8 \$/tonne-salt. We report these numbers here to encourage future research work into hybridization of NF with RO-ED systems.