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Techno-economic Analysis of Ion Concentration Polarization Desalination for High Salinity Desalination Applications

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Abstract

A techno-economic analysis is used to evaluate the economic feasibility of ion concentration polarization (ICP) desalination for seawater desalination and brine management. An empirical optimization model based on a limited set of experimental data, which was obtained from a lab-scale ICP desalination prototype, was established to calculate the required energy and membrane area for a given set of operating parameters. By calculating operating and capital expenses in various feed and product cases, the optimal levelized cost of water is determined over a range of feed salinities, mostly above seawater salinity (35 g/kg). Through these analyses, we study the economic feasibility of three applications: 1) partial desalination of brine discharge by ICP (feed varied from 35 – 75 g/kg) to common seawater RO feed level (35 g/kg) in a hybrid ICP-RO system; 2) the concentration of seawater desalination brine for salt production, and 3) partial desalination of oilfield wastewater. The economic feasibility of ICP desalination processes has been evaluated and the rough cost of treatment has been generated for several relevant applications. The approach taken in this work could be employed for other new and existing desalination processes, where *a priori* process modeling and optimization is scientifically and/or numerically challenging.

Keywords

Electromembrane desalination; high salinity desalination; brine management; economic analysis; ion concentration polarization

42 Nomenclature

43 *Roman symbols*

44	A_m	Total membrane area (mm ²)
	C	Concentration (g/kg)
	CE	Current efficiency
	d	ICP cell depth (mm)
	F	Faraday's constant (=9.65 × 10 ⁴ C/mol)
	I	Electric current (A)
	K_E	Electricity rate (\$/kWh)
	K_Q	Capital cost per unit membrane area (\$/m ²)
	L	ICP cell length (cm)
	Q	Flowrate (L/min)
	R	Cost of capital (%)
	SRR	Salt removal ratio
	T	Life time of equipment (year)
	V	Voltage (V)
	W	ICP cell width (mm)

45 46 *Greek letters*

47	μ	Dynamic viscosity (g/(cm·s))
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1. Introduction

Desalination processes for brackish water and seawater have greatly advanced over the past few decades with the development of membrane desalination technologies. Reverse osmosis (RO), the state-of-the-art membrane desalination technology, has reached an energy consumption (~ 3 kWh/m³) that is approaching the theoretical minimum (~ 1 kWh/m³, for 35 g/kg feed with 50% recovery) (Elimelech and Phillip, 2011; Lienhard et al., 2017) and electrodialysis, an electromembrane desalination technology, has achieved 1.65 ~ 1.85 kWh/m³ (for 32 ~ 33 g/kg feed) for seawater desalination (Aulick, 2014; “Bandwidth Study on Energy Use and Potential Energy Savings Opportunities in U.S. Seawater Desalination Systems,” 2017). In spite of these meaningful advances, desalination of high salinity brine, with salinity above seawater level (35 g/kg), has been a continuing challenge with few economically attractive options for treatment. In fact, the need for hypersaline brine desalination or concentration is growing, as the volume of brine generation increases. Two major sources of the brine are the concentrate from seawater desalination plants, and the wastewater (produced water) from modern oil / gas extraction processes. With ever-increasing desalination capacity due to economic growth, the volume of brine generation is steadily growing (Arnal et al., 2005; McGovern et al., 2013; Morillo et al., 2014). Moreover, efficient desalination operation and higher recovery of pure product water have created more concentrated brine (50 – 85 g/kg) (Lattemann and Höpner, 2008), which makes treatment more challenging. In the shale gas industry, management of produced water (8 – 360 g/kg) is one of the biggest challenges for economical and environmental reasons (Gregory et al., 2011; Shaffer et al., 2013; Thiel et al., 2015; Vidic et al., 2013).

Current methods of brine management include surface discharge, solar evaporation, underground injection, and internal reuse for enhanced oil recovery technologies (*e.g.* hydraulic fracturing).

Surface discharge is the most rudimentary method of disposal with minimal costs involved. However, the discharged brine can disrupt the ecosystem by increasing the local salinity of the surface water and by contaminating the surface water with toxic chemicals added during desalination or hydraulic fracturing (*e.g.*, anti-foulants, anti-scalants, surfactants, etc.) (Del Bene et al., 1994; Gacia et al., 2007; Roberts et al., 2010; Uddin et al., 2011). Solar evaporation involves evaporating brine from large, shallow ponds using natural solar energy and removing residual solid for disposal (Katzir et al., 2012). Evaporation ponds are easy to construct and to maintain, but require large land footprint (which raises capital cost), and may cause environmental liabilities in case of brine leakage (Ahmed et al., 2000; Gilron et al., 2003). Underground injection is the brine management practice employed to dispose majority of produced water from shale gas operation (Clark and Veil, 2009). Underground injection is limited by the number of available disposal wells and the transportation from produced water generation sites to wells (Rassenfoss, 2011). To reduce the brine volume in the shale gas industry, reuse of produced water has become increasingly popular (Shaffer et al., 2013). Internal reuse of produced water can reduce the demand for injection wells and freshwater, but can hurt the efficiency of the hydraulic fracturing process due to the high concentration of dissolved ions and other chemicals (Lutz et al., 2013; Nicot and Scanlon, 2012). Reuse needs to be accompanied by another disposal method, for the reused produce water will eventually need to be disposed when the reuse demand becomes low as the shale gas formations mature (Lutz et al., 2013; Shaffer et al., 2013; Vidic et al., 2013). As reviewed here, the current brine management methods are not sustainable to treat an increasing volume of high salinity brine. Therefore, desalination should be employed to reduce the volume of brine that needs to be managed by solar evaporation or underground injection.

94 Current desalination technologies for potential treatment of high salinity brine include reverse
95 osmosis, thermal desalination technologies, such as humidification-dehumidification (HDH)
96 multi-stage flash (MSF), multi-effect distillation (MED), and mechanical vapor compression
97 (MVC), and membrane distillation (MD) (Thiel et al., 2015). Techno-economic analyses have
98 been implemented to evaluate the economic benefits of technologies for desalination, water
99 treatment, and water reuse processes (Capocelli et al., 2019; Ghaffour et al., 2013; Prisciandaro et
100 al., 2016). Reverse osmosis (RO) is the most cost efficient desalination technology to treat saline
101 water with up to 35 g/kg TDS (Fritzmann et al., 2007; Greenlee et al., 2009; Tow et al., 2015).
102 However, for high salinity feed, the RO membrane modules are not normally designed for the
103 hydraulic pressure needed to overcome the large osmotic pressure, which arises from the large
104 salinity difference between the feed and the permeate (Fritzmann et al., 2007; Greenlee et al.,
105 2009). For this reason, high salinity brine is typically treated with thermal technologies, such as
106 MVC, HDH, or MD (Chehayeb and Lienhard, 2015; Swaminathan et al., 2018). These thermal
107 desalination technologies are more energy intensive than membrane processes at any salinity, and
108 all technologies require greater energy input at higher salinities than at lower salinities (Thiel et
109 al., 2015). Further, many thermal technologies (such as MSF, MED, MVC) require large amounts
110 of costly metal heat transfer surfaces, resulting in relatively high investment cost and large
111 footprint, compared to membrane desalination (Darwish et al., 2003; Ghaffour et al., 2013); the
112 exception is HDH which achieves lower capex through use of inexpensive polymers. To obtain
113 low-temperature desalination plants of the MSF and MED type are typically built next to power
114 plants to utilize the heat generated from these plants, and thus these plants are less mobile and
115 scalable (Shaffer et al., 2013). MVC is a more scalable thermal desalination technology, with
116 electrical energy consumption in the range of 10 - 30 kWh/m³ for treatment of brine with salinity

beyond seawater level (38 – 260 g/kg) (Chung et al., 2017; Koren and Nadav, 1994; Thiel et al., 2015; Veza, 1995). MVC is modular in design, and thus involves lower capital cost and is built at small scale more often than MSF and MED (Heins, 2010). Membrane distillation (MD) is a relatively new technology and is a thermally driven membrane desalination process. Its advantages are the use of low-grade heat source and the low sensitivity of performance to the feed salinity, which makes the process appealing for high salinity brine treatment (Alkhudhiri et al., 2012; Meindersma et al., 2006). Single stage MD with heat recovery can achieve thermal energy consumption of 70 - 90 kWh/m³ for seawater desalination (ALOBaidani et al., 2008). MD has also been proposed as the heat recovery element in MVC to improve its pure water recovery and specific energy consumption (Swaminathan et al., 2016). Roughly speaking, the alternative technologies for treating brine (salinity significantly higher than seawater) based on distillation processes are priced around \$20~50/m³, depending on the intake salinity and other parameters. (Alkhudhiri et al., 2012; ALOBaidani et al., 2008; Meindersma et al., 2006)

Electromembrane desalination, such as electrodialysis (ED), exhibits characteristics that can be particularly appealing for high salinity brine desalination. Since these processes are electrically driven, highly saline water will increase the ion conductance, making the process more energetically favorable (Chehayeb et al., 2017; Długolecki et al., 2010). In addition, electromembrane desalination enables ‘partial desalination’ because salt removal in the product stream can be easily altered. Due to this reason, brine treatment by ED has been demonstrated in several previous studies (Korngold et al., 2009; McGovern et al., 2014a; Nikonenko et al., 2014; Oren et al., 2010; Strathmann, 2010; Turek, 2003). Ion concentration polarization (ICP) desalination is an electromembrane desalination technology that has been developed by Han and coworkers (Kim et al., 2016; Kwak et al., 2016). ICP desalination (**Fig. 1a**) is a unipolar

electromembrane desalination process, employing one type of ion exchange membrane, whereas ED employs both cation and anion exchange membranes. In the case that a feed stream has sodium chloride as the majority salt, ICP desalination with cation exchange membrane (CEM)s can remove salt (*i.e.*, sodium chloride) more efficiently at a given current, because CEMs can generate stronger depletion layer than AEMs due to the difference of diffusivity (the diffusivity of chloride is higher than the diffusivity of sodium) (Kwak et al., 2016) (See **Fig. 1(b)**). The system was engineered to operate more energy efficiently, and its potential application was identified to be in partial desalination of high salinity brine (Kim et al., 2016; 2017).

In this work, we perform a more detailed techno-economic analysis to evaluate the economic feasibility of ICP desalination for seawater desalination and desalination brine management. This requires modeling prediction of efficiency at various flow rates and operating currents, for a given brine intake. Yet, such a model-based optimization is not yet readily available for ICP desalination nor electromembrane separation processes in general. While ICP desalination is utilizing the same Cation Exchange Membranes (CEMs) as in ED, its unique fluidic and membrane architectures prevent one from using widely accepted ED transport models. In addition, conventional ion selective membranes (e.g. CEMs) are generally optimized for brackish desalination applications (Strathmann, 2010), and it has been reported that high salinity intake water will generate many nonlinearity in membrane performances, which are scientifically challenging to model accurately. (Chehayeb and Lienhard, 2017; Narebska et al., 1984) In addition, it is often necessary to operate electromembrane processes near limiting current (in order to get sufficient salt removal required for brine treatment), for which there is no accurate transport models available. (Hattenbach and Kneifel, 1986)

Instead of *a priori* physics models, in this work we built an empirical model based on experimental data from a lab scale prototype, to estimate the energy consumption and the membrane area for a range of operating parameters that are relevant for brine desalination applications. Using this model, we optimized the operating conditions for the minimum water cost, given the feed and the product concentrations. We calculated the water cost for three applications: partial desalination of brine to seawater level by ICP, followed by conventional RO; and concentration of seawater desalination brine for salt production. The cost of water produced by ICP desalination was compared with the cost of water produced by other desalination technologies to help determine the competitiveness of ICP desalination.

2. Methods

2.1. Experimental

The experimental data was generated with a lab-scale ICP desalination device, whose configuration, fabrication, and operation were described and demonstrated in a previous work (Kim et al., 2017). The operation of ICP desalination is described in **Fig. 1(a)**. The channel dimensions were 2.5 mm in depth (*i.e.*, effective intermembrane distance), 8 mm in effective width (perpendicular to flow direction), and 30 cm in length (same as flow direction). The number of unit flow cells is four, which means five CEMs were used in total. Two FTCEM-E membranes (FuMA-Tech GmbH, Germany), which is mechanically reinforced, are installed in the outside and three Neosepta CMX membranes (Astom Co., Japan) are installed in the middle. The electric potential drop for the unit cell is precisely measured by employing chloridized wired electrodes (Ag/AgCl, A-MSystems Co., Sequim, WA) between the CEMs. This allows accurate measurement of electric potential drop across single unit cell, without being affected by the large potential variation occurring in the rinsing channel. A benchtop conductivity meter (Orion Star Series meter,

Thermo Fisher Scientific Inc., Cambridge, MA) is connected to a flow-through conductivity microelectrode (Microelectrode Inc., Bedford, NH) in order to monitor real-time salinity changes. The flow rate ratio of diluate to concentrate to feed stream was 1:1:2, and the overall recovery of the diluate was 50%. Varied parameters included feed salinity, flow velocity, and salt removal ratio (SRR):

$$SRR = \frac{C_{feed} - C_{diluate}}{C_{feed}} \quad (1)$$

where C_{feed} and $C_{diluate}$ are feed and product concentration, respectively. Since high salinity desalination applications of ICP desalination were considered, the tested feed salinity ranged from 35 g/kg to 100 g/kg of sodium chloride. The ranges of salt removal and the flow velocity were 10% – 90% and 0.25 – 3 mm/s, respectively. The current and voltage for each experiment were consequently varied to achieve the target SRR for the given feed salinity and flow velocity.

2.2. Estimation of electrical energy consumption

In order to estimate electrical energy consumption for any given set of feed salinity, SRR, and flow velocity, a relation must be built based on the experimental data. From the measured current and voltage values, the power and the electrical energy consumption were calculated, and the relation between the current and the power consumption were fitted to an equation. To estimate current for a given set of feed salinity, SRR, and flow velocity, another fitting was done to first estimate current efficiency (CE) as a function of SRR and flow velocity; the current efficiency was then used to calculate the corresponding current. The relation between current and current efficiency is shown in the equation below.

$$CE = \frac{zfQ_{diluate}(C_{feed} - C_{diluate})}{I} \quad (2)$$

where z is ion valence, F is Faraday's constant, $Q_{diluate}$ is volumetric flowrate of product stream and I is total current.

2.3. Levelized cost of water calculation

The Levelized Cost of Water (LCOW) was calculated with a simple cost model that consists of operating and capital costs. This method is a well-established method for cost analysis of electrodialysis and ICP desalination (Kim et al., 2017; 2016; McGovern et al., 2014a; 2014b). Electricity and pumping energies were used to calculate the operating cost. Total membrane area was utilized to estimate the capital cost since the capital cost of electrodialysis, which is very similar to ICP desalination in configuration and operation, depends strongly on the total membrane area (Strathmann, 1992). Equations below were used to calculate the electricity, pumping, and capital costs. The LCOW was obtained by summing the operating and capital costs and expressed in US dollars per cubic meter of feed solution. Parameters used in the cost analysis are listed in **Table 1**. The value of capital cost per unit membrane area (K_Q) was approximated with the K_Q values for ED and ICP desalination as guides (Kim et al., 2017; McGovern et al., 2014b; Sajtar and Bagley, 2012). More specifically, the value of K_Q for ED were \$480 /m²-membrane in McGovern *et al.* (McGovern et al., 2014b) and \$600 /m²-membrane from an industry quote (Nayar, 2016). ICP desalination presents the opportunity to reduce K_Q below that for ED desalination. ICP desalination needs only CEM, but ED desalination requires both cation and anion exchange membranes. The fouling in electromembrane processes mainly appears in the anion exchange membrane depending on wastewater quality (Lee et al., 2009), because most organic foulants in

many effluent streams are negatively charged, thus fouling anion exchange membranes due to deposition and/or adsorption. Also, ICP desalination requires only two pumps for feed and rinse streams, but ED desalination requires three pumps for the diluate, concentrate and rinse streams. However, the value of K_Q for ICP desalination was assumed to be 25% higher than that for ED, quoted from the industry (\$600 /m²), since ICP desalination is a new technology, which requires further development, and thus the manufacturing costs are higher. To obtain the optimal LCOW for a given set of feed and product salinities, the LCOW was calculated over a range of flow velocity (and resulting current), and the optimal operating conditions that result in the minimum LCOW were determined.

$$\text{Electricity Cost (\$/m}^3\text{)} = \frac{I \times V}{Q} \times K_E = \frac{P}{Q} \times K_E \quad (3)$$

$$\begin{aligned} \text{Pumping Cost (\$/m}^3\text{)} &= \frac{\text{Pumping Power}}{\text{Feed Flowrate per Cell}} \times N_{\text{cell}} \times K_E \\ &= \frac{12\mu QL}{wd^3} \times \frac{Q_{\text{total}}}{Q_{\text{cell}}} \times K_E \end{aligned} \quad (4)$$

$$\begin{aligned} \text{Capital Cost (\$/m}^3\text{)} &= \frac{\text{Used Membrane Cost (\$)}}{\text{Output Flow Volume per Life (m}^3\text{)}} \times \text{Anualized Factor} \\ &= \frac{A_m \times K_Q}{Q \times T} \frac{(1 + R)^T - 1}{T \times R} \end{aligned} \quad (5)$$

2.4.RO cost model

The LCOW of reverse osmosis (RO) was calculated using a model and configuration that was described in previous works (Mistry et al., 2011; Nayar et al., 2017; Thiel et al., 2015). The RO configuration selected (shown in **Fig. 2**) was a conventional 1-stage arrangement with a circulation

pump, a high pressure pump, a pressure exchanger and a booster pump. The RO model used seawater thermophysical property correlations developed by Nayar *et al.* (Nayar et al., 2016) to determine seawater osmotic pressure and density. The pinch pressure in the RO module (i.e., the difference between hydraulic pressure and peak module osmotic pressure) was assumed to be 10 bar with a pressure loss of 2 bar assumed in the RO module. The pumps were assumed to have an efficiency of 0.8 while the pressure exchanger was assumed to have an efficiency of 0.96 (Mistry et al., 2011). The energy requirements were determined by basic equations for pump work, described in Thiel *et al.* (Thiel et al., 2015) and Nayar *et al.* (Nayar et al., 2017). The capital cost of the RO system was assumed to scale as \$1206 day/m³ of RO water production (Nayar et al., 2017). This value was an average value based on seawater RO plants from around the world, sourced from DesalData (“Cost Estimator - DesalData (2017),” 2017). The capital cost was annualized using the same factor described in Section 2.3.

3. Results and discussion

3.1. Estimation of electrical energy consumption

ICP desalination with high salinity brine (salinity ≥ 35 g/kg) was performed with a fixed device configuration (i.e., fixed channel dimensions, cell numbers, and recovery ratio). Feed salinity, salt removal ratio, and flow velocity were varied, and the resulting current and potential drop, from which the power consumption was calculated, were measured. The power consumption was plotted as a function of the applied current over a variety of operating parameters. The results show a power-law relationship between the current and the power consumption (**Fig. 3**), which is described by $y = 0.416x^{1.86}$ with $r^2 = 0.985$, where $y = P$ in Watts and $x = I$ in amperes. The fitting equation indicates that the relation is similar to the Ohm’s law ($P = I^2R$) with a constant resistance R in the system. The total electrical resistance in the ICP desalination system includes

contributions from the membrane, the bulk solution, the diffusion boundary layer, and the electrical double layer (Długolecki et al., 2010). At high salinity above seawater level, the solution is highly conducting, so the membrane resistance becomes the main contributor to the total resistance (Długolecki et al., 2010). Therefore, the total resistance at high salinity is almost constant because the membrane resistance sets a lower limit for the resistance.

To estimate the electrical energy consumption from the power relation, the current must be calculated for the given operating conditions. The current is generally proportional to the rate of ion removal, but we must consider the current efficiency (CE), which represents how much of the total current is utilized for ion removal, as the CE varies with operating conditions. The *CE* trends as a function of the salt removal ratio and flow velocity are shown in **Fig. 4(a)** and **(b)**, respectively. When other parameters are set constant, the *CE* decreases with the salt removal ratio because higher ion removal requires a higher current density and thus thicker ion depletion layers, which cause undesirable phenomena, such as leakage of desalination, loss of membrane perm-selectivity or water splitting, reducing the performance of ICP desalination (**Fig. 1(b)**) (Nayar et al., 2017). The CE is a function of flow velocity, as well as the SRR, which is shown by the range of CE values for the given SRR in **Fig. 4(b)**. The *CE* exhibits a quadratic relation with linear flow velocity and peaks around 2 mm/s. This trend is a result of two effects: first, a high flow velocity reduces the thickness of ion depletion layers and enhances the *CE* (by minimizing the loss of depletion region into the middle channel); second, a high flow velocity requires a high ion removal rate to achieve the same salt removal ratio, leading to a high current density. High current density (high *SRR*) are shown to precipitously decrease the *CE*, due to many non-ideal behaviors of electromembrane separation (Luo et al., 2002; Yu et al., 2003).

A similar result can be also found in other works (Długołęcki et al., 2009) that the concentration polarization was greatly mitigated at flow velocity from 0 to 3 mm/s, but it was not much at flow velocity higher than 3 mm/s. The combined effect of the salt removal ratio (or operating current) and the flow velocity on the *CE* was captured by a multivariate fitted model based on an extensive set of experimental data operation membranes at various current and flow conditions. In **Fig. 4(c)**, the fitted model was plotted as a surface, and the experimental data were plotted as dots. Using this model, we can estimate the *CE* for a given combination of feed salinity, product salinity, and flow velocity. The *CE* was used to calculate the required current and the resulting power and energy consumption. For feed salinity beyond 100 g/kg (application 2 & 3, shown later) this model for current efficiency is also applied to estimate the power consumption and corresponding total cost by extending the current efficiency plot at salinity up to 200 g/kg. The accuracy of the current estimation was evaluated by calculating the absolute percentage deviation (equation below) of the estimated current from the measured current for the data set (**Fig. 5**). The average absolute percentage deviation, defined as follows, was 9.5%. This deviation mainly resulted from the difference between of the fitted current efficiency and the actual current efficiency value. The *CE* fitting model describes the general trend of *CE* as a function of the operating parameters described above, but it becomes more challenging to estimate the exact *CE* when more ions are involved (*i.e.*, higher SRR, concentration, and velocity), indicated by the larger deviation at higher current, and when the concentration polarization effect is more significant.

$$\text{Absolute percentage deviation} = \frac{|I_{\text{estimated}} - I_{\text{measured}}|}{I_{\text{measured}}} \times 100\% \quad (6)$$

3.2. Application 1: Pre-treatment of high salinity brine

The first application scenario of ICP desalination is partial desalination of high salinity brine to the typical seawater level (35 g/kg). Various sources of brine, or other feed waters, having salinities higher than 35 g/kg include: high salinity seawater in the Arabian Gulf (up to 50 g/kg), concentrate from RO/MSF desalination plants (60 to 70 g/kg), and produced water from oil/gas extraction processes (salinity widely varying with levels sometimes greater than 200 g/kg). As discussed in the Introduction, only limited options exist for treatment of this brine. Here, we performed a cost analysis to evaluate the economic feasibility of an ICP-RO hybrid system. ICP desalination is used to bring down the salinity of a brine source to 35 g/kg, a level at which a follow-on reverse osmosis system can operate efficiently. Based on the experimental data, the electrical energy was estimated as described in Section 3.1, and the pumping energy was calculated to obtain the total energy consumption. For a given set of the feed and product salinity, the flow velocity, and hence the applied current, was varied to obtain the energy consumption as a function of the current density.

In **Fig. 6(a)**, we show a case of partial desalination from the feed salinity of 50 g/kg to a product salinity of 35 g/kg. **Fig. 6(b)** shows a typical relation between the electrical and the pumping energy and the current. A high current leads to a greater electrical power and energy consumption. It also requires an increased flow and volumetric flowrate for the same salt removal ratio. Consequently, the pumping energy increases with the current. However, the ICP desalination generally operates with a low flow velocity, which is much slower than in electrodialysis, leading to very low pressure drops and a negligible contribution of pumping power to the total energy consumption. The energy consumption was used to calculate the operating cost; combined with the capital cost based on the required membrane area, the LCOW is plotted as a function of the current density in **Fig. 6(c)**. At a fixed salt removal ratio, a high current result in a fast flow and reduces the membrane area required, shown by the decreasing capital cost. Considering the contributions from the operating

and the capital costs, the optimal current, electrical energy consumption, and LCOW were determined to be 330 A/m², 6.5 kWh/m³-ICP-diluate, and \$1.4/m³-ICP-diluate, respectively.

We analyzed the LCOW of the partial desalination for a range of feed salinity and a fixed product salinity of 35 g/kg (**Fig. 7**). The LCOW increases with the feed salinity mainly because of the large amount of ion removal required. For similar reasons, the optimal current density increases with the feed salinity (**Fig. 7(a)**), as a high ion removal rate per membrane area is more cost efficient at high salinity. **Fig. 7(b)** shows the breakdown of LCOW to the operating and the capital costs. As the feed salinity increases, the operating cost accounts a greater portion of the LCOW due to the increasing importance of high salt removal rate. For a feed salinity of 75 g/kg, the optimal electrical energy consumption and the LCOW were 26 kWh/m³-ICP-diluate and \$4.60/m³-ICP-diluate, respectively. These values are similar to a comparable electrodialysis performance (feed and product salinity of 90 g/kg and 40.7 g/kg, respectively), which resulted in the electrical energy consumption of 20 – 22 kWh/m³ of diluate stream and a LCOW of ~\$5.5/m³ of diluate stream (McGovern et al., 2014a). Currently, complete treatment of high salinity brine can be done efficiently via MVC technology. The MVC cost for shale gas produced water treatment is \$22 – 39/m³ of product water (Slutz et al., 2012). Based on our cost analysis, the ICP partial desalination in combination with reverse osmosis (< \$2 /m³ of pure water) shows potential to be economically competitive for high salinity produced water treatment when compared to the current MVC technology.

Next, the LCOW from the ICP-RO hybrid process was analyzed. Here, partial desalination with ICP desalination is followed by complete desalination (*i.e.*, product salinity is 0 g/kg) with RO (see **Fig. 8(a)**). The feed and the product salinities were fixed; the feed salinity to the RO step, or the product salinity in the ICP desalination, was varied to determine the RO feed concentration for

an optimal ICP-RO operation. The ICP recovery ratio was kept at 50%, while the RO recovery ratio was varied with the feed salinity to result in a concentrate stream concentration of 70 g/kg. **Fig. 8** shows the ICP-RO water cost for the starting salinity (*i.e.*, the feed to the ICP-RO process) of 50 g/kg and 75 g/kg. The starting salinity of 50 g/kg represents the highest seawater salinities practically encountered in the northern parts of the Arabian Gulf while the choice of 75 g/kg reflects a typical salinity of brine from seawater desalination plants. When the RO feed concentration equals the starting salinity, the ICP desalination step was omitted, and the LCOW was calculated for the sole RO treatment. **Fig. 8(b)** shows that the total water cost decreases with a higher RO feed concentration and reaches a minimum for the solely RO operation. Generally, as the RO feed concentration increases, the LCOW of RO increases, and the LCOW of ICP desalination decreases. But in the case of 50 g/kg feed, the RO cost is much smaller than the ICP desalination cost for the entire range of the RO feed concentration, so the LCOW is minimized when the ICP desalination is not used. Furthermore, when ICP desalination is used to pretreat the RO feed with the ICP desalination recovery of 50%, the total volume of desalinated water was reduced to half, raising the water cost per volume of desalinated water. It has been shown that the LCOW of ICP desalination increases with increasing recovery (Kim et al., 2016), while the LCOW of RO should be independent of the ICP recovery. From our cost analysis, the RO-only desalination is more efficient than the ICP-RO hybrid for feed salinity up to 50 g/kg. However, practical seawater RO systems are used to treat feeds with salinities up to around 50 g/kg with the maximum operating pressure for the systems being around 68 bar, which corresponds to concentrate side concentration of 73 – 74 g/kg (Fritzmann et al., 2007). The maximum operating pressure of conventional seawater RO is guided by concerns around water costs, efficiency and membrane pressure limitations (Gottberg et al., 2017). The pressure limit in turn limit the brine

and feed salinity levels. The recovery rate of the RO-only desalination can be achieved at 28.6% at the feed salinity of 50 g/kg, but ICP-RO hybrid can increase the recovery rate to 42.9%. For treating feed salinities greater than 50 g/kg, the ICP-RO hybrid can be a cost-efficient solution.

We analyzed the ICP-RO water cost for a starting salinity of 75 g/kg, which can represent a concentrate stream from desalination seawater plants. As shown in **Fig. 8(c)**, the water cost follows a similar trend as in **Fig. 8(b)**, but the minimum cost is obtained when the ICP/RO transition concentration is 50 g/kg. This is because at a very high RO feed concentration, the RO recovery is reduced drastically, hence greatly increasing the specific cost of both ICP and RO. The water cost of \$8-9 per m³ of the pure product is much lower than the water cost of \$22-39 per m³ of product water for MVC, which is the current technology for treating brine at this concentration. The relatively low water cost of ICP-RO hybrid indicates that this technology may be an economically viable method for treatment of brine at 75 g/kg.

3.3. Application 2: Brine concentration for salt production

The second application is the concentration of brine for salt production. In most brine management practices, the diluate stream is considered the useful product. However, the concentrate stream can also be a useful product. For salt production applications, the concentrated brine stream is fed in to a crystallizer where salt is crystallized out (Turek, 2003). Crystallizers are expensive and so typically, seawater is concentrated first using desalination technologies to reduce the load of the crystallizer in salt production and reduce the overall cost (Chung et al., 2017). Chung *et al.* had compared the performance of state-of-the-art crystallizers to state-of-the-art brine concentrators and found that there was a greater scope for improving the energy efficiency of brine concentration (Chung et al., 2017). This led us to investigate the potential of ICP desalination for use as brine concentrator.

Fig. 9(a) describes a multi-stage ICP desalination for brine concentration. The starting salinity of 70 g/kg represents a typical brine concentration from seawater desalination plants, and the final salinity of 200 g/kg represents a typical feed to crystallizer in salt production plants. Since the ICP desalination has a maximum recovery of 50%, a single-stage ICP process can achieve a maximum of two-fold concentration, and the concentration from 70 g/kg to 200 g/kg needs to be performed in multiple stages. In the multi-stage ICP desalination scheme, the concentrate stream from each stage is fed to the next stage for further concentration, with the recovery of 50% at each stage. The LCOW is expressed as cost per final concentrate volume, and the optimized water cost will be converted to cost per unit salt mass. In the three-stage ICP process shown in **Fig. 9(b)**, the water cost for a later stage is higher than the earlier stage because the later stage is working at a higher concentration. The results in **Fig. 9(c)** show that the minimum water cost is obtained for the three-stage system, which consists of the optimized number of stages. The addition of ICP stage has the effect of lowering electricity cost for each stage, with the decrease in total electricity cost for the system. The increase in stage number reduces the salt removal ratio for each stage, preventing the ICP desalination from operating in a high current density condition, which causes inefficient salt concentration. However, the capital cost increases with an increasing number of stages. The optimal water cost is \$21.7/m³, which corresponds to \$95/tonne salt produced. Adding the crystallization cost of approximately \$40/tonne of salt(Nayar et al., 2017), the total salt production cost from the ICP-crystallizer is \$135/tonne salt. The cost of brine concentration by multi-stage ICP desalination is about twice of the brine concentration cost by electrodialysis (concentration from 35 g/kg to 200 g/kg), which is around \$60/tonne salt(Nayar et al., 2017). The large gap in the cost from the two processes is most likely to stem from that the recovery in ICP desalination is limited to 50% at each stage, whereas the recovery in ED is much higher. Furthermore, the ED

cost is obtained from a working salt production plant, where the process is well engineered and optimized, and thus the cost may be further lowered.

3.4 Application 3 : Brine desalination for reuse

The third application considered is the desalination of brine waste from the oil industry. The oil industry generates highly concentrated brine waste, ranging from 100 to 200 g/kg (Al-Shamari et al., 2013; Arthur et al., 2005). Because of the high salinity, the brine waste is disposed through underground injection or evaporation in a surface pit, with the potential for environment problems (AlAnezi et al., 2013). Therefore, the concentration of brine should be reduced (via partial desalination) to a level appropriate for disposal or reuse, or to the level amenable to other desalination technologies (around seawater level).

The schematic illustration of multi-stage ICP desalination in **Fig. 10(a)** shows similar configuration as the one shown in **Fig. 9(a)** for brine concentration. However, the brine desalination utilizes the diluate stream as the inflow for the next stage. The final diluate salinities for brine desalination are 10, 25, 40 g/kg, which represent salinity levels adequate for several reuse scenarios, such as Enhanced Oil Recovery, Pressure-Retarded Osmosis (PRO, 25 g/kg) and RO (40 g/kg), respectively (Huang et al., 2018; Piemonte et al., 2015; Royce et al., 1984; Shafer, 2011; Shaffer et al., 2013). In order to find the minimum LCOW, we calculate every single water cost with 1 % SRR increment for the multi-stage desalination from 1-stage to 4-stage. The SRR for the next stage is determined by the diluate salinity of the previous stage, while the final diluate salinity is fixed. **Fig. 10(b)** and **(c)** indicate the LCOW changes in two-stage and three-stage ICP desalination, respectively. Since the final diluate salinity determined by SRR of stage 2 (SRR-2), the final diluate salinity decrease requires increase in ‘SRR-2’ which gives increase in operating cost. The ‘SRR-1’ for the minimum LCOW increases with the final diluate salinity decrease.

Increase in 'SRR-1' facilitates to reduce the LCOW for stage 2 as providing a lower salinity feed water for stage 2, because higher salinity and 'SRR-2' cause loss of membrane permselectivity and undesirable chemical reactions, resulting in poor energy efficiency. (Kim et al., 2016) As a result, SRR-1 becomes a more dominant factor for a lower final diluate salinity. **Fig. 10(d)** indicates the minimum LCOW for 10 (d-i), 25 (d-ii) and 40 g/kg (d-iii) with various number of stages in system. Most of the LCOW for multi-stage arises in the final stage because most of salt is removed at the last stage. The total operating cost decreases slightly with increasing number of stages; as explained in the salt production application (Section 3.3) an additional stage reduces the energy consumption by distributing the energy load for salt removal to more stages. On the other hand, the capital cost is usually lowest for the system with 1 or 2 stages and increases significantly as the number of stages increases. The minimum LCOW is obtained for the two-stage system, \$25.9, \$20.5 and \$16.4 /m³ for 10, 25 and 40 g/kg, respectively.

The brine treatment using multistage ICP has achieved a competitive cost (\$16.4 ~ 25.9 m³) due to the cost flexibility associated with the salinity of the produced water compared to brine treatment costs with current technologies. The brine treatment costs by current technologies are as following : Multi-effect distillation or vapor compression distillation can obtain a pure water from brine (100 ~ 200 kppm) at a cost of about \$19 ~ 31 /m³ (Karapataki, 2010). Mechanical vapor recompression can treat a brine with 80,000 to 240,000 mg/L at cost \$22 ~ 39 /m³ (Slutz et al., 2012). The multistage batch-ED desalted the brine sequentially from 195 kppm to 0.24 kppm at a cost of \$29.35 /m³ (McGovern et al., 2014a). The cost of a crystallizer for zero liquid discharge is widely distributed from \$0.66 /m³ to \$26.41 /m³ depending on the feed salinity (Greenlee et al., 2009).

4. Conclusions

In this work, we utilized a set of experimental data from a lab-scale ICP desalination device to build a model that estimates the electrical energy consumption of ICP desalination at high salinity. This model was then used to calculate the operating cost, and the capital cost was calculated based on the effective membrane area required to reach the throughput. For a fixed feed and product salinity, the LCOW was calculated over a range of current, and the optimal operating conditions. This LCOW analysis was performed for three potential applications of ICP desalination in high salinity brine treatment.

The first application was partial desalination to seawater salinity (35 g/kg) so that the product can be treated by seawater reverse osmosis to make pure water. The cost analysis showed that the ICP partial desalination cost with feed salinity of 75 g/kg was comparable to electrodialysis performance at similar feed and product salinity. This cost is much lower than the complete desalination by MVC, a current desalination method for high-salinity produced water. Therefore, brine treatment by ICP desalination in combination with seawater RO at seawater levels of feed salinity can become an economically competitive option for high salinity brine desalination. The cost analysis of an ICP-RO hybrid shows that for the feed salinity range, at which RO is normally available, the ICP-RO hybrid is not economically feasible, mainly due to the reduced recovery in ICP-RO. The RO cost was relatively insensitive to increasing feed salinity (over the range of 10 – 50 g/kg). However, the ICP-RO hybrid can be an economically viable solution for treatment of brine from seawater desalination plants (~75 g/kg), which is at a concentration above the operational limit of conventional seawater RO.

The second application considered the concentration of brine (to 200 g/kg) by a multi-stage ICP process, for salt production. We varied the number of stages and found that the operation becomes more efficient when the number of stages is minimized. The optimal three-stage system resulted

in the LCOW of \$135/tonne salt produced. This cost is about twice the cost of brine concentration by ED. The high ICP cost can be contributed to the large volume of treatment required in early stages due to the ICP recovery limit of 50% per stage. To reduce the ICP cost for brine concentration, the salt retention in the concentrate stream should be enhanced, and the distribution of salt concentration (or salt removal) ratio at the stages can be optimized.

The third application was the desalination of brine (from 160 g/kg) by a multi-stage ICP process for reuse of brine waste from the oil industry. We applied a similar multi-stage ICP process with the second application, but the stream was sequentially desalinated. With the increase in the stage number, an additional stage reduced the operating cost with reduction in energy consumption by redistributing the energy load for salt removal. ICP desalination has the advantage of easily enabling partial desalination, which is the feature shared by electrodialysis, obtaining \$25.9, \$20.5 and \$16.4/m³ of the minimum LCOW to have 10, 25 and 40 g/kg of produced concentration, respectively. Yet, compared with electrodialysis, ICP system has more flexibility in dealing with fouling, since only more fouling-resistant CEMs can be utilized.

The concept of ICP desalination was introduced with microfluidic technology (Kwak et al., 2016), and then lab-scale ICP desalination was developed to demonstrate applicability to industries through increased throughput. In this paper, the economic feasibility of ICP desalination processes have been assessed with three different brine treatment applications. The feature of partial desalination by ICP desalination facilitates a variety of brine treatment scenarios, providing minimized LCOW with the optimized stage number. However, ICP desalination still suffers from a fixed recovery rate and a relatively high power consumption for brine treatment. In the future, both experiments and modeling should be employed to improve the economic feasibility, achieving better energy efficiency and improved the recovery rate.

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739 **Table 1.** Variables and parameters for cost analysis

Symbol	Description	Value	Unit
I	Current		A
V	Voltage		V
Q	Feed flowrate		L/min
K_E	Electricity rate	0.1	\$/kWh
μ	Dynamic viscosity	0.00972	g/(cm·s)
L	Cell length	30	cm
w	Cell width (effective)	8	mm
d	Cell depth (<i>i.e.</i> , intermembrane distance)	2.5	mm
Q_{total}	Total flowrate		L/min
Q_{cell}	Flowrate per cell		L/min
A_m	Total membrane area		m ²
K_Q	Capital cost per unit membrane area	750	\$/m ²
R	Cost of capital	10	%
T	Life time of equipment	20	year

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Fig. 1 (a) Schematic illustration of ICP desalination. Descriptions of ion movement (solid lines) and flow stream lines (dashed lines) under (b-i) low and (b-ii) high current density.

Fig. 2 Configuration of a 1-stage seawater RO plant (CP – circulation pump, HP – high pressure pump and BP – booster pump)

Fig. 3 Power relation between power consumption and applied current in ICP desalination. The experimental data were generated with a fixed device configuration over various parameters (*e.g.*, feed salinity, salt removal ratio, and flow velocity). The fitted equation is in the form of $P = a \times I^b$ where P and I are power consumption and current, respectively.

Fig. 4 Current efficiency as a function of salt removal ratio (SRR) and average flow velocity. The experimental data were generated with a fixed device configuration. (a) Current efficiency as a function of SRR, for a range of feed salinity and flow velocity, follows a linear relation. (b) Current efficiency as a function of average flow velocity, for a range of SRR and flow velocity, follows a quadratic relation. Only the data for 10%, 50%, and 70% were plotted to easily show the quadratic trend. (c) Current efficiency was plotted as a function of SRR and flow velocity, and the data was fitted to a polynomial surface.

Fig. 5 The estimated current using the CE fitting model for given feed salinity, SRR, and flow velocity. The (absolute) percent deviation of the estimated current from the measured current was plotted.

Fig. 6 Water cost optimization for partial desalination from feed salinity of 50 g/kg to diluate salinity of 35 g/kg. Recovery ratio (RR) = 0.5. (a) Schematic of ICP partial desalination used in **Fig. 1(a)**. The diluate salinity is fixed to 35 g/kg. (b) Electrical and pumping energy as a function of current. (c) LCOW as a function of current. Total cost is the sum of operating (electricity) cost and capital cost. All costs are expressed as U.S. dollar per diluate volume (m³).

Fig. 7 Water cost optimization for partial desalination for various feed salinity. Diluate salinity is fixed to 35 g/kg (seawater level). $RR = 0.5$. Total cost is the sum of electricity cost and capital cost. All costs are expressed as U.S. dollar per diluate volume (m^3) (a) LCOW as a function of current for various feed salinity. (b) Optimal LCOW for a range of feed salinity, broken down into contribution from operating and capital costs.

Fig. 8 ICP-RO cost analysis. (a) Schematic diagram of the ICP-RO hybrid. ICP desalination was performed on an ICP feed salinity of 50 g/kg (b) and 75 g/kg (c), and the ICP diluate stream was taken for complete desalination by RO. Hence, the RO feed concentration is equal to the diluate concentration of ICP. In all cases, the recovery in ICP was 50%, and the recovery of RO was varied to give a fixed concentrate stream concentration of 70 g/kg. When the RO feed concentration equals the overall feed salinity, the water cost is for a RO-only process, whose recovery is double of the ICP-RO hybrid.

Fig. 9 ICP desalination for brine concentration. (a) Schematic diagram of multi-stage ICP desalination for brine concentration. N is the number of stages. In all cases, the salt removal ratios for all stages in a system were about the same. (b) Water cost for 3-stage system, broken down by stage. (c) Water cost for systems with various stage numbers.

Fig. 10 ICP desalination for brine desalination from oil industry. (a) Schematic diagram of multi-stage ICP desalination for salt removal. N is the number of stages. Water cost variations for (b) 2-stage system with respect to SRR change at stage 1 and (c) 3-stage system with respect to both SRR change at stage 1 and stage 2.

Fig. 11 Minimum brine treatment cost from oil industry with the ICP desalination system. Various stage numbers, from 1 to 4, are evaluated to achieve fixed final dilute concentrations, (a) 10, (b) 25 and (c) 40 g/kg.