*Desalination and Water Treatment*
www.deswater.com

doi: 10.5004/dwt.2020.25598

189 (2020) 89–97
June

# Techno-economic comparison of pilot-scale EDI and BWRO for brackish water desalination

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Received 30 May 2019; Accepted 1 February 2020

**ABSTRACT**

In this paper, a pilot-scale electrodeionization (EDI) was designed and operated for brackish water desalination to produce drinking water. Technical and economic feasibility of pilot-scale EDI was investigated and compared to recently installed brackish water reverse osmosis (BWRO) mini-plant with equal production capacity. It was found that EDI could remove more than 98% of salt ions from brackish water at the optimum operating condition (45 V). From an economic point of view, EDI was economically potential as an alternative to the BWRO process which offered a lower capital investment (US$ 5076) than the existing BWRO plant (US$ 8447). The specific water production cost of the EDI process was US$ 0.50/m$^3$ which was comparable to the specific cost of BWRO process, US$ 0.42/m$^3$. The economic analysis from a long-term perspective shows that EDI is more economical than the BWRO up to around 20 y of operation. These results indicate that EDI is an attractive alternative to BWRO for brackish water desalination.

*Keywords:* Brackish water; Desalination; Drinking water; Economic; Pilot plant

## 1. Introduction

Commercially available technology for brackish water desalination can be generally classified into thermal- and membrane-based processes. The membrane is considered as a cost-effective process since it is less energy-intensive compared to the thermal processes [1]. With the advantages of the lower footprint, modularity, easy to operate, and more intensive process, membranes have started to replace conventional technologies in various fields [2–6]. Among membrane-based processes, reverse osmosis (RO) is the most widely used which accounts for 65% of the total installed plants in the world [2,7]. In RO, a pressure difference is continuously introduced into the system to overcome the osmotic pressure of the feed solution and to drive pure water transport through the selective membrane while salts are rejected. Supported by intensive research, advanced RO membrane which has a high salt rejection (up to 99.99%) is commercialized [8]. However, RO possesses major drawbacks such as high electric energy requirement and the needs of specific material for piping and housing associated with the high-pressure operation [9,10].

Electrodeionization (EDI) has been considered as a potential alternative for RO since it offers low-pressure operation [11,12]. EDI removes ions by selective transport through ion-exchange membranes under the influence of the electric field. Unlike the RO process which transports the large portion of fluid component (water), EDI transports salt ions which are the minor components. Therefore, EDI may have higher energy efficiency [11]. The main application of the EDI process is ultrapure water production for microelectronic and pharmaceutical industries [13–16]. However, the literature on EDI application for high salinity water is still lacking and needs more investigation. Moreover, the commercial EDI module also specifies low conductivity feed water (<40 μS/cm).

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The first application of EDI on brackish water desalination has been reported by Larchet et al. [17] where the effect of module construction was investigated. Further, Sun et al. [18] successfully produced drinking water from brackish water by using EDI with periodically reversed polarities. In this work, a pilot-scale EDI was designed and operated for brackish water desalination. The technical and economic feasibility of EDI was evaluated and compared to recently installed brackish water reverse osmosis (BWRO) mini-plant with equal production capacity.

## 2. Materials and methods

Pilot-scale EDI and BWRO mini-plant are presented in this section, including specifications of the plants, desalination methods, and analysis. Both EDI and BWRO systems were equipped with a sand filter and ultrafiltration (UF) and cleaning-in-place (CIP) facility. Commercial EDI and BWRO units were used in the plants.

### *2.1. Pilot-scale EDI specification, brackish water desalination by EDI, and analysis*

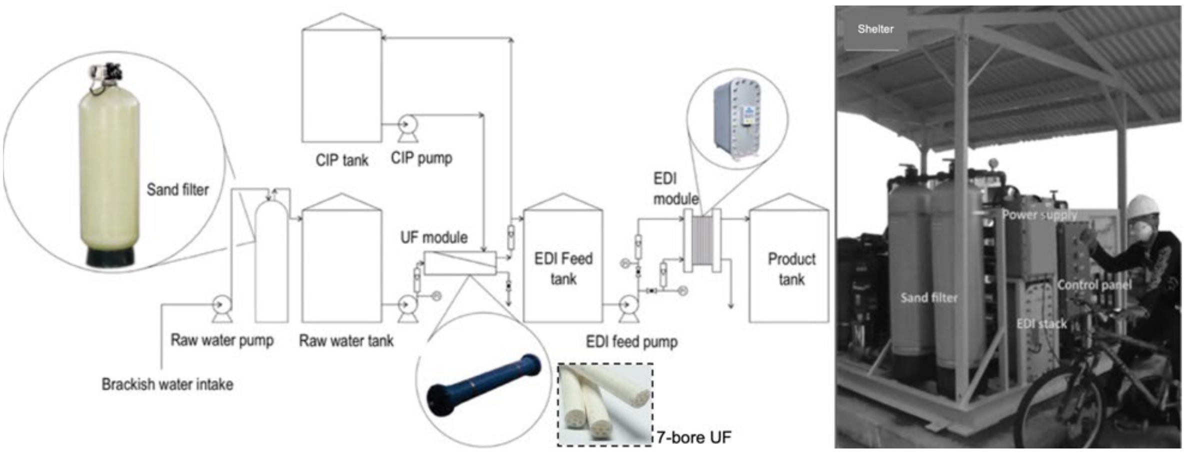
The process flow diagram and the photograph of pilot plant EDI are shown in Figs. 1a and b, respectively. Specifications of components used in the pilot plant are summarized in Table 1.

EDI plant took feed water from well water intake. Sand filter and non-modular UF membrane (supplied by GDP Filter, Indonesia) were used as pre-treatments. The UF was a multi-bore (7 bores) polysulfone-based capillary membrane. The pre-treatment step was used to remove unwanted components, such as particulate matters, colloidal components, etc. that may result in plugging or increasing pressure drop in EDI compartment. The pre-treatment also helped to remove uncharged components and organic matter that cannot be removed by EDI. Furthermore, a portion of bivalent substances can be reduced by the UF membrane [19]. Here, the UF membrane was designed to obtain 80% water recovery. The CIP system consisted of a CIP tank and a CIP pump for periodic cleaning. The periodic cleaning was required to remove foulant from the UF membrane surface as well as to the flux of UF. The CIP tank was filled with UF product or permeate. UF cleaning, that is, backwash or sometimes chemically enhanced backwash, was conducted when the normalized flux reduced to 10%–15% from its initial flux. An alkali solution (NaOH) with a concentration of 0.1% (w/v) was used for chemically enhanced backwash, while a general backwash was performed by flowing UF permeate from permeate side or shell side of the UF module to the feed side. Both the general backwash and chemically enhanced backwash were performed for 2 min at 0.15 MPa pressure. Chemically backwash was performed if the general backwash was ineffective to recover the flux.

An all-filled EDI module (from Zhejiang DD Water Industry Co. Ltd., Zhuji, Zhejiang Province, China) was designed to produce 1 $m^3$/h diluate. In this type of EDI, all

Table 1
EDI plant specification

| UF specifications | Values |
|---|---|
| UF module | GDP filter non-modular UF, 8 inch (0.2 m) × 2 m |
| Membrane material | Polysulfone |
| Average pore size (nm) | 10 |
| Area ($m^2$) | 120 |
| Recovery (%) | 80 |
| Operating pressure (MPa) | 0.15 |
| EDI specifications | Values |
| EDI stack | DDWI electropure EDI-P&F-4.0-HP |
| Capacity ($m^3$/h) | 1 |
| Recovery (%) | 75 |
| Operating pressure (MPa) | 0.15 |
| Interconnecting pipe, valve, and fitting | PVC |




(a) (b)

Fig. 1. EDI for brackish water desalination. (a) Process flow diagram and (b) photograph of pilot-scale EDI.

compartments including diluate and concentrate were filled with ion-exchange resins. The all-filled EDI provides the advantage of no brine circulation for the concentrate compartment. The brine circulation in the concentrate stream was used in an early EDI design in which the EDI module only used ion-exchange resin beads for the diluate compartment [15]. The brine circulation was employed to increase the conductivity of the concentrate stream which was also intended to improve the overall module conductivity. In a new EDI design (Fig. 2), all of the compartments were filled with ion-exchange resins. This new EDI design eliminates brine make-up and its associated equipment. Therefore, the system was simpler rather than the diluate-filled EDI type where ion-exchange resins are used only in the diluate compartment. In this system, the EDI unit was operated to achieve a total recovery of 75%. The relatively low water recovery was designed due to the high concentration of the brackish water. The concentrated water was then discharged. Generally, EDI is operated under a relatively high-water recovery that is, 85%–95%. Sometimes, the concentrated solution is recycled into a raw water tank to achieve an overall water recovery of 100%. However, this is only possible when EDI is operated under a very dilute solution such as a permeate of the RO membrane, especially in ultrapure water production. Direct current (DC) power for EDI was supplied from DC inverter which equipped with a manual variable voltage regulator. EDI control panel used an electrical interlock system to prevent overcurrent. The diluate concentration was analyzed by using digital total dissolved solids (TDS) meter (TDS meter). The trip logic system was activated when either over current or product water off-spec. Since the EDI system required a low operating pressure, polyvinyl chloride (PVC)-based piping system was sufficient.

TDS of feed and diluate of EDI was probed by using TDS-3 (from HM digital Taiwan). The removal efficiency of EDI was calculated by Eq. (1).

$$\text{Removal efficiency}\ (\%) = \frac{(\text{Feed TDS} - \text{Diluate TDS})}{\text{Feed TDS}} 100\% \quad (1)$$

The specific energy consumption (SEC) of EDI, W ($kWh/m^3$), was calculated by

$$W = \frac{E \cdot I}{Q} \quad (2)$$

where $E$ is applied voltage (V), $I$ is electrical current (A), and $Q$ is diluate flow rate ($m^3/s$).

*2.2. BWRO mini plant specification and brackish water desalination by BWRO*

The process flow diagram and the photograph of BWRO mini-plant are shown in Figs. 3a and b, respectively, while BWRO technical specification is summarized in Table 2. The BWRO mini plant consisted of UF and RO membranes. The brackish water intake was supplied from two water wells located approximately 100 m from the plant location. Every well was equipped with a single jet pump which was capable of delivering 3 $m^3/h$ brackish water. The intake system was designed to provide 2.5 $m^3/h$ for BWRO mini-plants while the rest was delivered for domestic applications such as plant watering and washing. The intake system was equipped with a back-washable 300 mesh strainer located at the discharge of a UF feed pump to protect the UF membrane. The UF system with a capacity of 2 $m^3/h$ and 80% recovery was used as pre-treatment which consisted of a single 8-inch (0.2 m) module with an effective length of 2 m. This type of UF membrane was chosen since it has a high flux and better mechanical strength [20].

The BWRO mini-plant was designed to achieve a total capacity of 1 $m^3/h$ with product TDS of <500 ppm. The

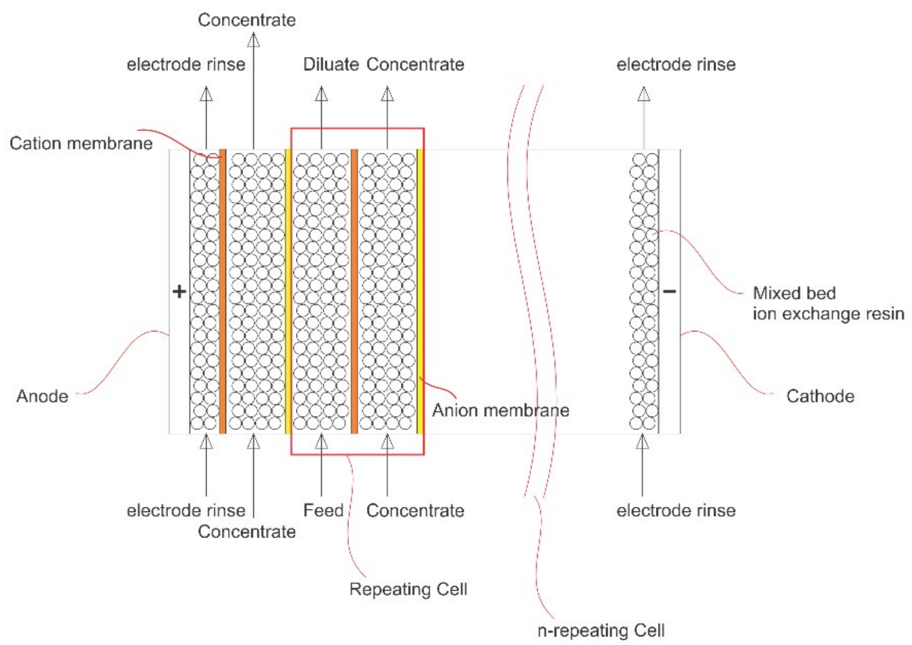
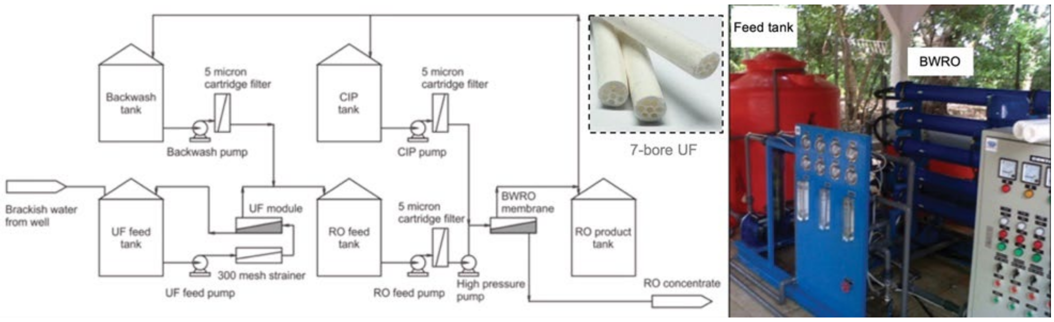



Fig. 2. Schematic of EDI configuration used in this study.




(a) (b)

Fig. 3. BWRO mini plant for brackish water desalination. (a) Process flow diagram and (b) photograph of BWRO plant.

Table 2
BWRO min plant specification

| UF specifications | Values |
|---|---|
| UF module | GDP filter non-modular UF, 8-inch × 2 m |
| Membrane material | Polysulfone |
| Average pore size (nm) | 10 |
| Area ($m^2$) | 120 |
| Recovery (%) | 80 |
| Operating pressure (MPa) | 0.15 |
| Capacity ($m^3$/h) | 2 |
| **BWRO specifications** | **Values** |
| RO module | CSM RO, RE-4040 BE |
| Nominal salt rejection | 99.7% |
| Total capacity ($m^3$/h) | 1 |
| Recovery (%) | 50 |
| Operating pressure (MPa) | 2.2 |
| Interconnecting pipe, valve, and fitting | SS 316 for high-pressure line and PVC for low-pressure line |

BWRO consisted of a single-stage configuration of five pressure vessels with five total elements. The membrane element has four in. (0.1 m) diameter and 40 in. (1.02 m) length. The BWRO unit was designed to achieve a water recovery of 50%. Cartridge filter with micron ratings of five microns was installed at the discharge of the BWRO feed pump to protect the RO element from dirt or particle from BWRO feed tank or piping. The system was also equipped with an online cleaning system for example, backwash and CIP system. The UF membrane was backwash every 4 h of continuous operation for 5 min using BWRO product while CIP of BWRO was based on a 15% drop in normalized permeate flow. The BWRO mini-plant was designed for manual operation. The high-pressure switch was installed in a high-pressure pump discharge to prevent overpressure in the RO system with a maximum allowable feed pressure of 2.5 MPa. The system is also equipped with a level switch in every intermediate tank to prevent overflow. Overload protection is installed to prevent overcurrent at every pump. Overpressure, low-level feed tank, and high-level product tank will cause total system shutdown while overcurrent will cause a partial shutdown of the specific pump.

### 2.3. Economic analysis

To assess the economic feasibility, several assumptions were made such as a continuous operation for 330 d/y, electricity price was US$ 0.038/kWh, and average service life of RO and EDI was 2 y. The capital expenditure (CAPEX) and operating expenses (OPEX) were the economic parameters that were considered.

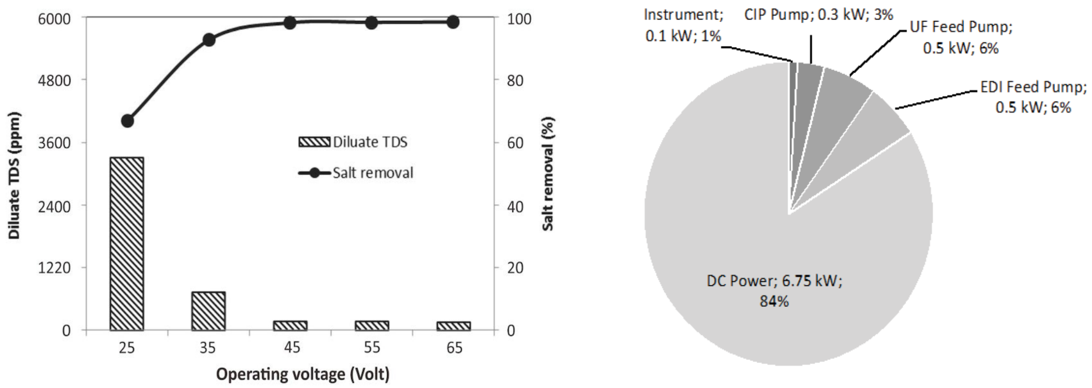
## 3. Results and discussion

In this section, the performance of pilot-scale EDI is firstly discussed, including the effect of operating parameters, practical operating zone map, and energy consumption. The discussion is then followed by the economic evaluation of pilot-scale EDI and BWRO mini plant.

### 3.1. Performance of EDI

The applied voltage is one of the important operating parameters of an EDI process. As the driving force of ionic migration, applied voltage determines the process efficiency as well as the final product quality. The applied voltage should be appropriately chosen to achieve the desired final product at high current efficiency. During the initial test, the applied voltage was varied from 25–65 V to find an optimum operating condition. A summary of the effect of applied voltage on diluate TDS is given in Fig. 4a. According to Fig. 4a, high quality of drinking water can be obtained at an applied voltage of 45 V or higher with a corresponding diluate TDS < 177 ppm.

It can be observed that at the low voltage regime (25–45 v), diluate TDS decreases significantly with the increase of electrical potential difference [21]. A steep increase in ionic migration is observed in this regime. As the applied voltage increases, the driving force for electromigration gets higher which in turn enhances the flux of ions from diluate compartment to the concentrate compartment yielding in the lower diluate concentration. However, above 45 v, the effect of increasing applied voltage on removal efficiency becomes negligible indicated by the insignificant change of




**(a)** **(b)**

Fig. 4. EDI performance at feed TDS of 10,000 ppm and feed flow rate of 1 $m^3$/h. (a) Effect of operating voltage on diluate TDS and (b) power requirement.

ion removal. According to Ohmic law, a current is linear to applied voltage while ionic removal is also proportional to the current. In this regard, the increase in applied voltage will increase the ion removal linearly. However, it is not observed in the high applied voltage implying the lower efficiency of current utilization. This may be due to the excessive water splitting process at a high voltage regime. At this regime, the excessive $H^+$ and $OH^-$ ions generated from water splitting reaction also migrate through the ion exchange membrane [22]. The current efficiency is dramatically reduced due to the extensive utilization of current for water splitting reaction. Also, the generated $H^+$ and $OH^-$ ions also act as charge carriers which consume some portion of the current. Therefore, the removal rate was insignificantly improved when the applied voltage was increased further.

Generally, there are two distinct operating regimes in EDI, namely enhanced electro-migration and electro-regeneration [13]. Electro-regeneration, which causes the formation of $H^+$ and $OH^-$ ions, occurs when the ion concentration is relatively low or at the outlet of the EDI module. When the applied voltage is too high, most of the ions are transferred to the concentrate compartment. Consequently, the concentration of ions at the outlet of EDI is very low. Since the applied voltage is similar along the inlet to the outlet of the EDI module, the high applied voltage at the outlet part of EDI leads to electro-regeneration. Furthermore, a more extensive water splitting may occur in the EDI module due to a larger bipolar point in the EDI compartments, especially for a mixed-bed type EDI [22]. The bipolar point can be found at a contact point between cation- and anion-exchange resin beads and a contact point between resin bead and ion-exchange membrane [22].

It is obvious that applied voltage is directly related to the ion removal rate. Increases applied voltage will lead to a higher product or diluate quality. However, one should note that higher applied voltage results in a larger energy requirement [18]. Therefore, EDI should be operated at a minimum applied voltage which can obtain the desired water quality. Operating EDI at a voltage of 45 V was chosen for normal operation to obtain better removal efficiency at lower energy consumption. This is because operating EDI at this applied voltage can produce diluate with low TDS and less energy consumption than 55 and 65 V.

Even though EDI has been commercially applied in ultrapure water production with an attractive technological and economical point of view, the application of EDI for brackish water desalination is quite challenging. Since EDI uses electrical energy to directly drive ionic migration, the energy consumption will depend significantly on the number of ions that should be transferred. Thus, the energy consumption was estimated to analyze the feasibility of the EDI application for brackish water desalination. The composition of the energy requirement of pilot plant EDI brackish water desalination is depicted in Fig. 3b. It is obvious that most of the power requirement of the process was due to the DC power supply, that is, 84% of the total electric power requirement (Fig. 4b). According to Faraday's Law, energy consumption is proportional to ion concentration, therefore relatively high energy consumption is required for the DC power supply. The presence of ion-exchange resins is expected then to reduce the overall EDI module and to achieve a lower energy consumption than the conventional ED module. Under a normal operation mode, the average electric power requirement throughout the process was 8 kW hence the SEC was 8 kWh/$m^3$. DC power supply for a single EDI stack required 6.75 kWh/$m^3$ while the rest were required energy for pumping. Compared to BWRO, the energy consumption of the EDI system for pumping is lower which is associated with lower operating pressure. The summary of a comparison of energy consumption between EDI and other processes in brackish water desalination is given in Table 3. The EDI application studied here required a larger amount of electrical power for ion transport which might appear disadvantageous in comparison with other brackish water desalination processes such as RO and evaporation. Yet, this application offers less complicated low-pressure technology. Comparing to the EDI process in the literature, the brackish water EDI plant in this report shows a larger SEC due to higher feed water TDS.

The practical range of operating variables should be precisely determined to achieve better process efficiency. The

Table 3
SEC of different membrane based brackish water desalination

| Technology | Capacity ($m^3/h$) | Feed TDS (ppm) | SEC ($kWh/m^3$) | Ref. |
|---|---|---|---|---|
| EDI | 1 | 10,000 | 8 | This study |
| EDI | 0.006 | 2,000–4,000 | 1.04–3.71 | [18] |
| Electrodialysis | – | 2,000 | 0.999 | [26] |
| Photovoltaic powered–hybrid NF/BWRO | 0.155 | 1,500 | 5.02 | [27] |
| Photovoltaic powered–BWRO | 0.171 | 1,500 | 17.6 | [27] |

NF = nanofiltration

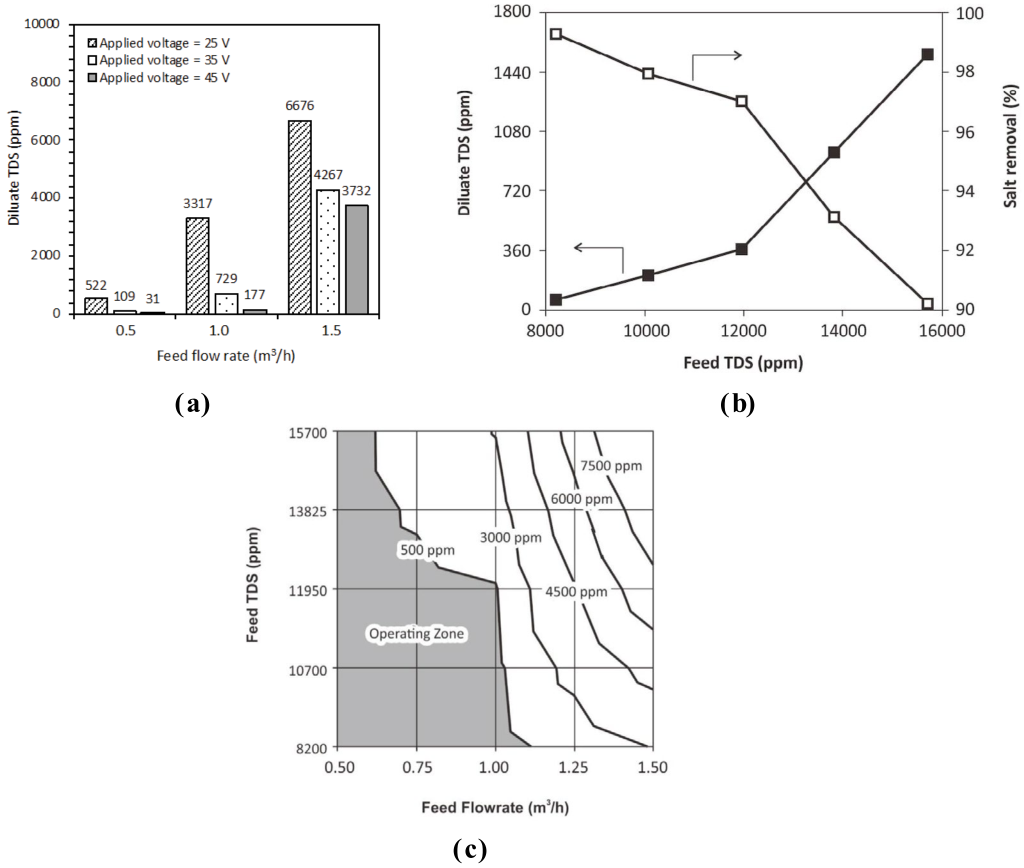



Fig. 5. EDI performance at various operating parameters and practical operating zone map for EDI. (a) Effect of flowrate and applied voltage on diluate TDS (feed TDS = 10,000 ppm), (b) effect of feed TDS on diluate TDS (flowrate = 1 $m^3/h$ and applied voltage = 45 V), and (c) map of a practical operating zone of EDI (the contour represents the value of diluate TDS).

effect of operating voltage, flowrate, and feed TDS is presented in Figs. 5a and b. The possibility of increasing plant capacity was investigated by varying feed flowrate under a constant applied voltage. Fig. 5a shows that the diluate TDS increases with increasing feed flow rate. This is obviously due to the increase in the total load of salt ions to be transported. The decrease in removal efficiency is also attributed to a shorter residence time at the higher feed flow [23–25]. Thus, operating above the normal capacity (1 $m^3/h$) is not recommended.

The TDS value of feed water was observed to be ranging from 8,200 ppm (rainy season) to 15,700 ppm (dry season) due to climate change. Fig. 5b reveals that with an increase in feed TDS at 15,700 ppm, the removal efficiency declined from 93.30% to 90.17% followed by increasing diluate TDS from 177 to 1,543 ppm which exceeds the drinking water standard recommended by World Health Organization (WHO) [28]. At the same hydrodynamic and electrical condition that is, flow rate and voltage, the number of ions transported through the ion exchange membranes is relatively unchanged. Thus, separation efficiency becomes inversely proportional to feed concentration. Generally, the ion-exchange membrane has high preferences under a low concentration of feed solution. At this high preference, the zone boundary in the EDI system exchanged sharply with high efficiency. At a high feed concentration, the exchange zone boundary of the EDI system becomes dispersed. The dispersed exchange zone induces an easy breakthrough of ions and then degrades water quality. Thus, a high feed concentration leads to steep performance degradation [21]. Also, permselectivity of the membrane is decreased at a higher salt concentration resulting in the lower separation performance [29]. Since operating at higher than 45 V gives an insignificant improvement in product quality, thus lowering feed flowrate is preferred. Finally, by analyzing the effect of operating conditions on EDI performance, the practical operating zone can be mapped (Fig. 4c). This map can be used as a preliminary guideline for efficient EDI operation in brackish water desalination.

### 3.2. Economic evaluation

CAPEX and OPEX of EDI and BWRO plants were summarized in Tables 4 and 5. It is shown that the initial CAPEX of the EDI system is lower than BWRO while energy cost is slightly higher. Elimination of the high-pressure pump in the EDI system significantly reduced investment cost up to 39% (Table 5). Moreover, since EDI was operated at low-pressure where a low-cost polymeric piping material such as PVC or HDPE can be used, pipeline and instrumentation cost is nearly 4 times lower than BWRO.

Table 4
Technological specification and design parameters of EDI and BWRO plant

| Parameters | EDI | BWRO |
|---|---|---|
| Total plant capacity ($m^3$/h) | 1 | 1 |
| Total recovery (%) | 75 | 50 |
| Hours of operation (h) | 12 | 12 |
| Annual volumetric water production ($m^3$/y) | 3,960 | 3,960 |
| Membrane lifetime (y) | 2 | 2 |
| Pretreatment | | |
| Number of UF module | 1 (non-modular, 8 in. (0.2 m)) | 1 (non-modular, 8 in (0.2 m)) |
| UF membrane material | Polysulfone | Polysulfone |
| Membrane area ($m^2$) | 120 | 120 |
| Recovery (%) | 80 | 80 |
| Operating pressure (MPa) | 0.15 | 0.15 |
| Main unit | | |
| Number of EDI stack | 1 | – |
| RO system | | |
| Number of RO pressure vessel | – | 5 |
| Number element per pressure vessel | – | 1 |
| System configuration | | |
| Array | – | 1 |
| Pass | – | 1 |
| Plant availability (%) | 90 | 90 |

In EDI, from OPEX estimation, it was found that electrical cost has the highest contribution (61%) followed by membranes replacement cost (37%) and maintenance (2%) (Table 5). According to this economic analysis, the specific water production cost of the EDI process was US$ 0.50/$m^3$ which is comparable to the BWRO plant (US$ 0.42/$m^3$).

The economic analysis from a long-term perspective is summarized in Table 7. Under the assumption that the costs for EDI and BWRO in Table 5 and Table 6 remain constant (for 20 y), it was found that EDI is more economical than BWRO up to around 10 y, since the initial investment CAPEX in year 0 for EDI is much lower than BWRO. However, with cheaper maintenance OPEX cost, BWRO becomes comparable with EDI from year 11 to 20.

As previously explained that one of EDI advantages over BWRO is lower piping costs associated with low-pressure operation. By assuming that the piping system must be replaced entirely every 10 y of operation which adds an extra $53 on EDI OPEX and $204.5 on BWRO OPEX. The EDI process is more economical than BWRO up to around 20 y (Table 7).

Table 5
Component of the capital cost of EDI and BWRO plant

| Component | EDI | BWRO |
|---|---|---|
| Pressure vessel | – | $568 |
| RO membrane | – | $1,136 |
| EDI stack | $1,464 | – |
| Power supply and electrical control panel | $1,566 | – |
| High pressure pump | – | $3,182 |
| RO feed pump | – | $379 |
| EDI feed pump | $379 | – |
| UF feed pump | $379 | $379 |
| CIP pump | $227 | $227 |
| Tanks | $530 | $530 |
| Interconnecting pipe + instruments | $530 | $2,045 |
| Total CAPEX | $5,076 | $8,447 |

Table 6
Summary of economic calculation

| No | Parameter | EDI | BWRO |
|---|---|---|---|
| I | CAPEX ($) | 5,076 | 8,447 |
| II | Production capacity ($m^3$/y) | 3,960 | 3,960 |
| III | Annual operating cost | | |
| | Total energy cost ($/y) | 1,211 | 948 |
| | Total membrane replacement cost ($/y) | 732 | 568 |
| | Total chemical cost ($/y) | 0 | 120 |
| | Service and maintenance ($/y) | 38.86 | 32.72 |
| | Total cost ($/y) | 1,981.86 | 1,668.72 |
| IV | Total specific water production cost ($/$m^3$) | 0.50 | 0.42 |

Table 7
Economic analysis from a long-term perspective

| Year | Type of exp. | EDI exp. ($) | BWRO exp. ($) | EDI cumulative exp. ($) | BWRO cumulative exp. ($) | Cost difference BWRO-EDI ($) | With additional piping replacement cost | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | EDI exp. ($) | BWRO exp. ($) | EDI cumulative exp. ($) | BWRO cumulative exp. ($) | Cost difference BWRO-EDI ($) |
| 0 | CAPEX | 5,076.0 | 8,447.0 | 5,076.0 | 8,447.0 | 3,371.0 | 5,076.0 | 8,447.0 | 5,076.0 | 8,447.0 | 3,371.0 |
| 1 | OPEX | 1,981.9 | 1,668.7 | 7,058.0 | 10,116.0 | 3,058.0 | 1,987.2 | 1,873.2 | 7,063.2 | 10,320.2 | 3,257.1 |
| 2 | OPEX | 1,981.9 | 1,668.7 | 9,040.0 | 11,784.0 | 2,745.0 | 2,034.9 | 1,873.2 | 9,098.0 | 12,193.4 | 3,095.4 |
| 3 | OPEX | 1,981.9 | 1,668.7 | 11,022.0 | 13,453.0 | 2,432.0 | 2,034.9 | 1,873.2 | 11,132.9 | 14,066.7 | 2,933.8 |
| 4 | OPEX | 1,981.9 | 1,668.7 | 13,003.0 | 15,122.0 | 2,118.0 | 2,034.9 | 1,873.2 | 13,167.7 | 15,939.9 | 2,772.1 |
| 5 | OPEX | 1,981.9 | 1,668.7 | 14,985.0 | 16,791.0 | 1,805.0 | 2,034.9 | 1,873.2 | 15,202.6 | 17,813.1 | 2,610.5 |
| 6 | OPEX | 1,981.9 | 1,668.7 | 16,967.0 | 18,459.0 | 1,492.0 | 2,034.9 | 1,873.2 | 17,237.5 | 19,686.3 | 2,448.9 |
| 7 | OPEX | 1,981.9 | 1,668.7 | 18,949.0 | 20,128.0 | 1,179.0 | 2,034.9 | 1,873.2 | 19,272.3 | 21,559.5 | 2,287.2 |
| 8 | OPEX | 1,981.9 | 1668.7 | 20,931.0 | 21,797.0 | 866.0 | 2,034.9 | 1,873.2 | 21,307.2 | 23,432.8 | 2,125.6 |
| 9 | OPEX | 1,981.9 | 1,668.7 | 22,913.0 | 23,465.0 | 553.0 | 2,034.9 | 1,873.2 | 23,342.0 | 25,306.0 | 1,963.9 |
| 10 | OPEX | 1,981.9 | 1,668.7 | 24,895.0 | 25,134.0 | 240.0 | 2,034.9 | 1,873.2 | 25,376.9 | 27,179.2 | 1,802.3 |
| 11 | OPEX | 1,981.9 | 1,668.7 | 26,876.0 | 26,803.0 | –74.0 | 2,034.9 | 1,873.2 | 27,411.8 | 29,052.4 | 1,640.7 |
| 12 | OPEX | 1,981.9 | 1,668.7 | 28,858.0 | 28,472.0 | –387.0 | 2,034.9 | 1,873.2 | 29,446.6 | 30,925.6 | 1,479.0 |
| 13 | OPEX | 1,981.9 | 1,668.7 | 30,840.0 | 30,140.0 | –700.0 | 2,034.9 | 1,873.2 | 31,481.5 | 32,798.9 | 1,317.4 |
| 14 | OPEX | 1,981.9 | 1,668.7 | 32,822.0 | 31,809.0 | –1,013.0 | 2,034.9 | 1,873.2 | 33,516.3 | 34,672.1 | 1,155.7 |
| 15 | OPEX | 1,981.9 | 1,668.7 | 34,804.0 | 33,478.0 | –1,326.0 | 2,034.9 | 1,873.2 | 35,551.2 | 36,545.3 | 994.1 |
| 16 | OPEX | 1,981.9 | 1,668.7 | 36,786.0 | 35,147.0 | –1,639.0 | 2,034.9 | 1,873.2 | 37,586.1 | 38,418.5 | 832.5 |
| 17 | OPEX | 1,981.9 | 1,668.7 | 38,768.0 | 36,815.0 | –1,952.0 | 2,034.9 | 1,873.2 | 39,620.9 | 40,291.7 | 670.8 |
| 18 | OPEX | 1,981.9 | 1,668.7 | 40,749.0 | 38,484.0 | –2,266.0 | 2,034.9 | 1,873.2 | 41,655.8 | 42,165.0 | 509.2 |
| 19 | OPEX | 1,981.9 | 1,668.7 | 42,731.0 | 40,153.0 | –2,579.0 | 2,034.9 | 1,873.2 | 43,690.6 | 44,038.2 | 347.5 |
| 20 | OPEX | 1,981.9 | 1,668.7 | 44,713.0 | 41,821.0 | –2,892.0 | 2,034.9 | 1,873.2 | 45,725.5 | 45,911.4 | 185.9 |

## 4. Conclusions

In this work, the performance of pilot-scale EDI in brackish water desalination was investigated and compared to the existing BWRO mini-plant. The results show that the EDI process has been successfully operated for brackish water desalination. At an optimum applied voltage of 45 V, EDI was able to remove more than 98% of ions. Even though the EDI process exhibits a higher SEC than the BWRO process, this application offers a less complicated low-pressure operation. EDI offers a lower capital investment (US$ 5076) than the existing BWRO plant (US$ 8447) and comparable water production cost (US$ 0.5/m$^3$ for EDI and US$ 0.42/m$^3$ for installed BWRO). EDI is also more economical than BWRO evidenced by the result of economic analysis from a long-term perspective (up to 20 y operation).

## Nomenclatures

| | | |
|---|---|---|
| $E$ | — | Applied voltage, V |
| $I$ | — | Electrical current, A |
| $Q$ | — | Diluate flow rate, m$^3$/s |
| $W$ | — | SEC of EDI, kWh/m$^3$ |

## Abbreviations

| | | |
|---|---|---|
| BWRO | — | Brackish water reverse osmosis |
| CAPEX | — | Capital expenditure |
| CIP | — | Cleaning-in-place |
| DC | — | Direct current |
| EDI | — | Electrodeionization |
| HDPE | — | High-density polyethylene |
| NF | — | Nanofiltration |
| OPEX | — | Operating expenses |
| PFD | — | Process flow diagram |
| PVC | — | Polyvinlyl chloride |
| RO | — | Reverse osmosis |
| SEC | — | Specific energy consumption |
| TDS | — | Total dissolved solids |
| UF | — | Ultrafiltration |
| WHO | — | World health organization |

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