Ultra-pure water production by integrated electrodialysis-ion exchange/electrodeionization

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Abstract. Ultra-pure water (UPW), a highly treated water free of colloidal material and of a conductivity less than 0.06 μS, is an essential component required by modern industry. One of the methods for UPW production is the electrodialysis-ion exchange (ED/IE) system, in which the electrodialysis (ED) process is used as a preliminary demineralization step. The IE step can be replaced with electrodeionization (EDI) to decrease the volume of post-regeneration lyes. In this paper, the electrodialysis process carried out to relatively low diluate conductivity was investigated and the costs of UPW production were calculated. The optimal value of desalination degree by ED in the ED/IE and ED/EDI systems was estimated. UPW unit costs for integrated ED/IE and ED/EDI systems were compared to simple ion exchange and other methods for UPW production (RO-IE, RO-EDI). The minimal UPW unit costs in ED/EDI integrated system were estimated as $0.37/m^3 for feed TDS 600 mg/L and $0.36/m^3 for feed TDS 400 mg/L at 64 m^3/h capacity, which was lower than in the comparable ED/IE integrated system ($0.42-0.44/m^3). The presented results suggest that an ED/EDI integrated system may be economically viable.

Keywords: electrodialysis; ion exchange; electrodeionization; ultra-pure water; cost analysis

1. Introduction

Ultra-pure water (UPW) is a highly treated water, free of colloidal particles, having an electric resistance of 18 MΩ/cm, total organic carbon content lower than 0.05 mg C/L and a reactive silica content lower than 2 μg SiO$_2$/L (Bennet 2006). Rapid growth of modern industry (i.e., pharmaceutical, semiconductor, electronic) causes an increase in demand for ultra-pure water. Traditionally utilized methods for obtaining ultra-pure water include distillation, ion exchange and integrated reverse osmosis/electrodeionization (RO-EDI) and reverse osmosis/ion exchange (RO-IE) systems. The ion exchange (IE) process allows the production of low conductivity water (Takeda *et al.* 2010, Yuan *et al.* 2000, Yu and Luo 2003), however has negative environmental impact, since the regeneration step generates high amounts of wastes. Electrodeionization is a separation process which combines electrodialysis and ion exchange. Ion exchange resins are placed between the ion selective membranes of the electrodialyzer. This allows in situ resin regeneration due to the creation of hydroxyl and hydrogen cations by the water splitting caused by

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electric overpotential (Tanaka 2007). Ion exchange resin also increases the conductance of resin-filled compartments. Since the electrodeionization process, contrary to ion exchange, does not require regeneration and thus does not create a need for post-regeneration lyes to be utilized, the EDI system may seem to be a more favorable choice over the IE system. RO-EDI systems can be more economical than RO-ion exchange, especially at higher RO permeate salinities (Tanaka 2007). RO-EDI systems are a common type of UPW production system for analytical laboratory purposes. Wood *et al.* (2010) have pointed out that the elimination of hazardous chemicals improves workplace health and safety conditions, and decreases the hidden costs of paperwork and environmental monitoring associated with the storage, use, neutralization, and disposal of hazardous chemicals. EDI systems have the advantage of continuous production of ultra-pure water, the quality of which can be easily regulated with applied current and feed volumetric flow (Trvznik *et al.* 2006), contrary to the IE system's product quality. Electrodeionization is also suitable for removal of species, which are mostly non-ionized in the pH range of typical feed waters – as are boron and silica. The silica removal rate for EDI is typically 98-99% (Hernon *et al.* 1999, Grebenyuk and Grebenyuk 2002, Wen *et al.* 2005). Boron, poorly ionizable and poorly removed by typical ion-exchange resins, can be removed down to the level of 0.05 ppb (Hernon *et al.* 1999). EDI is also suitable for removal of ammonia and carbon dioxide (Hernon *et al.* 1999, Grebenyuk and Grebenyuk 2002).

The reverse osmosis step is one of the methods which allows for a decrease in the load on electrodeionization or ion exchange. For instance, the "Grudziądz" Power Station (Bodzek and Konieczny 2005) has a reverse osmosis-ion exchange system, which treats water of TDS in the range of 480-576 mg/L giving product of TDS < 2.4 mg/L. "Żerań" Power Station (Klimanek and Koszarz 2001) uses a multi-stage reverse osmosis system, which treats feed of TDS 307 mg/L and produces permeate of TDS 6.6 mg/L. Liu *et al.* (2002) have demonstrated a RO-EDI system capable of producing 1 m³/h of water having resistivity of 17 MΩ/cm from a feed water having conductivity of 1800 μS/cm. A pretreatment consisting of ultrafiltration, activated carbon and ion-exchange softening, and removal of Ca^{2+} in order to prevent scaling was used. The total recovery of the system was 55%. Annual energy consumption of the EDI system was calculated at 560 kWh, compared to 2000 kWh in an ion-exchange design of similar capacity. Based on the presented data, we have estimated the RO permeate conductivity at 18 μS/cm. RO-EDI systems have been succesfully implemented in Krasnodar (Panteleev *et al.* 2012) and Putilovo (Gromov *et al.* 2011) power plants, with water production capacity of 33 and 40 m³/h, respectively. Wang *et al.* (2000) have described a RO-EDI system capable of producing 20 L/h water of conductivity varying from 12 to 18 MΩ/cm (15 MΩ/cm on average). The RO permeate conductivity was less than 20 μS/cm. An RO-EDI system for ultrapure water for boron trace analysis and for general UPW production was described by Darbouret and Kano (Darbouret 2000, Kano 2004). Working on feed water having conductivity of 569 μS/cm, the RO-EDI system was able to produce 10 L/h of 5 MΩ/cm water with a RO permeate conductivity of 12.3 μS/cm. The EDI diluate was further purified with ion exchange, allowing to reach the ultra-pure water resistivity of 18.2 MΩ/cm. An RO-EDI system was also described by Su *et al.* (2010). The RO step was put in place to lower the feed water hardness down to 11.7 mg/L as $CaCO_3$, thus allowing the EDI to produce water having a resistivity of 14.6 MΩ/cm, with an energy consumption of 0.26 kWh/m³. The industrial-scale electrodeionization solutions are being sold by various producers, with capacities reaching 150,000 GPD (Winda 2013).

Since reverse osmosis membranes are more sensitive to fouling than ion-exchange membranes, electrodialysis performed down to relatively low diluate salinity can be used as a pretreatment step

before ion exchange (Slesarenko 2003, 2005). ED has a much lower environmental impact. However, the costs of ultra-pure water production are relatively high, because the process becomes inefficient at low salinity due to high electric resistivity. In the case of using ED, it is possible to substantially reduce the consumption of expensive reagents for regeneration of the ion exchange bed and to increase the cycle time. Even when using the simplest one or two - stage water treatment by electrodialysis before its delivery to ion exchange, the consumption of acid and alkali for regeneration of the ion exchange bed is substantially reduced. A specific power consumption of equal to 0.2-0.3, 0.3-0.7 and 0.7-1.0 kWh per kg of removed salt was found for river, brackish and sea water, respectively (Slesarenko 2003). Based on the presented data, we have calculated that feed water was desalinated by ED to ca. 45-60 mg/L. Electrodialysis reversal (EDR) was also applied as a preliminary desalination step before IE. EDR is less sensitive for fouling and scaling than classical ED, so only a simple pretreatment is needed, and operation at a very high water recovery ratio is possible (85-95%). Final treated water salinity level was even higher than found in the aforementioned ED, e.g., 111.5 mg/L (Kurowski 1994). Chin (1996) has compared RO and EDR as a demineralization step before ion exchange. The EDR diluate conductivity was 180 μS/cm with 85% recovery, while RO permeate conductivity was 50 μS/cm with 60-70% recovery. The author reported water splitting and precipitation of metal hydroxides in the EDR step, which suggest too high current was applied.

The problem with electrodialysis, though, is that high electric resistance of diluted solutions dramatically increases the process costs when a high degree of desalination is to be reached. Previous research (Turek *et al.* 2007, Bandura-Zalska *et al.* 2009) showed that thanks to thin intermembrane spacers during electrodialytic boron removal, the treated water was at the same time deionized to approximately distilled water level. We believe that our solution, ED with thin spacers, will be thus a better first step of ultrapure water production, because of the lower salinity of ED diluate, resulting in the lower salinity of the EDI or IE feed. In this paper, the results of electrodialysis experiments and data concerning the electrodeionization process given by Fedorenko (2003a, 2003b, 2004) were used for optimization of an ED-EDI system and estimation of UPW production cost. Thus, we decided to evaluate the ED process carried out to a relatively low diluate conductivity as a first step of ultrapure water production. We would like, however, to focus only on obtaining water with an adequate high electric resistance of 18 MΩ/cm. The question of other impurities would not be examined, as we expect in case of modern water treatment plants ultrafiltration pretreatment would be used, which should decrease the problems with fouling caused by bacteria, colloidal silica and other colloids (Cheng *et al.* 2009). Achieving very low ED diluate conductivity means that most of the bicarbonate and ammonium ions has been removed. Results are then compared to an ED-IE design, based on the data provided by the DOW company (2011) and Venkatesan and Wankat (2011).

2. Experimental

The experiments were performed in batch circulation mode, schematically presented in Fig. 1. An ED stack was equipped with AMX and CMX Neosepta membranes arranged in four unit cells. A thin 0.40 mm spacer was used to decrease the inter-membrane distance, as well as the diluate cell resistance.

Operating parameters were chosen based on the previous research (Turek 2007): linear flow velocity was set to 2 cm/s, voltage drop per unit cell was set to 1.2 V. Results of the experiments
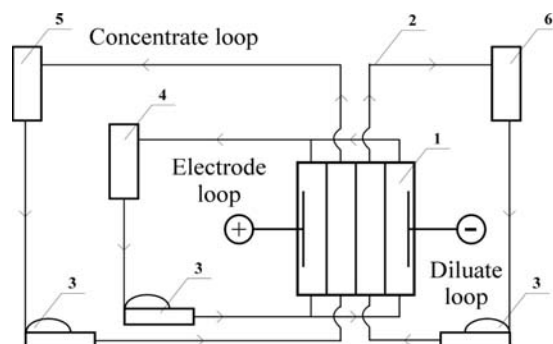
Fig. 1 ED experimental setup. 1 – ED unit, 2 – conductometric detector,   3 – peristaltic pumps,
4 – electrode solution container, 5 – concentrate container, 6 – diluate container

are presented in Tables 1-4. It was possible to reach a very low, at least for ED, value of TDS 1.15 mg/L.

Results were used for estimation of unit cost calculation of industrial-scale ED plants. The estimation assumptions were: a membrane cost of $100/m^2, effective membrane area of 70%, membrane durability period of 35000 h, energy costs of $0.06/kWh of AC current, AC/DC rectifier electric efficiency of 95%, a pumping efficiency of 85%. The contribution of membranes in the total ED cost was assumed to be 30%.

Table 1 Results for series 1 (Initial TDS 400)

Time [min]	Diluate conductivity [μS]	TDS [mg/L]	Current density [A/m^2]
0	833	400	22
10	537	258	11
17	295	142	3.3
26	8.3	3.99	2.0
35	3.1	1.49	1.3
45	2.4	1.15	1.1

Table 2 Results for series 2 (Initial TDS 400)

Time [min]	Diluate conductivity [μS]	TDS [mg/L]	Current density [A/m^2]
0	833	400	31
5	694	333	20
12	451	217	9.3
21	8	3.84	3.8
30	3.5	1.68	2.0
40	3.3	1.58	1.3
51	2.6	1.25	1.1

Table 3 Results for series 1 (Initial TDS 400)

Time [min]	Diluate conductivity [μS]	TDS [mg/L]	Current density [A/m^2]
0	833	400	22.2
10	538	258	11.1
21	47	22.6	5.1
31	11	5.28	2.7
41	4.2	2.02	1.6
51	3.1	1.49	1.3
61	2.6	1.25	1.1

Table 4 Results for series 4 (Initial TDS 600)

Time [min]	Diluate conductivity [μS]	TDS [mg/L]	Current density [A/m^2]
0	1250	600	44.4
3	763	366	31.1
13	104	49.9	7.6
23	12	5.76	2.4
33	2.4	1.15	1.1

3. Electrodialysis costs estimation

The energy cost was calculated using the equation

$$C_E = C_{current}\left(E_{AC} + \frac{E_{DC}}{\eta_{AC/DC}} \right) \tag{1}$$

where $C_{current}$ denotes unit cost of AC electric energy assumed as \$0.06/kWh, E denotes energy consumption and $\eta_{AC/DC}$ denotes electric rectifier efficiency. The ED energy consumption per 1 m^3 of purified water during the time t of the experiment was calculated using the equation

$$E_{DC} = \frac{UA}{\eta_{AC/DC}V_{dil}} \int_0^t i\,dt \tag{2}$$

where E_{DC} denotes required DC energy consumption in kWh per 1 m^3 of purified water, U denotes applied voltage across the membrane stack, V_{dil} denotes diluate volume and the i denotes the applied electric current density, A – membrane area (given as active area divided by the percent of membrane effective area). Next, the pumping energy consumption per 1 m^3 of purified water was calculated using the equation

$$E_{AC} = \frac{Q_{ED}\Delta P_{ED}t}{\eta_p V_{dil}} \tag{3}$$

where Q_{ED} denotes volumetric flow through the electrodialyzer, η_p denotes pump efficiency and ΔP_{ED} denotes pressure drop within the ED stack, which was calculated based on the empirical correlation

$$\Delta p = 0.0862ul \tag{4}$$

where pressure drop is given in kPa, u denotes linear flow velocity through the diluate and concentrate compartments in cm/s, l denotes effective membrane length in cm.

Capital and maintenance cost were calculated with equation

$$C_{cap} = \frac{nC_m At}{0.3V_{dil}t_{life}} \tag{5}$$

where n denotes the number of membranes, C_m – cost of 1 m^2 of membrane and t_{life} denotes membrane durability period.
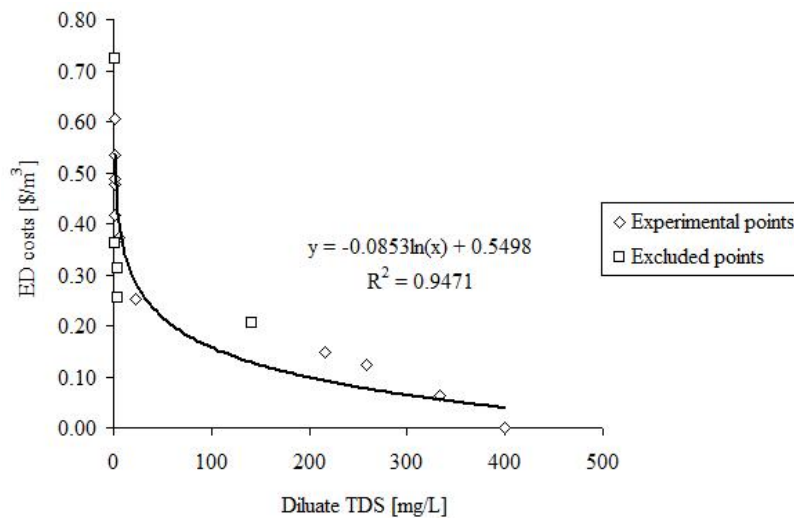

Fig. 2 Dependence of the ED demineralization costs on diluate TDS for series 1-3 with initial TDS 400
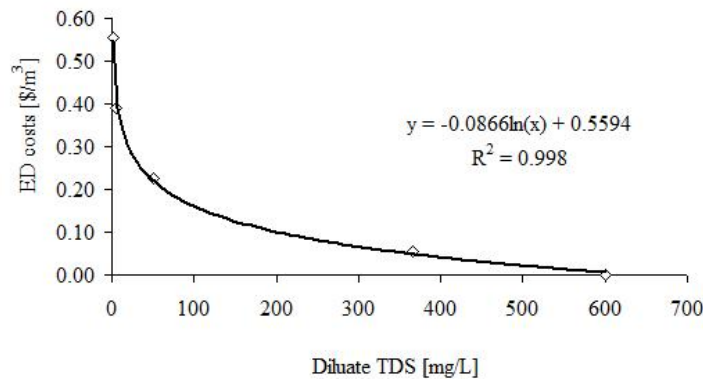

Fig. 3 Dependence of the ED demineralization costs on diluate TDS for series 4 (initial TDS 600)

Total cost of electrodialysis was calculated as a sum of energy, capital and maintenance costs. Figs. 2 and 3 show the calculated dependence of total ED demineralization costs on diluate TDS. The costs of electrodialysis rapidly increase when the final diluate TDS decreases, which is expected behavior for an ED process progressed down to low diluate conductivity.

4. ED/EDI cost estimation

The following EDI design was proposed: small scale design, four Yongjieda 4T/H EDI modules (Hangzhou 2010), each producing 4 m^3/h, totalling Q_{UPW} = 16 m^3/h, large scale design, 16 modules totaling Q_{UPW} = 64 m^3/h. Working pressure was proposed as ΔP_{EDI} = 3 bar, pump efficiency η_{EDI} = 85%, conversion R = 85%, maximum feed salinity 25 mg/L – a typical EDI feed water requirement (Wood *et al*. 2010), product resistance 18 MΩ/cm, voltage drop U_{EDI} = 50 V per each module, apparatus cost \$1500 per each module, apparatus life-time of 3 years with 350 working days per year, being 20% of total maintenance and investment costs. Pump requirements for concentrate and electrolytic loops were: volumetric flow 0.3 Q_{UPW}, pressure drop 3 bar. The formula for calculating operating current was recalculated with the assumption that total exchangeable anions as ppm CaCO$_3$ equals TDS in mg/L, taking into account all four modules and multiplying the result by 1.25 safety factor, resulting in

$$I = 0.6 \cdot TDS + 4.6 \qquad (6)$$

The DC consumption per m3 of ultra-pure water produced was thus calculated using the equation

$$E_{EDI} = \frac{4U_{EDI}I_{EDI}}{Q_{UPW}} \qquad (7)$$

The required feed volumetric flow with assumed 85% conversion was (Fedorenko 2003a)

$$Q_{in} = \frac{Q_{UPW}}{1 - \left(\frac{100}{85} - 1\right)} \qquad (8)$$

The pump power requirement was thus

$$P_p = \frac{(Q_{in} + 0.3Q_{UPW})\Delta P_{EDI}}{\eta_{EDI}} \qquad (9)$$

The total cost of the ED/EDI process was calculated using equation

$$C_t = 0.06\left(E_{EDI} + \frac{P_p}{Q_{UPW}}\right) + \frac{4 \cdot \$1500}{0.2 \cdot 3y \cdot 350d \cdot 24h \cdot Q_{UPW}} + C_{ED}\frac{Q_{in}}{Q_{UPW}} \qquad (10)$$

where C_{ED} denotes electrodialysis costs as a function of ED/EDI feed TDS, as shown in Figs. 2 and 3. Fig. 4 shows the calculated dependence of UPW unit cost on ED diluate TDS in an integrated ED/EDI system. The minimum lies at TDS higher than 25 mg/L, which was assumed to
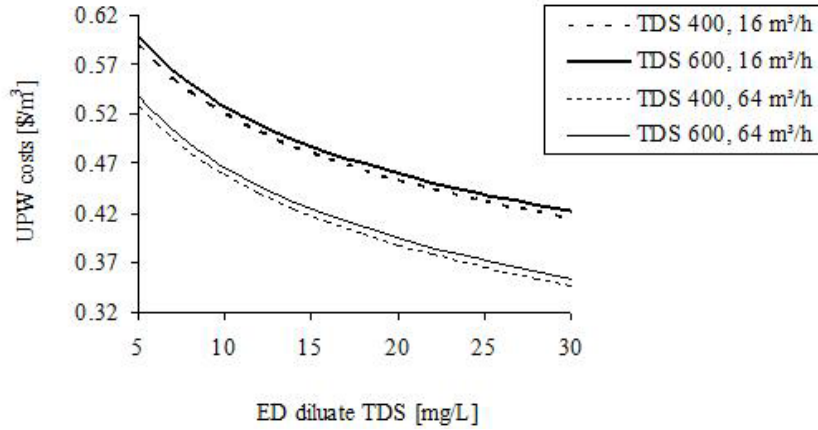
Fig. 4 The dependence of UPW production costs on ED diluate TDS, results for feed TDS 400
and 600 mg/L, production capacity 16 m^3/h and 64 m^3/h

be maximal EDI feed salinity, so the proposed ED diluate TDS is 25 mg/L, which would result in UPW production cost of \$0.43/m^3 for feed TDS 400 mg/L and \$0.44/m^3 for feed TDS 600 mg/L in case of small scale design and respectively \$0.36/m^3 and \$0.37/m^3 for large scale design.

5. ED/IE cost estimation

The IE costs were estimated based on theoretical design. Assuming that ED diluate is mostly sodium chloride and the UPW concentration should be close to zero, the required amount of ion-exchange resin was estimated as

$$m_{IE}^{+/-} = 1.2 \frac{TDS}{M_{NaCl}} \frac{Q_{IE} t}{c_{IE}^{+/-}} \qquad (11)$$

where $m_{IE}^{+/-}$ is the required mass of resin (cationic or anionic) [g], *TDS* is the ED diluate salinity (mg/L), Q_{IE} is the flow through the column [m^3/h], M_{NaCl} is the equivalent mass of sodium chloride [g/eq], $c_{IE}^{+/-}$ is the cationic or anionic resin ion-exchange capacity [eq/g], t denotes the service run time [h], which was assumed as 2 h. Assuming strong acid and strong base Dowex® resins – respectively 2 eq/L with 1.28 g/ml particle density and 1.3 eq/L with 1.08 g/ml (DOW 2011), the ion exchange capacities were calculated as c_{IE}^+ = 1.56 meq/g and c_{IE}^- = 1.2 meq/g. A 20% excess was assumed. Both regeneration time t_{reg} and conditioning time t_f were assumed as equal to 2 bed space-times

$$t_{reg} + t_f = 4 \frac{V_{IE}}{Q_{IE}} \qquad (12)$$

To compare ED/IE with an ED/EDI system, as described in the previous section, the amount of produced UPW during the complete cycle (demineralization-regeneration-flush) was assumed equal in both systems

$$Q_{IE}t = Q_{UPW}\left(t + t_{reg} + t_f\right) \tag{13}$$

Assuming column void fraction as 0.5, the required bed volume was estimated as

$$V_{IE} = 2\left(\frac{m_{IE}^+}{\rho^+} + \frac{m_{IE}^-}{\rho^-}\right) \tag{14}$$

where ρ denotes the resin density (cationic/anionic). Taking into account Eqs. (11), (12), (13) and (14), the final equation for the volumetric flow through the bed can be stated as

$$Q_{IE} = Q_{UPW}\left(1 + 4 \cdot 1.2\frac{TDS}{M_{NaCl}}\left(\frac{1}{c_{IE}^+\rho^+} + \frac{1}{c_{IE}^-\rho^-}\right)\right) \tag{15}$$

It was assumed that the ratio of bed depth to column diameter should be equal to 6, so the bed length was calculated as

$$L = \sqrt[3]{\frac{144V_{IE}}{\pi}} \tag{16}$$

The pressure drop on the IE column was then calculated based on a generalization of the example data on pressure drop per bed depth (DOW 2011), reformulated as

$$\Delta P_{IE} = \frac{5.4Q_{IE}}{\pi L} \tag{17}$$

The pumping cost per 1 m^3 of UPW was then calculated as

$$C_p = C_{DC}\frac{Q_{IE}\left(t + t_{reg} + t_f\right)}{Q_{IE}t}\Delta P_{IE} \tag{18}$$

The chemical cost was calculated assuming a 50% excess of acid and base, \$200 per tone of technical grade concentrated hydrochloric acid (ca. \$0.019 per 1 mole of HCl) and \$300 per tone of solid sodium hydroxide (ca. \$0.012 per 1 mole of NaOH). The costs of dissolving regeneration agents down to the level required by the resin was neglected. The total cost of chemicals per 1 m^3 of UPW was calculated as

$$C_{chem} = 1.5\frac{0.017\,m_{IE}^+c_{IE}^+ + 0.012\,m_{IE}^-c_{IE}^-}{Q_{IE}t} \tag{19}$$

The equipment cost was estimated based on the equation given by Venkatesan and Wankat (Venkatesan and Wankat 2011)

$$C_{eq} = \frac{(MS)}{280}101.9\left(\frac{1}{6}L\right)^{1.066}L^{0.802}\cdot 2\cdot\frac{1}{\frac{350}{365}\cdot 25\cdot Q_{IE}t} \tag{20}$$
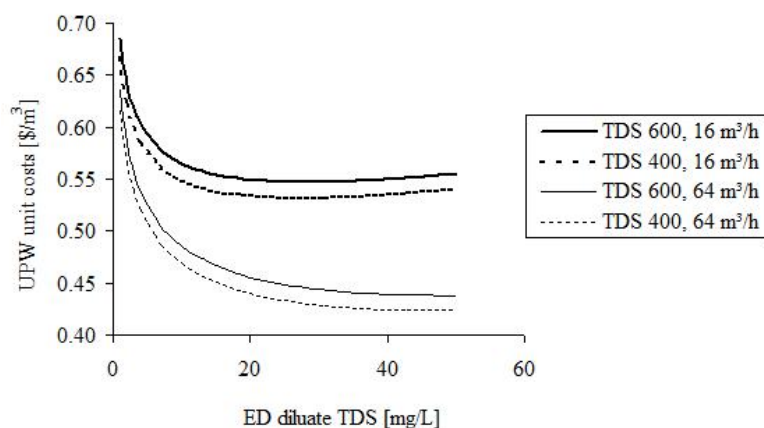
Fig. 5 The total costs of UPW production in integrated ED-IE system

where *MS* denotes the assumed M&S index (1468.6) and the 25 years of plant life, 350 days per year. Fig. 5 presents the UPW unit costs, calculated as a function of ED diluate TDS.

6. Results discussion

The presented results indicate that an integrated ED/EDI system may be of potential use for UPW production. Cocker *et al.* (1994) gives the UPW production cost for a reverse osmosis/ion exchange system as \$0.44-0.49/m^3, \$0.40-0.44/m^3 and \$0.39/m^3 for ca. 1 000, 2 000 and 4 000 m^3/d plants respectively, with a feed TDS range of 100-600 mg/L and a product resistance higher than 10 MΩ/cm. For comparison, the estimated results of the authors' ED/EDI integrated system producing 1024 m^3/d of UPW (large scale design, 64 m^3/h) of resistance 18 MΩ/cm are \$0.36-0.37/m^3 for feed TDS range 400-600 mg/L, which suggests that an ED/EDI system is a better solution for UPW production, especially given that a bigger-scale ED/EDI plant should result in lower unit costs. Goffin and Calay (2000) have compared the energy consumption of EDI and IE for the removal of ammonia in steam generators blow-down. They have found that the investment costs of EDI installation was 7 MBEF (ca. \$230 000) and the energy consumption was 0.7 kWh/m^3 at capacity ca. 75 l/h, while in case of IE the operating costs, without the effluents, were given as 850 kBEF/y (ca. \$28 000/y). Hu *et al.* (Hu 2008) have measured the energy consumption of an installation producing 102.8 m^3/h of UPW – 9.15 kWh/m^3 if the UPW transporting energy was to be neglected. Assuming \$0.06/kWh, that would generate the costs \$0.55/m^3. That is much higher than the values calculated for ED-EDI system, however our calculations do not include all of the factors that contribute to the energy consumption. Unfortunately, Hu *et al.* focused on measuring energy consumption of the various systems in a semiconductor plant and did not discuss what method was used for UPW production.

Liu and Martin (2006) have investigated a possible application of membrane distillation (MD) in ultra-pure water production, testing a two-stage MD installation producing 36.6 m^3/h of UPW. The energy consumption was 438 kWh/m^3 (thermal) and 1.9 kWh/m^3 (electrical). The total UPW cost was calculated as \$1.13/m^3. Pandya (1992) describes an ion exchange system treating feed of TDS 216.5 mg/L, giving 22 m^3/h of a product of electric resistance higher than 5 MΩ/cm, for

which the production costs were recalculated as $0.39/m^3 for an existing installation and $0.27/m^3 for a projected installation. This is less than the authors' presented design, however an ED/EDI system would give a product of higher resistance (18 MΩ/cm), with smaller environmental impact due to ion exchange resin regeneration. The costs of ED demineralization significantly increase when desired diluate TDS is in the low salinity range. Combination of electrodialysis and electrodeionization processes allows production of ultra-pure water at relatively low costs. The optimal values of TDS in an ED diluate stream were estimated as 25 mg/L, giving $0.44/m^3 for feed TDS 600 and $0.43/m^3 for feed TDS 400 when the 16 m^3/h capacity was assumed and respectively $0.36/m^3 and $0.37/m^3 when 64 m^3/h capacity was assumed. The estimated costs, compared to other methods for UPW production, suggest that using ED as a pretreatment step for EDI might be economically justifiable; however, the process needs further investigation. The ED/IE design exhibits higher UPW production costs than an ED/EDI design of comparable capacity. When a 16 m^3/h capacity was assumed, the minimal costs were $0.55/m^3 for an ED diluate TDS of 30 mg/L and feed TDS 600 mg/L, $0.53/m^3 for an ED diluate TDS of 27 mg/L and feed TDS 400 mg/L. For a 64 m^3/h capacity, the minimal costs were $0.44/m^3 for an ED diluate TDS 49 mg/L and feed TDS 600 mg/L and $0.42/m^3 for an ED diluate TDS 47 mg/L and feed TDS 400 mg/L. This may be even higher, given that the post-regeneration lyes disposal was not taken into account, but they are generally 17-25% higher than in ED-EDI system. Wenten *et al.* (Wenten 2013) have investigated the integrated RO-EDI system for the production of high pressure boiler feed water. The calculated production costs of water having conductivity 0.3-0.4 μS/cm was $0.41/m^3 at capacity 120 m^3/h ($0.53/m^3 if depreciation of equipment was taken into account), which was lower than existing ion exchange installation ($0.66/m^3). The presented costs are lower than the UPW production costs in ED-EDI system, however Wenten *et al.* (2013) assumed the electricity costs to be $0.038/kWh, which seems quite low. Also, the capacity was twice as much as the capacity assumed for ED-EDI system.

The presented results are based on the bench-scale experiments with model solutions and may not accurately describe the demineralization of real-life water, which can contain colloids, biological impurities and other contaminants that reduce the membrane life-time and increase the energy consumption. On the other hand, novel ED and EDI solutions can improve the performance of the ED-EDI system. A thinner, 0.26 mm intermembrane spacers with low residence time variance and have shown higher conversion than the 0.40 mm spacers without the drastic increase in the pressure drop on the module (Turek *et al*. 2013), which could result in smaller membrane area required. EDI performance can be improved with an ion-exchange textiles, which can result in higher current efficiency and faster ion-exchange (Yeon *et al*. 2004).

7. Conclusions

The costs of ED demineralization significantly increase when desired diluate TDS is in a low salinity range. Combination of electrodialysis and ion exchange or electrodeionization processes allows production of ultra-pure water at relatively low costs. The optimal values of TDS in an ED diluate stream were estimated as 25 mg/L, giving a UPW production unit cost of $0.37/m^3 for initial TDS 600 and $0.36/m^3 for initial TDS 400 if a 64 m^3/h capacity was assumed. These costs are lower or comparable to the costs of ultra-pure water production by IE alone and by other methods, including RO/IE and ED/IE, which indicates that performing ED down to relatively low diluate salinity as a pretreatment step of electrodeionization may be economically justified.

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