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Design of ILEDR for brackish groundwater: A literature review approach

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ABSTRACT

A literature review was done in details in desalination by using electrodialysis reversal (EDR). All available data – source of water entered into pretreatment and/or directly into EDR, physico-chemical characteristics of water, targeted pollutants, methods and the reasons for the pretreatments, specifications used of EDRs, developments/improvements in EDR, fouling causing compounds and methods to control fouling, membrane types, membrane assembly, spacer, and concentrate management, method, and recovering of ions were searched and summarized in Tables 1–3. From the tables, the concentrations profiles of the targeted pollutants were observed in different sub-processes (i.e., pretreatment, ERD dilute and concentrate streams). The percentage removals of individual ion from different literatures were compared. The concentrations of fouling causing compounds (for examples, $CaSO_4$ and $MgSO_4$) in concentrated streams were eliminated by individual separation process with the specific different types of permselective membranes in four different stages. All these summarized data were used as tools, references, and comparisons to design and select the sub-processes in ILEDR project to treat the brackish groundwater into the drinking water. Design includes designs I (using a single type of membrane pair) and II (using four different types of mono- and di-valent permselective membrane). Permselective membrane design shows 7% cost saving while comparing to the single type of membrane design with the same water recovery rate, the same demineralization rate, and the same membrane life. The saving increases up to 15–18% if membrane life is considered in the design.

Keywords: $CaSO_4$ and $MgSO_4$ scaling eliminating by design; Desalination; Inland electro-dialysis reversal; Membrane fouling; Permselective membrane

1. Introduction

Electrodialysis reversal (EDR) is a modified vision of electrodialysis (ED) to reduce the membrane fouling and scaling by reversing the electric polarity in a fixed time interval. To avoid fouling and scaling, the polarity of the electrodes is reversed in every 15 min, and the flows are simultaneously switched so that the concentrate channel becomes the dilute channel, and vice versa [1]. By revering the electrical driving force pushes the charged particles off the membrane surface back into the waste stream [2]. During the polarity reversing, the reactions in electrodes are chemically reversed [3]. Hydrogen and hydroxide ions are produced in negative electrode in Fig. 1, and hydroxide raises the pH of water, resulting calcium carbonate $CaCO_3$ precipitation [3]. Acid, oxygen, and some chlorine are produced from positive electrodes in Fig. 1, the acid has ability to melt any calcium carbonate in attendance to hamper scaling [3]. Thus the EDR process is

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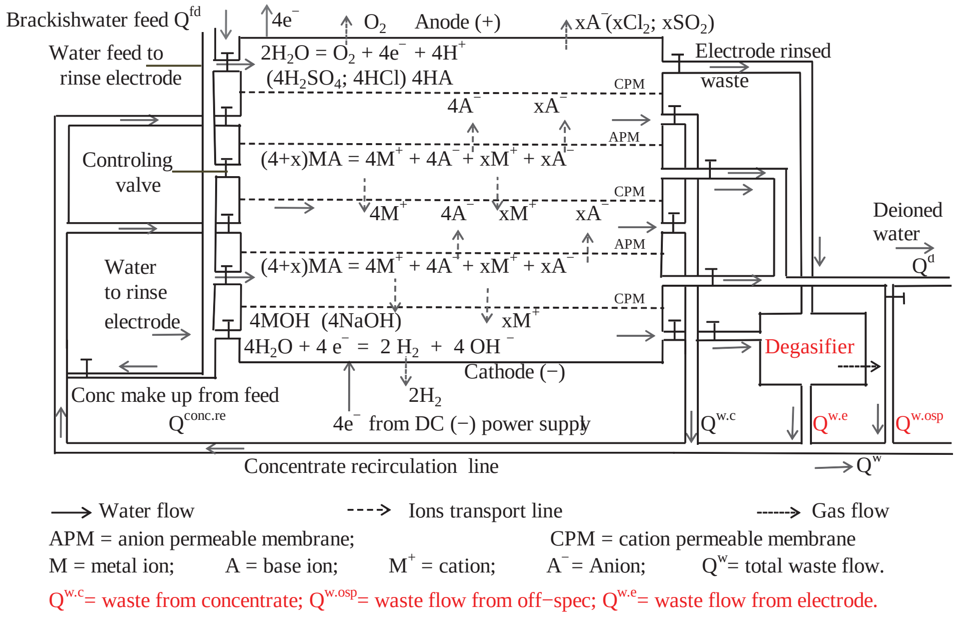


Fig. 1. Chemical reactions in standard electro-dialysis reversal array of alternating cation and anion membrane separated by alternating product and brine compartment, Referred from [17].

self-cleaning by reversal the chemical reactions, electricity driving force, and the flows. The *effectiveness* of current-reversal process of EDR are shattering and detaching polarization films four times/hour to prevent polarization scale; contravention up freshly precipitated scale or seeds of scale and flushing them back into waste stream before they can cause scratch to membrane; reducing slime formations on membrane surface; reducing problems associated with the use of chemicals; cleaning electrodes with acid automatically during alternative anodic operation [4] in reversal. There is a period (1–2 min [4]) instantaneously following the polarity reversal, the waters from both dilute and concentrate are required to send back to the feed stream or diverted to waste because these waters are *''off specification''*.

EDR *efficiency* depends on method to prevent membrane from *fouling*; ion's charge and mobility, solution conductivity, relative concentrations, applied voltage, and the characteristics of the ion-exchange membranes, especially its permselectivity [5]; system design [6,1]; materials used in membrane [7].

EDR is not cost effective at plant capacity lower the 40 m^3/h while comparing with R.O; Elyanow and Persechino recommended that high recovery (94%), elevated SDI (6–12), potential for biofouling, hard-to-treat, high hardness, lower salinity waters (from 200 to 5,000 mg/L TDS) are preferential to EDR technology, while lower recovery (60–85%), low SDI (4–5), high TDS waters (TDS > 5,000 mg/L) and low plant capacity (<40 m^3/h) are preferential towards RO [8]. Chemical and acids additions can effectively control $CaCO_3$ scaling but less effectively $CaSO_4$ scaling [9]. However, due to its polarization reversal frequency to clean up the scaling especially from calcium sulfate, EDR can operate without any chemical addition with *calcium sulfate* levels greater than *saturation*; the upper design limit for calcium sulfate level is 150% [3] or 175% [10] $CaSO_4$ saturation in which there is no chemical addition requirement for cleaning [3,9]. EDR is branded or gradually known for its excellence to desalt Ca^{2+} and SO_4^{2-} dominated brackish groundwater.

2. Objectives of this study

The preliminary objective of this study is to summarize all the technical available data including inflow characteristics, with or without pretreatment requirement, membrane types and properties, fouling prevention and cleaning, inflow velocity, spacer design, water recovery, TDS or target pollutant treatment efficiency, concentration management technologies, metal recovery and its marketable quality in EDR from literature. With the summarized information, the final *objective* is to design an inland EDR (ILEDR) desalination project which will be usable in Alamogordo, New Mexico, the United States of America.

2.1. Literature review in pretreatment

EDR is a direct current driven membrane separation process in which the charged species are exchanged throughout from a semi-ion-membranes while the de-charged diluted water passes between-in the membranes [2]. A positive charged cation from wastewater is pushed throughout from semi-permeable cation-membrane and migrate toward a semi-permeable-anion-membrane by electrostatic attraction; a negative charged anion is pushed throughout from semi-permeable-anion-membrane and migrate toward semi-permeable-cation-membrane [8]. The de-charged clean water is collected in-between the channel within the considered membranes, and concentrated ions wastewater is collected from channel outside the considered membrane in Fig. 1. *Non-charged particles* such as silica (SiO_2) [11], silicon (Si), bacteria, and organic contaminants are not removed by EDR [12,5,3]. Boron is also not removed by EDR [13]. These particles may cause fouling in membrane which may reduce the efficiency of EDR performance; these particles may be removed in *pretreatment* from feed water before desalting in the EDR stack. The requirement of pretreatment is based on characteristics of feed water. Certain levels of concentrations of $CaCO_3$ and $CaSO_4$ in concentrate stream may cause fouling in membrane; iron, manganese, sulfides [14,11], and CO_2 [15–17,11] may cause fouling in anode electrodes by oxide the ferrous and manganous species [18]. Chemical additions can effectively control the $CaCO_3$ scaling but less effectively control in $CaSO_4$ scaling [9].

2.2. Literature review in case studies in EDR without pretreatment

Tables 1 and 2 summarized eight case studies [19–24,14,8] including real field scale without pretreatment including Washington, Iowa, groundwater EDR desalination treatment by [19]. From Table 1, in case of [19], all pollutants except *pH and silica,* are removed in EDR without any kind of pretreatment, and all the pollutants are met with the primary drinking water standard after EDR without the pretreatment. After the EDR units, the water flows through an aerator where odor-causing gases and carbon dioxide are removed. The removal of carbon dioxide (CO_2) and carbonate (CO_3^{2-}) in aerator raises the finished *water pH* and causes the water to be stable or slightly alkaline in Table 1, resulting in some deposition of scale in the water mains not in EDR [19]. This indicated that the changes in pH is closely relative to the changes in carbon dioxide associated compounds in water, and the removal and/or deposits of carbon dioxide associated compounds is affected and shown in pH changes. The pH in the concentrated stream is used as an indicator to monitor and avoid scaling problems [5]. The rise, in silica-SiO_2 [12] and silicon-Si [19] in dilution stream after the EDR treatment in Table 1, is due to the mass balance of silica in EDR feed and EDR dilute stream with water amount loss in dilution into concentrated stream.

All EDR units require multi-media sand filter and 10 μm cartridge filter as minimum pretreatment to prevent total solid entering into EDR. Table 3 summarizes data from pretreatment with EDR performance from literatures [25 and 26]. Iron (1.5 mg/L) and manganese (0.41 mg/L) were necessary to pretreat from feedwater with 2.5 to 2.8 mg/L 2% solution of potassium permanganate injection to oxidize iron and manganese followed by permanganate greensand filtration to the levels of 0.008 and 0.03 mg/L respectively before desalting in EDR [25]. Table 4 summarizes design criteria and performance of full scale EDR in Prairie, Canada from groundwater characteristics and pretreatment to EDR from [18]. Because the groundwater does not contain TS, and there is no pretreatment requirement for TS in this project, however, iron (1 mg/L) and manganese (0.5 mg/L) were required to pre-treat to 0.3 and 0.2 mg/L, respectively, due to their potential stack and electrode fouling. The fouling starts from the accumulation of Cl_2 and O_2 from the anode electrodes, and these Cl_2 and O_2 oxide the ferrous and manganous species following by precipitation. Silicon (Si) 12 mg/L and Silica (SiO_2) 9.5 mg/L in feed groundwater were not required pretreatment [19 and 25]; however, SiO_2 30 mg/L in feedwater required pretreatment [18]. The pretreatment requirement for some ions with the tolerance limits are summarized in Table 5.

2.3. Literature review in membrane

EDR has been *improved* in design, membrane quality. Membrane has been improved from crushed ion-exchange resins in an inert matrix to acrylic-based anion exchange membranes to increase the resistance toward organic fouling and chloride tolerance [27]. Acrylic-based membranes also reduce the tendency of crystallization centered on styrene-divinyl-benzene membrane [7]. Membranes are flat with a smooth surface and reinforced with synthetic fiber. Both anion and cation transfer membranes has low electrical resistance; insolubility in aqueous solutions; semi-rigid construction for ease of handling; resistance to change in pH from 2 to 9; ability to function continuously at ambient temperatures above 115°C; resistance to osmotic swelling (volume expansion due to water uptake) in solutions between 220 and 60,000 mg/L salt; long life expectancy (average 10 years for anion and 15 years for cation membranes); resistance to fouling; impermeability to water under pressure; available in many sizes and configurations. Membrane typically

Table 1
Performance inside EDR without pretreatment or after pretreatment: data collected from literature

		[19]; groundwater				[20]										[14]				[8]			
Units in mg/L – otherwise indicated	PDWS	In	Dilu	Conc	% re	In	Dilu			Conc			% re			In	Dilu	Conc	%re	In	Dilu	In	Dilu
Water recovery, %			95.5	4.5			90.0	90.3	90.7	10.0	9.7	9.3	90.0	90.3	90.7		55.6						
Radium, pCi/L		7*	3.3		55.0																		
pH		7.3	8.05		NA	7.05				7.35	7.40	7.57								7.36	7.20	7.62	6.72
Total dissolved solids		1200	628	13,457	50.0											5,532	997	11,186	90				
Silica (SiO_2)		***9.5***	***10***		***NA***																		
Silicon (Si)																				1.6	1.67	3.43	3.47
Potassium (K^+)		22	9.2	296	60.0											101	41	188	59	34.5	9.0	40.5	8.8
Sodium (Na^+)		200	140	1486	33.1											635	149	1197	77	158.5	67.4	178.6	61.7
Calcium (Ca^{2+})		110	42	1567	63.5	284*	59.22	52.94	58.8	2,307	2,435	2,480	81.2	83.2	81.2	816	133	1445	84	101.5	14.0	111.7	9.1
Iron		0.8	0.6	5	28.3																		
Manganese		0.04	<0.02		54.6																		
Magnesium (Mg^{2+})		47	19	647	61.4	130	76.22	72.1	66.4	614	669	750	47.2	14.6	16.4	248	18	569	93	30.2	4.9	30.7	3.1
Lithium (Li)																				0.015	0.007	0.015	0.007
Bromine (Br)																				0.971*	0.212	0.810*	0.120
Rubidium (Rb)																				0.012	0.003	0.014	0.003
Strontium (Sr)																				1.656	0.233	1.809	0.191
Nitrate ($NO^-{}_3$)	10(N)	<0.1	<0.1		14.0											39	11	152	72				
Fluoride	4	1.3	1.2	3	11.8																		
Chloride (Cl^-)		60	46	360	26.8	359	88.78	40.3	27.8	2,791	3,326	3,589	77.7	59.0	37.4	1,485	288	3,332	81	259.8	75.6	285.7	57.4
Sulfate ($SO_4{}^{2-}$)		600	230	8,529	63.4	1066*	364.1	300.3	289	7,383	8,194	8,644	69.3	74.6	75.4	1,814	168	3,911	91	163.9	49.1	146	41.9
Bicarbonate ($HCO_3{}^-$)		290	221	1,769	27.2	297	213.2	181.2	156.9	1,051	1,375	1,663	35.4	23.3	21.4	394	189	392	52				
Hardness as $CaCO_3$		480	183	6,844	63.6																		
Alkalinity as $CaCO_3$																				228.1	55.4	288.4	56.3
Electrical conductivity 2																11,430^3	2,030^3	18,390^3		1576^3	413^3	15343	3593
Temperature, F		72	72																	77	76	73	
Flow rate, gpm		819.5	783		4.5		Lab scale with 1.72 dm^2			Active membrane area						0.396	0.220	0.176					
Linear flow velocity, cm/s							8.60	8.80	90	0.73	0.67	0.63				2 to 7				Continuous		Continuous	

(continued)

Table 1 (continued)

		[19]; groundwater				[20]				[14]				[8]			
Units in mg/L – otherwise indicated	PDWS	In	Dilu	Conc	% re	In	Dilu	Conc	% re	In	Dilu	Conc	%re	In	Dilu	In	Dilu
Type of membrane										Cation-CR67-HMR-412							
Effective membrane, m^2										0.022							
Membrane thickness, mm						Neosepta AMX, CMX, and CMS				0.5							
Chemical added						Without adding any scale formation inhibitor				Anion-204-SXZL-386							
Polar cycle, min						22				15–20							
Spacer tk, mm						0.19				1							
Dead volume in EDR						Eliminated											
$CaSO_4$ supersaturation																	
$CaSO_4$ supersaturation								361 364 368									
LSI 8		Non-corrosive						1.79 1.94 2.12						0.1	−1.4	0.4	−1.8
Ryznar index 6														7	9.9	6.7	10.3
Current density, A/m^2							60.3 67.0 70.0			1st-stage = 20 to 90;							
Limiting current density %							81.5 91.8 94.0			2nd-stage = 20 to 50							
Electrodialyzer Prod Cost, $/1000 gal		1.11				I M Na_2SO_4											

*Targeted pollutant, % re = % removal, 1: Dilu = diluted stream. Meeting the primary drinking water regulations, 2: Unit in μS/cm, 3: at 20°C, 4: at 76–77 F. 5: LSI = Langelier saturation index at 20°C, LSI<0 aggressive water; LSI > 0 scaling water, LSI = 0 equilibrated water, 6: Conc = Concentrated steam, 7: PDWS = primary drinking water standard, 8: Ryznar index at 20°C, 6.0<RI<7.0, slightly aggressive water, 7.5<RI<9.0, very corrosive water, RI > 9.0 intolerable corrosive water.

Table 2

Performance inside EDR without pretreatment or after pretreatment: data collected from literature

Unit in mg/L – otherwise indicated	PDWS[1]	[22]				[21]				[24]							[23]			[23]		
			Dilu[3] [48]	Conc[4]	% re[5]	In[2]	Dilu[3]	Conc4	% re[5]	In[2]	Dilu[3]	Conc[4]			% re[5]		In[2]	Dilu[3]	Conc[4]	In[2]	Dilu[3]	Conc[4]
Source of water										Groundwater							Groundwater					
Water recovery, %		100	75	25		100	59	41		100	30	50	70	50	70	50	Batch mode 250 L/h			Batch mode 600 L/h		
pH		7.8	6.2			7.1	7.1	7.0		7.44	7.38	7.12					7.45	8.1	6.8	7.1	8.30	7.4
Total dissolved solids										817	572	406	1,159	1,622	99.3	99.3						
Manganese						9	7	11.9														
Magnesium (Mg^{2+})									22	71.1	57	20.5	100	140	98.8	98.8	8	5.6	44	5	3.6	28
Sodium (Na^+)						16	15.5	16.7	3								56	50.5	188.5	22.5	16	198
Calcium (Ca^{2+})		166	1.4	659.8	99	174	101	279.0	42	124	68.2	49.0	176	246	99.2	99.2	113	61.5	820	180	96.5	1800
Electrical conductivity 6		3860*	305		92					1,140	799	567					700	488	4,590	931	538	7,520
Nitrite (NO_2^-)						0.1	0.04	0.2	60													
Nitrate (NO_3^-)	10(N)					62*	15	129.6	76	73.2*	14.3	8.2	104	146	99.4	99.4	90*	26.5	1312	155*	16	2,491
Fluoride (F^-)	4	16	0.9	61.3	94																	
Chloride (Cl		888	13	3513	98	16	5.5	31.1	67	92.2	46.1	36.0	131	183	99.2	99.2	42	13	850	53.5	12	1750
Sulfate (SO_4^-)		437	86	1490	80	150	120	193.2	20	207	201	190	292	409	98.9	98.9	30	24.5	34	127	116	433
Bicarbonate (HCO_3^-)						370	231	570.0	38	62.2	47	26.8	88	123	98.8	98.8	350	262	612.5	262.5	187	366
Hardness as $CaCO_3$										12.1	8.2	5.3	17	23	96.8	96.8						
Alkalinity, meq/L																	5.74	4.3	10.04	4.3	3.07	6
COD		41	20		51																	
Flow rate, m^3/d		350	320	30																		
Polar cycle, min		60				20				20	48/28.8	14.4	14.4									
Effective membrane area, m^2		1.60 x 0.80				3.33				500 cm^2 for each = 1000 cm^2 for total.							0.3/each = 0.4 cm^2 tot 3.3/each = 6.7 cm^2 tot					
Type of membrane						Mono valent ion-selective				ACS for anion; CMX-Sb for cation exchange							Monovalent anion-selective membr, neosepta ACS					
Membrane thickness, mm						ACS, CMX-S; Tokuyama Co.				membrane made by Tokuyama Corp.							CMX by Tokuyama. CMX-S, Tokuyama.					
Chemical added		5% HCl in mem cleaning				HCl to avoid $CaCO_3$ scaling				Stack is flushed with 0.1 M HCl to remove							HCl to get nuetral pH avoid $CaCO_3$ scaling					
Chemical used L/m^3		0.3																				
Voltage, V		125								11.6			23.8									
Current, A		60								1.81			3.04									
Power supply, desalination kWh		0.75[7]								0.37[8]			0.43[8]									
Power supply in pumping kWh		13.5[7]																				
Module arrangement		3 stages in series				Single pass				2 stacks in series							2 stacks			2 stacks		
Total # of membranes (pairs)		450				60				60							20			120		
Temperature, °C						24.7	24.5										18	20.5	21.8	21.4	21.8	21.9
EDR trade mark						EUR 6-60				Pilot scale supplied by Eurodia Corp.							Pilot scale EUR2			Pilot scale EUR 6-60		
LSI[7]										−0.02		−0.03		−0.09								

* Targeted pollutant, 1 PDWS = primary drinking water standard, 2 Water into EDR, 3 Dilu = diluted stream, 4 Conc = concentrated steam, 5 % re = % removal, 6 Unit in μS/cm, 7 calculated value, 8 measured value.

Table 3
Performance in pretreatments plus EDR: data collected from literature

Reference	[25]					[26]; No chemical added			
Unit in mg/L – otherwise indicated	Composition original water	Pretreatment	Dilu 1	Conc 2	% removal	EDR in	EDR product	EDR Conc 2	% remov
Water recovery, %	Groundwater		86	14		**RO brine**	85.0		85
pH	7.4		7.2	7.8	4	8.50	4.40		
Total dissolved solids	1,600		500	5,600	69	**4,579**	553		90
Silicon (Si) / Silica (SiO_2)	Si 12		***Si 12***	***Si 12***		SiO_2 **175**	SiO_2 90	SiO_2 **175**	56
Potassium (K^-)	8		2.8	41	65				
Sodium (Na^+)	240		110	1,180	54	1,430	184		89
Calcium (Ca^{2+})	170		41	792	76	147	6		97
Iron	1.5	Yes 4, 5	0.005	0.046	100				
Manganese	0.41	Yes 4, 5	0.001	0.033	100				
Magnesium (Mg^{2+})	88		21	513	76	9	1		91
Barium (Ba)	0.007		<0.001	0.027	100				
Aluminum (Al)	0.039		0.011						
Borom (B)	**0.37**		**0.37**	0.4	0				
P	0.31		0.14	2.7	55				
As	0.023		0.001	0.03	72				
Cr	0.005		<0.001	0.024	100				
Copper (Cu)	0.001		<0.001	0.003	100				
Mo	0.016		<0.005	0.01	100				
Ni	0.008		0.005	0.001	38				
Pb	<0.005		<0.005	0.031					
Ti	<0.001		<0.001	<0.001					
V	0.02		<0.001	0.1	100				
Chloride (Cl^-)	38		5			**878**	16		98
Sulfate ($SO^{2-}{}_4$)	735		246	4,325	67	**2040**	345		86
Bicarbonate ($HCO^-{}_3$)	579		341	2,400	41	62			100
Total hardness	788		289	4,088	63				
Calcium hardness	425		139	1,980	67				
Alkalinity as $CaCO_3$	475		279	1,962	41				
Electrical conductivity 3	2,200		800	7,720	64				
TOC						11.6	5.5		60
Temperature, F	44.6								
Silt density index	0.3–0.5							2.2	
Flow rate [gal/min]						70			
Up-front capital						$45,000 for EDR + $300,000 clarifier 6			
Total cost [$/1000gal]						1.25			

1: Dilu = diluted stream, 2: Conc = Concentrated stream, 3: Unit in μS/cm, 4 = 2% solution of potassium permanganate injection to oxidize iron and manganese followed by manganese greensand filtration. These pretreatments reduce iron and manganese from 1.5 and 0.4 to 0.008 and 0.03 mg/L, 5: A10 μm cartridge filter to remove any particulate matter, 6: Clarifier is used to remove SiO_2 from the dilute stream for EDR as post treatment after EDR.

has 0.004–0.024 in. (0.1–0.6 mm) thickness and is either homogeneous or heterogeneous [3]. Membranes are made from polymer or cellulose blending with ion-exchanging resin; the ions exchanging resin must be at least half of the polymer. The blended mixture is required to heat and then, poured into a supporting open-wave cloth. The properties of ion-exchange membranes (AFN, ACM, AMH, and SB-6407) are shown in Table 6 from literature [28].

Numerous researchers have investigated the possibility of membrane improvement/modifications to reduce fouling potential. The investigations are graft polymerization, membrane stretching, surface modifications, membrane surface roughness, rendering the surface more hydrophilic rather than hydrophobic, altering the surface charge to make it more negative, and reducing the pore size distribution [29]. Through graft polymerization techniques, polymers are attached to the membranes to form brushes or barriers to foulants, preventing them from attaching to the membranes surface. Ultraviolet irradiation is the most common method of polymer grafting surface modifications. However, modified membranes typically exhibit lower initial fluxes, and do not exhibit the same degree of flux decline as unmodified membranes. The performance of the modified membrane is often lower due to the increased thickness caused by the coating layer or grafted polymer.

Table 4
Design criteria and performance of full scale EDR in Prairie, Canada

Parameters [mg/L unless stated]	Groundwater composition	Pretreatment	ERD* Feed	Product	Concentrate	NPDWS	NSDWS
Total flow rate [m^3/d]			2,290	1,929	361		
pH [–]	7.1		7.6	7	7.2	6.5–8.5	
Conductivity [μS/cm]	2,150						
Silt density index [–]	0.3–0.5						Noncorro
Actual TDS	1,620		1,675	433	7,250		500
Design TDS			1,600	490	7,500		
Calcium [Ca^{++}]	**180**		160	31	550***		
Magnesium [Mg^{++}]	90		75	17	333		
Sodium [Na^{+}]	250		240	90	900		
Potassium [K^{+}]	<1						
Iron [Fe^{++} or Fe^{+++}]	**1**	**	0.3		0.3		0.3
Manganese [Mn]	**0.5**	**	0.2				0.05
Barium [Ba^{++}]	0.01					2	
Strontium [Sr^{++}]	1.3						
Aluminum [Al^{+++}]	<0.05				0.05–0.2		
Silica [SiO_2]	**30**	**					
Sulphate [$SO_4{}^{2-}$]	729		713	117	3,333***		250
Chloride [Cl^{2-}]	29		38	9	167		250
Alkalinity [$HCO_3{}^{-}$]	**580**		548	236	1,362***		
Fluoride as F [F^{-}]	<0.01					4	2
Total operation cost with or without blending [$/$m^3$]							0.24

*Trademark Ionics aquamite V; two electrical stages and five hydraulic stages.
Cell pair configuration //60/57/40//60/40//. Reversal every 20 min.
**Two 0.51 m (20 in.) diameter dual media (green sand/ anthracite) filter were used;
NaOCl 1.7 mg/L as Cl_2 was added to the feedwater followed by a potassium
permanganate body feed (1.3 mg/L)
***A scale inhibitor (Flocon 100) was injected into the brine recycle stream of EDR,
at a concentration of 10 mg/L to prevent $CaSO_4$ and $CaCO_3$ scaling of membrane.
Dilute flow = 2,056 m^3/d; brine make-up = 207 m^3/d; brine pump flow = 1,538 m^3/d;
Dilute flow loss = 69 m^3/d; electrode waste = 26 m^3/d; off-spec. product = 59 m^3/d.
Pressure stack = 331–380 kPa; recovery = 84.2 %.
DC power = 1.9 Wh/gal; DC energy = 0.5 kWh/m^3.
Source water temp = 7°C; Operation temp = 6°C; design temp = 7°C.
All data referred from Harries et al. [18].
NPDWS = National primary drinking water standards
NSDWS = National secondary drinking water requlations. noncorro = non-corrosive.

Table 5
Typical water impurities, tolerance limit, and corresponding pretreatments

Impurities	EDR tolerance limits	Pretreatments*	References
Turbidity	2 ntu	Filtration	[3]
Iron	0.3 mg/L	Oxidation and/or filtration	[3]
Maganese	0.1 mg/L	Oxidation and/or filtration	[3]
Hydrogen sulfide	0.3 mg/L	Oxidation	[3]
Chlorine	0.5 mg/L continuous	Carbon adsorption or sodium bisulfate	[3]

*Pretreatment requires if impurity >tolerance limits.

Table 6
The properties of ion-exchange membrane

Membrane	AFN	ACM	AMH	SB-6407
Thickness, cm	0.15–0.20	0.11–0.13	0.26–0.28	0.152
Exchange capacity, meq/g dry membrane	2.0–3.5	1.4–1.7	1.3–1.5	2.15
Electrical resistance, ohm/cm^2	0.4-1.5	4.0–5.0	11.0–13.0	0.3-1.2
Water content, %	40-55	13–18	17–22	50-55
Characteristic	Resistant against organic fouling	Low H$^+$ transport	High mechanical chemical strength	–
Burst strength, kg/cm^2	2–3.5	2–3	12	–

Referred from Akgemci et al. [28].

Ion beam irradiation is mostly used to modify/reduce the roughness of the membrane surface in which ions bombard the membrane surface which smooth out the peaks on the membrane surface [30]. Lower flux decline and less cake layer accumulation was experienced on membranes that had been tested using ion bean irradiation. A fifteen percent reduction in flux decline was demonstrated in this study.

Fouling directly depend on polarization concentration which can be decreased by increasing the mass transfer away from the membrane surface and by reducing the flux through the membrane. High flow velocities and increased turbulence are two common ways to increase the mass transfer coefficient of the membrane system. Traditional membrane feed *spacers* are diamond shaped and are rather susceptible to plugging and fouling, which increases the pretreatment requirements adding cost and complexity to the treatment process.

The ability of membrane to withstand $CaSO_4$ 440% saturation in new aliphatic anion selective membrane (AR 204 ZXZL) at 93.5% calculated water recovery rate (R_c) was demonstrated by [31] in 300 h-test with the high 42% SO_4^{2-} feedwater (SO_4^{2-} 560.7 mg/L in 1829 mg/L TDS) without any pretreatment with acid and SHMP dosed in concentrate. Although the AR 204 ZXZL has the ability to withstand $CaSO_4$ 440% saturation, the metered water recovery rate (R_m) was 86.0% because of high leakage from calculated water recovery rate 93.5%. Due to this hydraulic leak, the metered power consumption in rectifier is up to 1.057 kWh/m^3 product water for R_c 93.5% and R_m 86.0% in Fig. 2. The metered power consumption for R_m 88.14% without acid and SHMP dosing was 0.808 kWh/m^3 product water with the same feedwater above [31]. With the research measured data from [31] in Fig. 2, there is no advantage and there are disadvantages for higher R_c with the additional acids and SHMP adding in concentrate stream due to the hydraulic system leakage and the finite amount of water transferred through the membrane along with ions from dilute into concentrate stream [32] because R_m was not increased and metered kWh/m^3 increased.

2.4. Literature review in spacer

EDR is a multi-stacks device where semi-permeable-cation-membrane and semi-permeable-anion-membrane are alternatively sited between a spacer. The spacer serves as a flow path for the water, supports the membranes, and creates *turbulent flow*. Ion-exchange-membrane is size specific and permselectivity [5]. If more selective membranes are developed and available, ED and EDR designs are more controlled by spacer *hydrodynamics* than by membrane properties [33]. The tortuous-path-flow sheet spacers are tortuous path that makes several 180° bends between the entrance and exit ports of a compartment, determine an effective transfer area of membrane, and create *the turbulence* flow [5]. Water flows along the spacers' flow paths across the surface of the membranes rather than through the membranes as in RO. Spacer design was improved in design to optimally promote *turbulence* which helps minimize colloidal deposition [6]. A very thin membrane-to-membrane distance determined by a 0.19-mm net spacer was applied to achieve a high concentration degree along the membrane in single pass mode [1] without any chemical additives and scale formation inhibitors. The authors [1,10] eliminate the stagnant-region in EDR to reduce the mean cell residence time and its variance.

Spacers usually have the sizes from 18 in. × 40 in. (0.5 m × 1 m) to 3.3 ft × 6.5 ft (1 m × 2 m) and the thickness from 0.02 to 0.04 in (0.5–1.0 mm) [3]. Flow velocity in tortuous-path-flow spacer varies from 35.4 to 68.9 ft/s (18–35 cm/s); maximum pressure for EDR system is restricted to 50 psi (345 kPa). Flow pressure has to maintain in spacers; spacer design and flow velocity are among the controlling parameters in the design of EDR or the current density at which a system

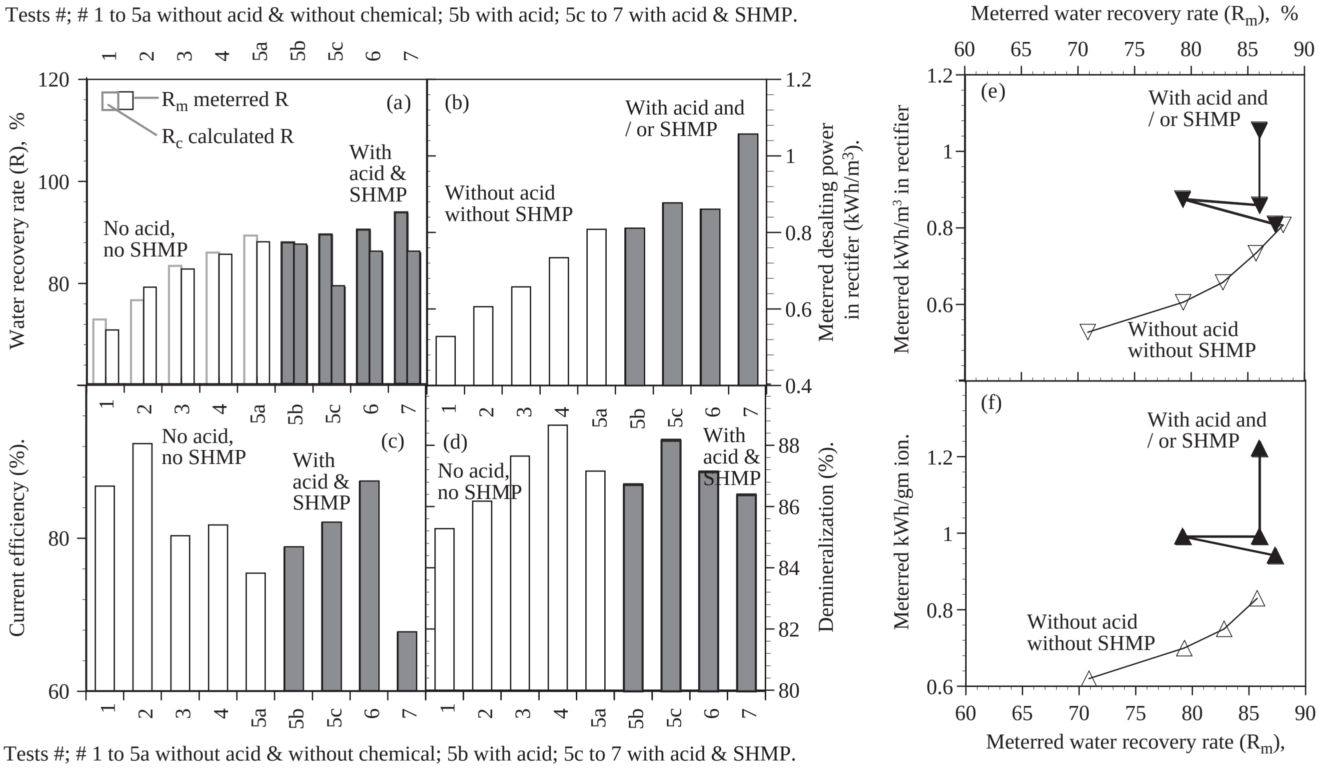


Fig. 2. (a) Tests in different water recovery rates (R) in with and without acid and/or SHMP addition; (b) tests in metered desalting power vs. different *R*; (c) current efficiency tests; (d) demineralization tests; (e) metered *R* vs. metered kWh/m^3 product water; (f) metered *R* vs. metered kWh/g ions. (Pointed markers are measured pilot scale data, are referred from [31]).

should operate. Due to the temperature limit of spacer used, the *upper operating limits of water temperature in membrane stacks* vary; the temperature of 113 F (45°C) is for membrane stack with the low-density polyethylene spacer and 140 F (60°C) is for high quality spacers [3]. There are two types of spacer available – sheet flow spacer and tortuous path flow spacer. The tortuous path flow spacer is designed for the high current density due to the high flow velocities in the channels with the high residence times of the solution in the cells at the same time [33]. The flow velocities are normally 3–10 cm/s in sheet-flow spacer and 15–50 cm/s in tortuous path flow spacer. The pressure drops are 0.5–2 bars in sheet-flow and 1–4 bars in tortuous-path-flow.

2.5. Literature review in electrode

Electrodes are fabricated with inert metal, usually with platinum coated, due to the corrosive acid produced from anode compartments [3] in Fig. 1. The ruthenium-coated titanium *electrodes* and stainless steel electrodes were evaluated in polarization measurement and an accelerated life test by [34] in EDR. The ruthenium-coated titanium electrode performed better in electrochemical characteristics and longer service life than stainless steel electrode because of containing iridium and more titanium. Another development in electrode is that Cl^- ion can attack to the oxide film on the stainless steel electrode for dilute NaCl solution. NO_3^- or SO_4^{2-}/HCO_3^- ions can be added into the NaCl dilute solution to prevent/inhibit the attack on the oxide film on stainless steel electrode. The author [4] also injected hydrochloric acid (110–132 lb/d) into the electrode recycle stream to prevent scaling forming in electrodes. Moreover, these salt or hydrochloric acid was used in cleaning in every 1,500–2,000 h of operation.

2.6. Literature in staging

The required level of charged ions removal in EDR is designed by *staging* the membrane stacks in series. The objectives of *staging* are to grant adequate membrane area and detention time within the stack to achieve the designed level of charged ions removal. *Hydraulic staging* is designed to raise the number of passes the water makes along a membrane stack in EDR. The maximum charged ions removal of each hydraulic stage is 50–67% which is based on the manufacturer recommendation. The increasing charged

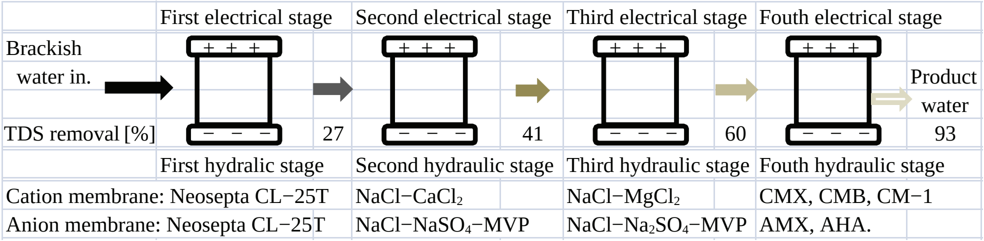


Fig. 3. EDR system with four hydraulic and four electrical stages.

ions removal requires the addition of more *hydraulic stages or membrane stacks* [3] in the system that requires high capacity as in Fig. 3. However, in the small systems, the membrane stacking can be modified in a single stage by inserting one or more thick inter-stage membranes as shown in [3].

Each stack normally reduces 30–60% total dissolved solids at flow rates of 100–130 gpm (6.3–8.2 L/s) and pressures of 50 psig (345 kPa) [3]. Stacks consist of 450–500 cell pairs is normally designed in the plant with 50,000 gal/d (189 m^3/d) of dilute product water. A cell pair is constructed of anion permeable membrane, concentrate spacer, cation permeable membrane, and dilute stream spacer. *Electrical staging* defines the number of electrodes in a membrane stack; typically a single membrane processes one pair of electrode.

2.7. Literature review in limiting parameters

The *limiting parameters* are required to consider for designing a successful EDR plant; these are limiting current density (polarization); current leakage; back diffusion; Langelier saturation index (LSI); calcium sulfate saturation; pressure drop; differential pressure; water transfer; and temperature [3].

Current density may be defined as an electrical flux [A/cm^2]; it is a measure of the mobility of individual ions. The ions freely moving rate through a solution is named as the limiting factor. There should be balance in ions moving rate through the solution and through the membrane. Current density depends on flow velocity in the spacers; the temperature of fluid; and ionic composition. Ions pass through membrane rate is faster than the rate of ions flowing through solution if *current density* increases. On the other hand, H^+ and OH^- are dissociated from water if *voltag*e increases; this process is named the *polarization* point and is experienced in dilute stream. While polarization becomes obvious, the pH of dilute stream changes; H^+ and OH^- pass through the positive and negative exchange membrane, respectively. OH^- has possibility to combine with metal ions (for example magnesium) to form metal salts ($Mg(OH)_2$) as scale, and the scale reduce the efficiency which requires frequent cleaning to prevent the fouling in EDR. Some disadvantages may occur from the operation at the current density which is high enough to cause polarization. These disadvantages include energy utilization increases while electrical resistance increases; currency efficiency decreases due to the un-designed transfer of H^+ or OH^-; further increases of current become inefficiency in transferring ions in polarization areas. To avoid the above disadvantages, a reasonable level of safety factor is included in the design; a factor of 70% of the limiting current density is used as the maximum allowable current density for brackish water operation in a commercial EDR system [3].

An excess of *electric current* at electrodes can leak through a near-by membrane (a heavy anion or cation membrane) into higher conductivity water which is concentrate while comparing with dilute. These *current leakages* can melt membrane and spacers; a factor of 80% is factored into design to prevent these current leakages as design practices restrict the voltage to 80% of the current that would cause burning. The occurrence appears to be color transforms of membranes due to electric current seep out through the membranes and it is essential to address the mechanism of this phenomenon and techniques to prevent such a trouble. The limit of restriction is calculated by temperature of water; source of water characteristics; the size of membrane stack, and the internal manifolds which divide stream into concentrate and dilute.

If the concentration ratio between concentration and dilute is larger than 150, ions move from concentrate into dilute as *back diffusion* against the direction of DC current. The *back diffusion* can be eliminated by lowering the water recovery and restricting the polarization; restricting the polarization may be done by reducing the capacity and quality of recovery dilute water [3]. The pressure of dilute stream is operated 0.5–1 psi (3.4–6.9 kPa) higher than the concentrate stream to prevent the ions back diffuse from the concentrate into dilute stream; the difference in these two pressures is named as *differential pressure*.

LSI is an indicator used to measure the scaling potential of a water source. Scaling in concentrate stream is resulted from the accumulation of calcium (Ca^{2+}), total dissolved solids, and bicarbonate (HCO_3^-). LSI is calculated by the difference between the actual pH and the pH at which the scaling occurs [3]. A positive LSI shows the possibility of scaling and a negative value shows non-scaling possibility [35,3]. To prevent possible scaling, HCl is needed to add in concentrate stream to bring the LSI to +1.8 in the concentrate stream for calcium carbonate control

$$\mathrm{LSI} = \mathrm{pH}_{\mathrm{actual}} - \mathrm{pH}_{\mathrm{scale\ occuring}} \qquad (1)[3]$$

$$\begin{aligned} \mathrm{LSI} = {} & 0.995\log[\mathrm{Ca}^{2+}] + 0.016\log[\mathrm{Mg}^{2+}] \\ & + 1.041\log[\mathrm{HCO}_3^-] - 0.197\log[\mathrm{S}] + \mathrm{pH} - 11.017 \\ & \text{At } 25\,^{\circ}\mathrm{C\ LSI} \end{aligned}$$

where Ca^{2+}, calcium ion concentration; Mg^{2+}, magnesium concentration; HCO_3^-, bio-carbonate concentration; and S, and TDS concentration are in ppm.

Pressure drop through membrane is a function of the type of spacer, flow rate per stage, and number of cell pairs per stage. It includes the sum of the pressure drops through in each hydraulic stage. Typically, EDR requires 50 psi (345 kPa) maximum pressure at the inlet. The pressure will be used up by each hydraulic stage, designing the pressure through the entire system is a major challenge. Techniques to manage pressure drop through systems include adjusting both the numbers and types of spacers; and *feedwater* flow.

2.8. Literature review in preventing $CaSO_4$ scaling with and without any chemicals

The feedwater containing Ca^{2+} 284 mg/L and SO_4^{2-} 1,066 mg/L, was not pretreated and directly feed into EDR to recover 90.0%, 90.3%, 90.7% of product water and generated 2,307, 2,435, 2,480 mg/L of Ca^{2+} and 7,383, 8,194, 8,644 mg/L of SO_4^{2-} in respective concentrate streams without adding scale inhibitor in Neosepta AMX, CMX, and CMS membranes, polar reversal 22 min, and spacer thickness 0.19 mm in modified EDR [20]. The EDR was modified [20] to reduce the dead-end volumes inside the EDR for reducing the mean-ion-resident time (MIRT) in concentrate stream. The feedwater which has Ca^{2+} 816 mg/L and SO_4^{2-} 1,814 mg/L was also not pretreated with cation-CR-67-HMR-412 and Anion-204-SXZL-386 membranes, spacer thickness 1 mm, and polarity 15–20 min [14]. However, the above tests were performed in lab (1.72 dm^2 effective membrane area [20]; and membrane effective area 0.022 m^2 and 0.396 gpm feedwater [14]) in Table 1, and these need to be tested in pilot scale levels. Ca^{2+} 508 mg/L and SO_4^{2-} 1,550 mg/L in feedwater were desalted in Dell city field scale, Texas with Ionics demineralizer-S.N. 638 without any chemical addition for SO_4^{2-} scaling in concentrate stream with SO_4^{2-} 3,000 mg/L [9]. If amount of Ca^{2+} and HCO_3^- in the concentrated water is equal and/or more than 279 and 570 mg/L in mono valent ion-selective ACS, CMX-S membrane from Tokuyama Co., with 20 min polar reversing, the scaling inhibitor (e.g., HCl) is required to add in the concentrated stream to avoid precipitation of $CaCO_3$ in water and scaling in membrane [12]. In case of Prairie, Canada, a scale inhibitor (Flocon 100) 10 mg/L was injected into concentrate stream to prevent $CaSO_4$ and $CaCO_3$ scaling in membranes from Ca^{2+} 550 mg/L, SO_4^{2-} 3,333 mg/L, and HCO_3^- 1,382 mg/L in concentrate stream in Ionics aquamite V with reversal every 20 min [18]. The author [25] injected sodium hexametaphosphate as scale inhibitor into the concentrate recycle stream at a dose of 8 mg/L to prevent scaling in membranes.

2.9. Literature reviews in higher recovery with concentrate stream recycle

2.9.1 High water recovery rate requires waste to recycle back into feed of concentrate

Because there must be maintained as slightest as pressure difference (0.5–1 psi) between the concentrate and dilute streams, the practical flow rates through these two streams are fundamentally equal. These equal flows would result to a 50% recovery of the dilute product water since dilute water quantity equal to only half of the brackish water fed. To gain the higher recovery (>50%) in ED and EDR operation, a small fraction of concentrate flow from concentrate stream necessitates to recycle into the feed concentrate line to equilibrium this identical flow and almost equivalent pressure in both streams that are shown Fig. 1. Theoretically, this fraction is limited by the solubility of ion which is the least solubility in the concentrate stream; the concentrate stream may be recycled up to a level at which the least soluble ion starts to precipitate. Practically, this concentration level is coped by adding a portion of new feed water to the concentrate recycle by wasting the equal amount of the concentrate from the concentrate stream [10].

EDR is user friendly membrane system. During the first design year, membranes are fresh and the performance in EDR is over-designed; the quality of product water from EDR may below the design' water concentrations. One may take advantage of this situation by blending the product water with feed water to increase the additional water recovery rate without any additional cost.

2.9.2. The optimal water recovery rate for high SO_4^{2-} feedwater

Due to the high *water recovery rate (R)* required in arid regions, the unwanted ions are build-up in the concentrate stream along with the recycled water from concentrate into EDR feed [10]. The antiscalant, sodium hexametaphosphate (SHMP) is normally added in concentrate to lower the $CaSO_4$ saturation for preventing $CaSO_4$ precipitation; acids addition is to lower the LSI for avoiding $CaCO_3$ precipitation in concentrate stream. Because the costs of HCl and SHMP dosage in Dirab and Labakha-hawaita, Saudi Arabia are 15 and 83 times higher than power cost [32] in Table 10. The author[38] analyzed to operate equal dimensional EDR with the optimal *R* up to $MIRT_c < 130$ min for eliminating chemical usage to lower the cost; the optimal *R*, depends on species of Ca^{2+}, Mg^{2+}, HCO_3^-, and SO_4^{2-} ions concentrations and TDS in feedwater; polar reversal interval (PRI), can be determined from Monograph proposed by [38].

The optimal *R* was also researched in pilot scale testing by [31] with high 42% SO_4^{2-} feedwater (SO_4^{2-} 560.7 mg/L in 1,829 mg/L TDS) without any pretreatment. The tests in [31] included nine tests (#1, 2, 3, 4, 5, 5a, 5b, 5c, 6, 7) with different *R*; tests # 1 to 5a was without any acid and without any SHMP; test 5b with the acid; and test # 5c to 7 were with acids and SHMP. The results finds from [31] are summarized in Figs. 2(a)–(f). Fig. 2(a) compares the metered *R* with calculated *R*. Fig. 2(b) compares the power consumption in different *R* with and with acid and/or SHMP. Fig. 2(c) shows current efficiency. Fig. 2(d) depicted the % demineralization in different tests. Fig. 2(e) compares power consumption per m^3 of product water with different *R*. Fig. 2(f) analyzes power consumption per gm of ions removed with different *R*. By increasing R_c from 73.0% to 89.4%, the hydraulic leaks are so small that R_m equals to R_c. When R_c is increased from 89.3 to 93.6 by adding acid and SHMP in concentrate stream to avoid sulfate fouling and carbonate precipitation, the hydraulic leaks are considerable that R_m is significantly smaller than R_c in Fig. 2(a). Fig. 2(b) shows the tests in metered power kWh/m^3 product water. One can conclude from Figs. 2(a) and 2(b) that there are disadvantages when the *R* increases from 89.3 to 93.6 in Tests # 5b to 7 with the addition of acid and SHMP; these disadvantages are hydraulic leaks are significant and power consumption (kWh/m^3) increases. The optimal R_m 88.1 can be selected from Figs. 2(e) and 2(f) with the maximal *R* without any acid and chemical from test #1 to 5a.

2.10. Literature review in species, dose, and $MIRT_c$ in concentrate stream of EDR

Species, dose, and mean ion resident time ($MIRT_c$) in the concentrate stream of EDR desalination were analyzed by [38]. In the classical EDR, the dimensions, flow, and velocity of dilute and concentrate are theoretically equal and practically approximately equivalent; LSI and $CaSO_4$ saturation are normally use as indicators to control the scaling and fouling processes in concentrate stream as such LSI < +2.16 for preventing $CaCO_3$ from fouling and $CaSO_4$ saturation level < 200 for averting $CaSO_4$ from precipitation. If LSI is more than allowable limit, acid is required to add in concentrate stream to keep $CaCO_3$ continuing dissolving; if $CaSO_4$ saturation level in concentrate is more than the allowable limit, sodium hexametaphosphate (SHMP) is normally added in concentrate stream to maintain $CaSO_4$ enduring dissolving. EDR, however, was successfully modernized to operate at slower velocity in concentrate stream than velocity in dilute to gain the higher water recovery rate without adding any anti-scalant and without acid; this new EDR operated LSI at 2.29 and $CaSO_4$ saturation level 358.9% at lower dose and lower MIRT in concentrate stream. Dose and MIRT were proposed to address the controlling process in [38]'s monograph which states the requirement to add acids (HCl or H_2SO_4) and SHMP in concentrate stream of EDR are determined by does of (Ca^{2+} +0.016 Mg^{2+}); (Ca^{2+} or SO_4^{2-}); and MIRT. Once can draw conclusion from [38]'s monograph that acid and SHMP are not necessary if the $MIRT_c$ < 130 min.

2.11. Techniques to lower the desalination cost

One way to reduce the desalination cost in EDR is to lower/eliminate the acid and SHMP usage in concentrate stream by lowering dose and MIRT because chemical cost is 14 times higher than desalting power cost in Saudi Arabia demonstration unit (Table 10). EDR in Saudi Arabia demonstration unit consumed 8 L/d of 36% HCl; 0.12 kg/d of SHMP; and desalting power 0.65 kWh/m^3 of product water in 105 m^3/d of product water [32]; with the price of 36%HCl 5.905 $/L (http://www.sciencelab.com/page/S/PVAR/SLH 2206); of SHMP 5.73 $/kg (http://www.chemistrystore.com/Chemicals_S_Z-Sodium_Hexametaphosphate.html);

0.12\$/kWh [39], the cost of acid, SHMP, and kWh are 47.24; 68.64; 8.19 \$/d, respectively.

EDR was successfully operated at LSI 2.29 and 358.9% $CaSO_4$ saturation level with the higher water recovery rate *R* 79.1% without adding any anti-scalants and without any acids in [40] lab by slowing down the velocity in concentrate stream to gain the lower MIRT and lower dose in single pass without any recirculation. Moreover, [23] also demonstrated to operate ED without any recirculation in single pass to gain higher *R* 89.7% by decreasing the volume of concentrate stream for low dose and low MIRT. The advantages of operating EDR in lower dose of TDS, lower MIRT are to reduce the contact time between foulants and the surface of membrane in the concentrate stream, to reduce the dose to which is not high enough to be toxic to the membrane, to increase the life of membrane, and to reduce the current power to attract the ion opposite charge from dilute to concentrate (lower dose TDS) stream. [20] reduced the thickness of cell to 0.19 mm and eliminated all the dead-end volume inside the cell to gain the lower MIRT and its variance and then, achieve the higher *R* (90.0%; 90.3%; 90.7%) in single pass mode [20] without any chemical additives and scale formation inhibitors in high concentrations of Ca^{2+} (284 mg/L) and SO_4^{2-} (1,066 mg/L) from feedwater.

2.12. Literature reviews in modeling

Refs. [41,42,39] were the first to model ED. Refs. [42, 39] modeled ED with the empirical LCD which generated from Lab data. They modeled the same concentrations of brackish water in feed (0.05833 eq/L) and in dilute product (0.005833 eq/L) with the same constants a (25,000 A s^b m^{1-b}/keq) and b (0.5) in LCD calculations, however values of constants a and b are far from the suggested table values from [42]; more researches are needed in this area. Actually, the current density regulates ions transport rate through ion-exchange membranes. The maximum current density terms as limitation current density (LCD) which is the current corresponding to the ions reduction at the surface of membrane on the dilute desalination stream at which the concentration approaches to zero. LCD is defined by the physico-chemical characteristics of feed water; electro-chemical properties of ion-exchange membrane; and hydrodynamic conditions in EDR as follows:

$$\mathrm{LCD} = i_{\lim} = zFC_s^dD / \{(t_m - t_s)\delta\} = zFC_s^dk / (t_m - t_s) \tag{3}$$

where *z* is the number of charges of ions in feedwater, *F* is Faraday constant, 96,485,342 C/mol = 96, 485, 342 A s/mol, C_s^d is concentration of bulk solution in dilute, eq/L, *D* is diffusion coefficient, δ is thickness of boundary layer, t_m is ion transport # through membrane, t_s is ion transport # in solution, *k* is mass transfer coefficient The ions concentrations C_s^d that charged by current have to be enough for the current transport at the membrane interface in ED/EDR cells if the ions concentration is not enough, water splitting may result by the excess current [43].

Ref. [38] modeled EDR with the equations from ED in [42,[34]; In ED, the ions concentrations in concentrate and dilute are not much different. Due to the water recovery rate (*R*) in EDR, the ions concentrations in concentrate in EDR are higher than ions concentration in dilute. The differences in ions concentration between concentrate and dilute depends on *R* [38]. The difference is not significant when *R* 5 to 7; this difference becomes significant when *R* > 7. The higher concentration difference between concentrate and dilute may alter the ions migrate from dilute into concentrate stream; this effect has not yet included in modeling of EDR in literature.

2.13. Membrane life of EDR

Refs. [42,39] modeled with EDR membrane life 5 years; however, [32,44,26] recommended life of EDR membrane 10 design year, and [3] specified 10 years for anion and 15 years for cation membrane. In reality, the life of membrane shall depends on the material used in fabricate membrane; ion dose, ion species, MIRT interacting with membrane. These doses of ions and $MIRT_c$ in concentrate that interact with membrane vary with water recovery rate in EDR [38]; the lower *R* generates lower dose and lower $MIRT_c$ around the membrane in concentrate and vice versa. The ability of membrane to withhold the different ions concentration and different MIRT may be different. The species, dose, and MIRT effects in membrane life have not been researched in literature yet.

EDR successfully competed with RO and nano filtration (NF) in side by side comparisons; the cost per thousand gallons (US\$/kgal) in EDR, RO, and NF are 0.23, 0.27, and 0.29; EDR gained the higher recovery rate (EDR 88.7% vs. RO 73.6% vs. NF 74.1%) with lower cost (EDR 0.061 vs. RO 0.071 vs. 0.077 \$/m^3 of product) by acid adding in the processes [45]. The membrane life was not included in this comparison although EDR is primarily designed to be reversed periodically to prevent the accumulation of $CaSO_4$ fouling and $CaCO_3$ scaling agencies on the membrane surface. This reversal results EDR more forbearing to membrane deterioration and fouling than other membrane technologies such as RO [46].

2.14. Concentration polarization

Concentration polarization occurs in ion-exchange membrane while scaling forming compounds and hydroxides are precipitated and deposited on the surface of membrane; the scaling layers restrict the ions diffusing through the membrane [47]. This concentration polarization consumes high energy and reduces the efficiency of EDR processes. In the membrane separation process, solutes are continuously rejected by membrane, and solutes are accumulated at the surface of the membrane at where solute concentration is always higher than that of the bulk solution. This is defined as concentration polarization (solute build-up). Concentration polarization is reversal and can be minimized by velocity adjustment in cell pair, pulsation, ultrasound, or electric field [48].

Concentration polarization is negligibly low when the potential drop varies by only a tiny fraction from the simple ohmic drop basing on the normal conductivity in a solution when a current is flowing. On the opposite site, concentration polarization is high when as a result of sharp concentration gradients the potential drop is significantly higher than the ohmic drop basing on the normal conductivity [49]. From this one can conclude that concentration polarization is caused by diffusion layers when the potential drop is high enough to differentiate the potential drop by ohm. Concentration polarization may be reduced by proper design and choice of operating conditions. The thickness of diffusion layer is not affected by applied potential, brine concentration, and membrane resistance; the concentration polarization can be managed low by designing sufficient low potential or sufficient low y (short channel length, high velocity flow, and large channel spacing) [49].

2.15. Fouling

Fouling is caused by the rainfall of colloids, an entity having at least in one direction a dimension of between 1 nm and 1 μm [50,51] on the surface of membrane. Colloids in natural water carry negative charged that attracted to positive charged anionic membranes and form the gelatinous films on the membrane surface and lose their charges there. It is constantly the anionic membranes that are suffered in fouling. Colloids are fed and deposited in dilute side of membrane. The degree of fouling occurs in ED and EDR depends on the type of membrane used and the quantity and quality of organic colloids carry in feed water. For example, organic colloid 0.03 mg/L in feed water is dangerous to the anionic membrane in ED and EDR [52,53]. Current reversal and mechanical cleaning of the membrane surface cleared for only a short of period because the voltage raising to its original value before reversal and/or mechanical cleaning instantaneously as high as it had been before reversal and cleaning. The experiment in [53] shown that the current reversal do not remove off the fixed humic acid penetrated inside the body of membrane.

Fouling (microbial adhesion, gel layer formation, solute adhesion) is subjected to the type of membrane. Polyethylene, Neg. Inst.; Parchment, Neg. Inst.; Cast Polyethylene, Neg. Inst.; Selemion, AMT. 10; Cellophane, Neg. Inst.; Neosepta AV-4T are less sensitive to fouling. Ionic 111 B-2 L 183; Ionac MA-3575; A.M.F. A-63 membranes are medium sensitive to fouling. Ionac MA-3148; Aminated P.V.C (negev Inst.) are very sensitive to fouling [53].

Fouling is not reversal. Fouling causes by adsorption of feed components that plugs the pore in membrane and deposes solid on the membrane surface following crystallization and compaction of membrane structure. These adsorption, plugging, and deposition consist of chemical reactions between membrane and components from feed, gel coacervation and microbial growth [47]. The formation of gelatinous films in membrane surface by colloids is very adherent, and the films require special agents to clear [54]. Therefore, it is better to avoid the colloids in feed water by pretreatment.

Fouling, scaling, poisoning, and concentration polarization are primary causes to shorten membrane life time and to decrease permeate flux. Fouling can be minimized by installing proper membrane type and/ or by membrane pretreatment using surfactants, polymers, and enzymes [47]; by practicing intermediate mode of operation [55,47]; by pre-treating using ultrafiltration to reduce SDI_{15} to 0.001. Scaling is precipitated by $CaCO_3$, $MgCO_3$, $Mg(OH)_2$, and $CaSO_4$ on the concentrate side of the surface of cationic membrane, and to a lesser degree to the surface of anionic membrane [53]. Poisoning is originated by the attachment of multivalent and/ or larger counter-ions inside the pore and/ or body of membrane.

2.16. Classical and advance pretreatment

Pretreatment (1) are located prior to or upstream of EDR, (2) effectively change the feedwater characteristics, substances composition, and properties of certain aquatic constituents, and (3) improve the performance of EDR in desalting natural waters and wastewater effluents [51]. The reasons for using pretreatment of the feedwater to low pressure membranes (LPMs) are to enhance the removal of aquatic contaminants, such as micropollutants and disinfection byproduct (DBP) precursors; and to reduce the fouling in surface of membrane. Both validations, in turn, influence the cost and acceptability of EDR and membrane use [51]. Classical pretreatment train, including flush mixing tank, flocculent tank, sedimentation, and media filter, grants water with an average SDI_{15} of 3.1 from seawater [56].

Advance pretreatment may include ultrafiltration. Ultrafiltration eliminates more foulants than dual-media filtration (DMF). Ultrafiltration has abilities to pre-treat seawater of (conductivity 48.7 mS/cm at 20°C; turbidity 0.7 to 5 NTU; color <2.5 mg/L Pt/Co; TOC 0.3 mg C/L; UV254 0.75/m; phytoplankton 20,000-60,000 cell/L; SiO_2 1 mg/L; hydrocarbons < 50 μg/L) from SDI_{15} 13–25 to 0.8 with 60–150 L/(h m^2) at 20°C in 30 min cycle period in aquasource membrane. DMF is capable of pretreatment to the same seawater from SDI_{15} 13–25 to 2.7–3.4 [57]. Moreover, ultrafiltration generates a constant reliable quality of water than DMF which product has larger fluctuation which varies with the turbidity of feed water. However, it is not feasible to operate UF at both high flux rate and high turbidity feed water because severe membrane fouling and plugging in fibers of ultrafiltration.

2.17. Additional wastes streams from EDR

Besides the concentrate waste from EDR after the recirculation point, there are two more waste streams from EDR-waste from rinsing electrodes and off-spect product (OSP). HCl and H_2SO_4 are produced at anode electrode (Fig. 1), and some of these are used in cleaning naturally after each polar recycle. A few residues of O_2, Cl_2, HCl and H_2SO_4 always remain in the anode which is made of titanium and plated with platinum. The residues of HCl and H_2SO_4 are required to rinse with feedwater to prevent shorter life of anode [3]. Similarly, H2 and OH^- are produced near the cathode electrode (Fig. 1). The residue of OH^- always scales in cathode electrode, and scales have to be rinsed with a few portion of feedwater to prevent cathode from scaling. In the both of anode and cathode electrodes compartments, there are an electrode, an electrode water-flow spacer, and a heavy cation membrane. The electrode water-flow spacer serves two purposes – as a barrier which seals the main path of the stack to avoid electrode waste migrates into; and as a mechanism (thicker spacer) to promote the water velocity for self-cleaning of OH^- scaling from cathode. The rinsed water consists of the residues of O_2, Cl_2, and H_2 that are required to degasify in de-gasifier for safely disposing off. The degasified rinsing water from degasifer collected as electrode wastes ($Q^{w.e}$) which are combined into the waste stream (Q^w) (Fig. 1).

Due to its polar reversal between concentrate and dilute streams in a fitted time interval, an OSP water produces from EDR just after concentrate reverses into dilute. In this period, the dilute water contaminates with the residues from concentrate water from previous reversal, and the quality (TDS) of water from this period is always higher than the TDS of normal product. This water defined as OSP, and it is required to dispose off into waste stream or send back into feed stream. Ref. [3] defined this period as 30 s per hydraulic stage due to the period needed for source water entering this new dilute (previous concentrate) compartment to pass through and clean the entire membrane system. The author [58] stated that the average TDS concentration in OSP (C^{osp}) is found to be related with the number of hydraulic stage (N_{hs}), the concentrations in dilute and concentrate streams, and PRI interested and 20 min PRI (PRI_{20min}) in Eq. (4) by mass balance fitting of 62 set of literature data in Fig. 4

$$C^{osp} = 0.03 N_{hs}\,\mathrm{PRI}\left(C^c + C^d\right)/\mathrm{PRI}_{20min} \tag{4}$$

Ref. [3] recommended to delay to open the outlet valves to a level which average TDS concentration of product water is less than the salinity of source water after the reversal. Ref. [3] also recommended to recycle the OSP into feed stream to increase the water recovery rate. However, the method of recycle of OSP into feed increases the TDS of feed water in EDR and subsequently increases the desalting energy.

2.18. EDR application

EDR is especially known for its excellent to desalt high SO_4^{2-} [31] and high NaCl brackish groundwater into drinking water without adding any chemicals and acids. However, with today technology, EDR can successfully desalted wastewaters from coal-mine [59]; from WWTP [2] with ultrafiltration as pretreatment for agricultural purposes; from steel plant with sand filtration as pretreatment [60]; reverse osmosis blowdown [26]. EDR is also capable of removing specific dissolved solids and pollutants such as THMs-Formation Potential (THMs-FP) [61]; bromide [61] and radium [19].

3. Designing alamogordo IEDR desalination project

EDR project design requires information relative to product water quality and quantity requirement, characteristics of the source water to be treated, and the percentage of recovery [3]. The designs of pumps, pipes, valves, and the number of membrane stacks in line are based on the product water quantity requirement.

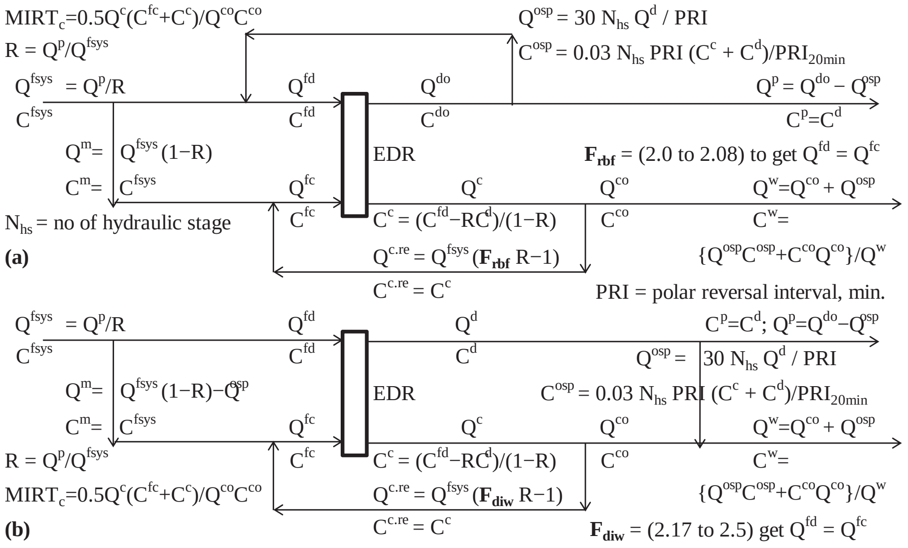


Fig. 4. Mass balance diagrams – (a) OSP re-circulates into feed; (b) OSP wasted directly (Referred from Myint 2010b).

Percentage of water recovery and water qualities in feed and product calculates the membrane stacks configuration or staging. EDR full scale design requires pilot scale studies which provide data to verify the model; aid in burgeoning specific design criteria; and pin down to cost data allowing the best process to be selected for a full-field-scale project. The equations used to design the ILEDR are summarized in Table 7.

3.1. Characteristics of groundwater

The characteristics of groundwater to be treated in ILEDR project are summarized in Table 8. From Table 8, the major pollutants are Ca^{2+} (581), Na^+ (591), Mg^{2+} (396), SO_4^{2-} (,2933), Cl^- (646), HCO_3^- (216) mg/L. These ions concentrations result in total dissolved solid (5,609 mg/L) which is equivalent to 0.085 keq/m^3 in Table 8.

3.2. Proposing pretreatment

Table 1 shows that pretreatment was not required for iron and manganese with their concentrations of 0.8 and 0.04 mg/L; Table 3 shows that pretreatment was required to remove iron and manganese from 1.5 and 0.41 mg/L to 0.008 and 0.03 mg/L [25]. From analyses in Tables 1 and 3, pretreatments of iron are not required in our design in ILEDR with the iron concentration of 0.15 mg/L; however, the pretreatments of are required for manganese because of the concentration (41 mg/L) is much greater than 0.1 gm/L tolerant limit. Therefore, pretreatment in oxidizing with 2% solution of potassium permanganate followed by manganese greensand filtration is needed for ILEDR.

3.3. No adding scale inhibitor

Table 1 shows that Ca^{2+} 284 mg/L and SO_4^{2-} 1,066 mg/L in feedwater was not required pretreatment with Neosepta AMX, CMX, and CMS membranes, polarity 22 min, and spacer thickness 0.19 mm [20] without adding scale inhibitor in concentrate stream in dead-end volume eliminated inside of EDR. Ca^{2+} 816 mg/L and SO_4^{2-} 1,814 mg/L in feedwater was not required pretreatment with cation-CR-67-HMR-412 and Anion-204-SXZL-386 membranes, spacer thickness 1 mm, and polarity 15–20 min [14]. From the analyses in Table 1, pretreatment do not require for the feedwater in ILEDR with Ca^{2+} 481 mg/L, Mg^{2+} 396 mg/L, sulfate (2,933 mg/L), and HCO_3^- 216 mg/L with Neosepta AMX, CMX membranes as long as using polar reversal 15 min and spacer thickness 0.19 mm in dead-end volume minimized EDR.

3.4. Percentage of water recovery

Our project likes to recover as much as water quantity from EDR without adding any chemical into the system. After pretreatment, the targeted main pollutants in raw groundwater are Na^+, Ca^{2+}, Mg^{2+}, Cl^-, HCO_3^-, and SO_4^{2-}. The water recovery rate and the percentage of removal of individual dissolved ion from EDR in our project are referred from the analyses in Table 7 between raw groundwater characteristics and the available literature data [26,14] that have the same similar amount of targeted main pollutants with our project. The percentage of water recovery for our project is set to be 56% from the analyses. From the

Table 7
Equations used in design

Equations used in cost model	Ref.	Eq.
TDS concentration in concentrate out, $\mathrm{keq/m^3} = C_s^{\,c} = (C_s^{\,\mathrm{fd}} - RC_s^{\,d})/(1 - R)$	[42]	(1)
TDS concentration feed to concentrate, $\mathrm{keq/m^3} = C_s^{\,\mathrm{fc}} = (C_s^{\,\mathrm{fd}}\,(1 - R)/R) + (C_s^{\,c}\,(2R - 1)/R)$	[42]	(2)
Practical limiting current density, $\mathrm{A/m^2} = i_{\mathrm{prac}} = si_{\mathrm{emplim.th}}$	[42]	(3)
$g = [\{1\} + \{\Lambda(\rho_A+\rho_C+\rho_{\mathrm{solu}})\,(C_s^{\,\mathrm{fd}} - C_s^{\,d})/tk_{\mathrm{ce}}\}]zFC_s^{\,d}utk_{\mathrm{ce}}\,\alpha$		(4)
$h = [\,\{(C_s^{\,d}/C_s^{\,c}) + 1 + (\Lambda C_s^{\,d}\,(\rho_A+\rho_C+\rho_{\mathrm{Solu}})/tk_{\mathrm{ce}})\}si_{\mathrm{lim}}\,\beta\zeta]$		(5)
$L_{\mathrm{prac.tot}},\ \mathrm{m} = \mathrm{g/h}$		(6)
Ionic strength in solution, $\mathrm{keq/m^3} = I_{\mathrm{stre}} = 0.5\mathrm{Sum}((z_{\mathrm{i}})^2[\mathrm{C_i}])$		(7)
Activity coefficient in dilute $= \gamma^{\mathrm{d}} = \exp(-0.509z_{\mathrm{i}}^{\,2}(I^{0.5}/(1+(I^{0.5}))))$ if $I < 0.1$	[71]	(8)
Activity coefficient in dilute $= \gamma^{\mathrm{d}} = \exp(-0.509z_{\mathrm{i}}^{\,2}(I^{0.5}/(1+(I^{0.5}))- 0.2I))$ if $I < 0.5$	[71]	(9)
Activity coefficient $= \gamma^{\mathrm{c}} = \exp[-0.509z_{\mathrm{i}}^{\,2}\{I^{0.5}/(1+0.3287a(I^{0.5})\}+bI]$ if $I < 1.$	[71, 72]	
Specific desalting energy consumption $= E^{\mathrm{des}}{}_{\mathrm{s}}$, $\mathrm{kWh/m^3}$ product		
$E^{\mathrm{des}}{}_{\mathrm{s}} = [utk_{\mathrm{ce}}\alpha(C_s^{\,\mathrm{fd}} - C_s^{\,\mathrm{d}})^2z^2F^2/(\zeta^2L_{\mathrm{prac\ tot}})][\,tk_{\mathrm{ce}}/\{\Lambda\,(C_s^{\,\mathrm{fd}} - C_s^{\,\mathrm{d}})\,\} + (\,\rho_A+\rho_C+\rho_{\mathrm{Solu}})]F_{\Delta\psi}$	[58]	(11)
Factor corrected for concentration differential $= F_{\Delta\psi} = \ln((C_s^{\,\mathrm{fc}} + C_s^{\,\mathrm{c}})\gamma^{\mathrm{c}}\,/(\mathrm{C_s}^{\mathrm{fd}}+\mathrm{C_s}^{\mathrm{d}})\mathrm{y}^{\mathrm{d}}))$	[73]	(12)
Cell pair voltage drop, $V = U_{\mathrm{cp}} = tk_{\mathrm{ce}}[(1/C_s^{\,\mathrm{c}}) + (1/C_s^{\,\mathrm{d}}) + \{\Lambda(\rho_A+\rho_C+\rho_{\mathrm{Solu}})/tk_{\mathrm{ce}}\,]i_{\mathrm{prac}}\beta((\Lambda)$		(13)
Total membrane surface area, $\mathrm{m^2} = 2N_{\mathrm{ce}}A_{\mathrm{prac}} = gg/h$	[42]	(14)
$gg = [\{1\}+\{\Lambda(\rho_A+\rho_C+\rho_{\mathrm{Solu}})\,(C_s^{\,\mathrm{fd}} - C_s^{\,\mathrm{d}})/tk_{\mathrm{ce}}\}]zFC_s^{\,\mathrm{d}}Q^{\mathrm{d}}$		(15)
Number of cell pair $= N_{\mathrm{cp}} = Q^{\mathrm{d}}/(w\ tk_{\mathrm{ce}}u\alpha)$		(16)
Total current through one cell pair, $A = I_{\mathrm{ce}} = zFQ^{\mathrm{d}}(C_s^{\,\mathrm{fd}} - C_s^{\,\mathrm{d}})/(\zeta N_{\mathrm{cp}})$	[39]	(17)
Total current through all cell pair, $A = I_{\mathrm{tot}} = zFQ^{\mathrm{d}}(C_s^{\,\mathrm{fd}} - C_s^{\,\mathrm{d}})/(\zeta)$		(18)
Potential drop per stack, $V = U_{\mathrm{st}} = U_{\mathrm{cp}}N_{\mathrm{cp}}/N_{\mathrm{st}}$	[42]	(19)
Number of stack in series $= N_{\underline{\mathrm{st}}} = L_{\underline{\mathrm{prac.tot}}}/L_{\underline{\mathrm{st}}}$		(20)

designed % removal of individual pollutant and percentage of water recovery, the amounts of individual pollutant concentrations in dilute and concentrate can be predicted in Table 8, respectively, with the mass balance equation as follow:

$$C_s^{d} = [Q_s^{fd}C_s^{fd} - \{(Q_s^{fd}C_s^{fd})(\%\ \text{ion removal rate})\}/100]\,/\,Q_{\text{prac}}^{d} \tag{4}$$

$$C_s^{c} = (C_s^{fd} - R_{\text{ppf}}C_s^{d})/(1 - R_{\text{ppf}})[42] \tag{5}$$

$$R_{\text{ppf}} = Q_{\text{prac}}^{p}/Q_{\text{prac}}^{f\text{sys}} \tag{6}$$

where C_s^{fd} is concentration of dissolved ions in feed flow of dilute stream into EDR, $C_s^{\,c}$ is dissolved ions concentration in concentrate flowing out from EDR, $C_s^{\,d}$ is concentration of pollutant in dilute stream which flow out from EDR, $Q_{\text{prac}}^{\ fd}$= Practical flow rate of feed water into EDR in dilute stream, $Q_{\text{prac}}^{\ \text{sys}}$ = Practical flow rate feed into EDR system, $Q_{\text{prac}}^{\ p}$is practical product flow rate from EDR system and R_{ppf} is water recovery rate which is the ratio of product to feed

3.4.1. Design of the ILEDR project

EDR is a modified version of ED to reduce the membrane fouling by reversing both the electric polarity and hydraulic flow; EDR requires extra units such as a timing control valve, automatic valves to swap the product and brine streams, and relays to turn around the polarity of the direct current supply when comparing with the ED. On the other hand, EDR gets rid of the acid tanks, complicated agent tanks and mixers, dosing pumps, and pH controllers from the ED process [9]. The design of EDR in membrane, spacer, cell pair selections; hydraulic stages; and electric stages are assumed to be the same as those in ED for these researches. The formula derivations for the designs can be referred from everywhere [39,41,42] and are not repeated in here. The equations used are summarized in Table 7.

3.4.2. Number of cell pairs and number of stage design

The number of cell pairs, number of stage requirement, membrane requirement, and specific power consumption are calculated from the total dissolved solids in Table 9. The equations used to calculate these data are referred from [58] and summarized in Table 7. Data from our permselective design for ILEDR are

Table 8
Predict percentage removal rate for pollutant and percentage water recovery rate

Design parameters	Design parameters from literatures [14]								For ILEDR designs I & II		
Unit in mg/L – otherwise indicated	[14] No chemical added				[14] No chemical added				Measured EDR	Predicted value	
	EDR in	Dilu[2]	Conc[3]	%Re	EDR in	Dilu[2]	Conc[3]	%Re		Dilu[2]	Conc[3]
Water recovery, %		55.6			**Brine**	85		85		56	
pH.					8.5				6.9–7.5		
Total dissolved solid	5,532	997	11,186	90	**4579**	553		90	5,375	446	10,819
Silica, SiO_2					175	90	175				
Potassium, K^+	101	41	188	59					42.5	0	90
Sodium, Na^+	635	149	1,197	76	1,430	184		89	591	0	1257
Calcium, Ca^{2+}	816	133	1,445	84	147	6		97	481	13	1,006
Iron, Fe^{2+} and Fe^{3+}									1.5	0.5	16
Managanese, Mn^{2+}									0.1	0.1	
Magnesium, Mg^{2+}	248	18	569	93	9	1		91	396	7	833
Nitrate, $NO^-{}_3$	39	11	152	72							
Chloride, Cl^-	**1,485**	288	3332	81	878	16		98	646	5	1,358
Sulfate, $SO_4{}^{2-}$	**1814**	168	3911	91	**2040**	345		86	**2933**	396	5730
Bicarbonate, $HCO^-{}_3$	394	189	392	52	62				216	19	435
Carbonate, CO_2-3									48	6	94
Flow rate, m^3/d	2.16				382				68.1	38.2	
Type of membrane	Cation-CR67-HMR-412. Anion-204-SXZL-386.								See Fig. 4.		
Linear flow velocity	2 to 7 cm/s								7.5 cm/s		
Effective membrane	0.022 m^2										
Membrane thickness	0.5 mm										
Polar reversal cycle	15–20 min								15 min		
Spacer thickness	1 mm								0.19 mm		
Dead volume in EDR									Eliminated.		

Feedwater = groundwater from Alamogordo city, NM, USA. 1: EDRin = concentration into EDR. % re = % removal. 2: Dilu = concentration in dilute out. 3: Conc = concentration in concentrate out.

compared with the data from classical whole-sum design in Table 9.

3.4.3. Cost-effective EDR design for ILEDR

AIEDR project is designed for two design – classical design I and mono–di-valent selected design II. All the design details are detailed in Table 9. To be cost effective and to ensure eliminating membrane scaling from $CaSO_4$ and $MgSO_4$ in EDR, four different types of membrane are used in our design II (Fig. 5) based on the specific targeted ions removal. Neosepta CIMS monovalent cation selective membrane and Neosepta ACS-2 monovalent anion selective membrane [62] are used as the first type to recover NaCl (Table 9). NaCl-$CaCl_2$-cation exchange membrane (Ca^{2+} selective; [63]) and NaCl-Na_2SO_4-anion exchange membrane of MVP type, (not to select $SO_4{}^{2-}$; [63] are used in our project as second set of membrane (Table 9). NaCl-$MgCl_2$-cation exchange membrane (Mg^{2+} selective; [63]) and NaCl-Na_2SO_4-anion exchange membrane of MVP type, (not to select $SO_4{}^{2-}$; [63] are used in our project as third set of membrane (Table 9). Neosepta AMX anion-exchange-membrane and CMX, cation-exchange membranes [20] are used in fourth set of design (Table 9). The purpose of the second-, third-, and fourth set of membranes are to separated Ca^{2+} and Mg^{2+} from $SO_4{}^{2-}$; these separation will eliminate the $CaSO_4$ and $MgSO_4$ scaling in the concentrate stream and recover the individual ion for the higher market values. The characteristics and properties of the membranes used in design II are summarized in Table 11. Although design II consists of four different kinds of membranes, however, all these four kinds of membranes are originated from a single brand Neosepta from the same manufacture, Tokuyama Soda, Tokuyama Corp., in Table 11. The purposes to choose the same brand Neosepta and the same manufacture for four different mono- and di-valent selective membranes are to maintain one of the most advantages of EDR which is easily to be assembled and disassembled the stacks to clean or change, both membranes and spacers.

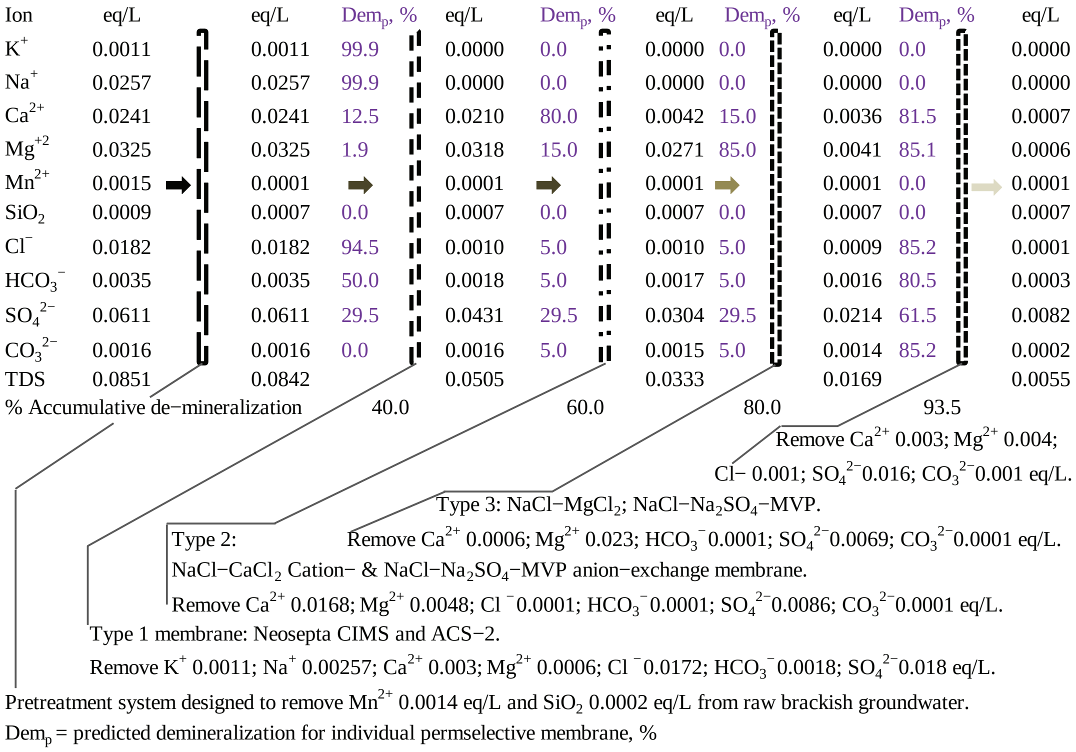
Table 9
Design of ILEDR Desalination Project: Compare Designs I vs. II

Comparison between designs I and II for ILEDR	Design I	Design II for ILEDR						
		Perm-selective; design to selective						
Input data/Type of design	Classic whole-sum	Part I	Part II	Part III	Part IV	Total	Unit/Reference	
Concentration feed in stack, C_s^{fstack}	0.0842	0.0842	0.0619	0.0497	0.0339	0.0842	keq/m^3	
Concentration dilute outlet, C_s^d	0.00583	0.0619	0.0497	0.0339	0.0058	0.0058	keq/m^3	
Recovery rate, $R = Q_p/Q^f_{prac}$	0.56	0.56	0.56	0.56	0.56	0.56		
Equivalent conductance of solution at 20°C, cond	10.5	10.5	10.5	10.5	10.5	10.5	S m^2/keq	
Thickness of cell chamber, tk_{ce}	0.00065	0.0007	0.0007	0.0007	0.0007	0.0007	m	
Linear flow velocity, check type of spacer, u	0.075	0.075	0.075	0.075	0.075	0.075	m/s	[42]
Faraday constant, $F = 96{,}485{,}342$ A s /keq							A S/keq	
Volume factor, $\alpha = tk_{ce}\,A_{appl}/\,(tk_{ce}\,A_{prac})$	0.8	0.8	0.8	0.8	0.8	0.8		
Area factor accounting for spacer-shadow effect, β	0.7	0.7	0.7	0.7	0.7	0.7		
Total area of resistance of membranes, $\rho_A + \rho_C$	0.0007	0.0007	0.0007	0.0007	0.0007	0.0007	Ohm m^2	[42]
Safety factor, s	0.7	0.7	0.7	0.7	0.7	0.7		
Current utilization, ζ	0.9	0.9	0.9	0.9	0.9	0.9		
Production capacity of plant, $Q^d{}_{prac}$ = 7 gal/min	0.00044	0.00044	0.00044	0.00044	0.00044	0.00044	m^3/s	
Effective width of cell, w	0.42	0.42	0.42	0.42	0.42	0.42	m	
Effective length of flow path per stack, L_{st}	0.85	0.85	0.85	0.85	0.85	0.85	m	
Electrochemical valence, z;	z = 1 for K^+, Na^+, Cl^-, HCO_3^-, and z = 2 for Ca^{2+}, Mg^{2+}, SO_4^{2-}, CO_3^{2-}.							
Output								
Feed flow rate, Q^f_{prac}	0.00079	0.00079	0.00079	0.00079	0.00079	0.00079	m^3/s	
Concentration of concentrate outlet, $C_s{}^c$	0.172	0.115	0.065	0.049	0.029		keq/m^3	
Concentration of concentrate inlet, C_s^{fc}	0.108	0.000	0.053	0.037	0.020		keq/m^3	
Theoretical empirical limiting current density	132.5	132.5	90.0	51.0	13.3		A/m^2	
Practical empirical limiting current density, i_{prac}	92.8	92.8	63.0	35.7	9.3		A/m^2	
Total length of flow path, $L_{prac} = g/h$	1.9	2.2	1.8	2.5	4.2		m	
Specific electrical power consumpti, E_s^{des}	4.20	0.38	0.80	0.89	0.06	2.13	kWh/m^3	
Cell pair voltage drop, U_{cp}	1.11	0.48	0.37	0.26	0.12		V	
Total membrane area, $2N_{ce}\,A_{prac} = gg/h$	44	49	40	55	94	238	m^2	
Total # of cell pair, $N_{cp} = Q_{dprac}/(w\,tk_{ce}\,u\,\alpha)$	27	27	27	27	27		Use 27.	[42]
Total current through one cell pair, I_{ce}	0.28	0.12	0.08	0.06	0.04		A	
Potential drop per stack, U_{st}	11.20	4.30	4.10	2.00	0.57		V	
Number of stacks in series, L_{prac}/L_{st}	3.0	5.0	3.0	4.0	6.0			
[3] Monovale = K^+; Na^+; Cl^-; HCO_3^-. [4] Correct applied into practical volume; [5] Correct the spacer effect into account.								
Comparison between classical and mono- and di-valent selected designs								
Electrical cost, \$/m^3 (with \$0.12/kWh [39]).	0.504					0.2556	\$/m^3	
Membrane and capital cost, \$ (150\$/m^2 [39].)	6600					35,700	\$	
Membrane life years 10 years [3, 26, 32, 44]	**10**					10	Year	
Cost per year for desalting electricity \$/yr	7019					3559	\$/year	
Cost per year for membrane, \$/yr	660					3570	\$/year	
Total desalting power and membrane cost, \$/yr	7679					7129	\$/year	
Cost saving, % (at same water recovery rate; same de-minerialization rate; same membrane life)						7	%	

Table 10
Cost comparison: power vs. chemical in EDR (equal dimension in dilute & concentrate)

Reference	Valcour, 1985		Passanisi & Reynolds, 2000
Country	Saudi Arabia		Tanaka, 2007
Plant/ place / town	Dirab	Labakha-hawaita	Port Hueneme, CA, USA
Feed temperature, °C	31	35	
Feed TDS, mg/L	1,273	2,148	1,000
Product TDS, mg/L	91	500	320
Waste TDS, mg/L	13,500	20,609	5,825
Product flow, m^3/d	105	2,000	3,783
Waste flow, m^3/d	10.4		480
Water recovery rate, %	91	90	95
Waste CaSO4 saturation, %	206	270	
Waste Langilier Index	2.00		1.30
Power consumption, kWh/m^3	0.65[1]	1.69[2]	0.337
36% HCl, L/d	8	8.5	
SHMP, kg/d	0.12	2.95	
Power consumption, $/$m^3$ prod	0.078	0.203	0.0233
36% HCl and SHMP, $/$m^3$ prod	1.14	16.93	0.0082
(Cost HCl & SHMP)/cost power	15	83	0.35

1 Desalination power; 2 Energy consumption (EDR system)
Unit price of 36% HCl 5.905 $/L (http://www.sciencelab.com/page/S/PVAR/ SLH2206)
Unit price of SHMP 5.73 $/kg
(www.chemistrystore.com/Chemicals_S_Z-Sodium_Hexametaphosphate.html)
Unit price of electricity 0.12 $/kWh (Tsiakis and Papageorgiuo, 2005)



Ion	eq/L	eq/L	Dem_p, %	eq/L	Dem_p, %	eq/L	Dem_p, %	eq/L	Dem_p, %	eq/L
K^+	0.0011	0.0011	99.9	0.0000	0.0	0.0000	0.0	0.0000	0.0	0.0000
Na^+	0.0257	0.0257	99.9	0.0000	0.0	0.0000	0.0	0.0000	0.0	0.0000
Ca^{2+}	0.0241	0.0241	12.5	0.0210	80.0	0.0042	15.0	0.0036	81.5	0.0007
Mg^{+2}	0.0325	0.0325	1.9	0.0318	15.0	0.0271	85.0	0.0041	85.1	0.0006
Mn^{2+}	0.0015	0.0001	→	0.0001	→	0.0001	→	0.0001	0.0	0.0001
SiO_2	0.0009	0.0007	0.0	0.0007	0.0	0.0007	0.0	0.0007	0.0	0.0007
Cl^-	0.0182	0.0182	94.5	0.0010	5.0	0.0010	5.0	0.0009	85.2	0.0001
HCO_3^-	0.0035	0.0035	50.0	0.0018	5.0	0.0017	5.0	0.0016	80.5	0.0003
SO_4^{2-}	0.0611	0.0611	29.5	0.0431	29.5	0.0304	29.5	0.0214	61.5	0.0082
CO_3^{2-}	0.0016	0.0016	0.0	0.0016	5.0	0.0015	5.0	0.0014	85.2	0.0002
TDS	0.0851	0.0842		0.0505		0.0333		0.0169		0.0055
% Accumulative de−mineralization			40.0		60.0		80.0		93.5	

Remove Ca^{2+} 0.003; Mg^{2+} 0.004;
Cl^- 0.001; SO_4^{2-} 0.016; CO_3^{2-} 0.001 eq/L.

Type 3: NaCl−$MgCl_2$; NaCl−Na_2SO_4−MVP.
Remove Ca^{2+} 0.0006; Mg^{2+} 0.023; HCO_3^- 0.0001; SO_4^{2-} 0.0069; CO_3^{2-} 0.0001 eq/L.

Type 2:
NaCl−$CaCl_2$ Cation− & NaCl−Na_2SO_4−MVP anion−exchange membrane.
Remove Ca^{2+} 0.0168; Mg^{2+} 0.0048; Cl^- 0.0001; HCO_3^- 0.0001; SO_4^{2-} 0.0086; CO_3^{2-} 0.0001 eq/L.

Type 1 membrane: Neosepta CIMS and ACS−2.
Remove K^+ 0.0011; Na^+ 0.00257; Ca^{2+} 0.003; Mg^{2+} 0.0006; Cl^- 0.0172; HCO_3^- 0.0018; SO_4^{2-} 0.018 eq/L.

Pretreatment system designed to remove Mn^{2+} 0.0014 eq/L and SiO_2 0.0002 eq/L from raw brackish groundwater.
Dem_p = predicted demineralization for individual permselective membrane, %

Fig. 5. Treatment trains with 4 different types of membrane based on the feed characteristics.

Table 11
Characteristics and properties of membranes used in design II

Membrane type	1: Neosepta membrane		2: Neosepta membrane		3: Neosepta membrane		4: Neosepta	
	Cation	Anion	Cation	Anion	Cation	Anion	Cation	Anion
Name	CIMS	ACS-2	$NaCl-CaCl_2$	MVP	$NaCl-MgCl_2$	MVP	CMX	AMX
Stability							High mechanically	
Material							Styrene-divinyl benzene copolymer	
Backing/ reinforce	Quaternary ammonium						Polyvinyl chloride	
Thickness, mm	0.15–0.20		0.23–0.24	0.22–0.24	0.23–0.24	0.22-0.24	0.17–0.19[4]	0.16–0.18[4]
Electric resistance, Ωcm^2, 25°C	2.0-2.5		5–6	3–5	5–6	3–5	1.5–1.8[4]	1.4–1.7
Transport number, 25°C	>0.99		0.86–0.88	0.90–0.91	0.86–0.88	0.90–0.91	0.98<[6]	
Water content, g/g dry membrane	0.25–0.30		0.20–0.22	0.28–0.3	0.20–0.22	0.28–0.3		
Ion exchange capacity, meq/g dry	1.4–2.0		1.1–1.2	1.9–2.0	1.1 to 1.2	1.9 to 2.0		1.7[5]
Fixed ion concentration, meq/g H_2O			5–6	6–7	5–6	6–7		
Diffusion constant × 10^8, cm^2/s.			4–5	4–5	4–5	4–5		
Burst strength, kg/cm^2			4–6	6–7	4–6	6–7	5–6[4]	4.5–5.5[4]
Allowable temperature, °C							0–40[4]	0–40[4]
Allowable pH.							0–10[4]–1–10[4]	
Water transport (10g dm^{-3} NaCl)							6 mol/F[4]	6 mol/F[4]
Manufacture	All membranes are Neosepta from the same supplier, Tokuyama Soda, Tokuyama Corp.,							
Reference	[64]	[64]	[63]	[63]	[63]	[63]	4 = [6s]	5 = [66]; 6 = [67]

3.4.3.1. Design II: reasons for selection of four different types of perm-selectivity membrane

The reasons for selection of Type 1 membrane (Neosepta CIMS monovalent cation selective membrane and Neosepta ACS-2 monovalent anion selective membrane) are: CIMS historically reduces high monovalent (K^+ and Na^+ in 99.9% of each) and do not reduce much of Ca^{2+} (12.5%) and Mg^{2+} (1.9%) in Table 12. Literature also recommended ACS is also very good to remove monovalent (Cl^-) at 94.2% and is not good to remove SO_4^{2-} (29.5%) in Table 11. Moreover, Neosepta CIMS and Neosepta ACS-2 demonstrated low energy consumption (155 kWh/t NaCl) and high permselectivity, $((Na^+ + K^+)/Cl^- = 94.5)$ when comparing with Neosepta CL-25T[a] and AVS-4T; and Neosepta CL-25T[b] and AVS-4T. Neosepta CL-25T[a] and AVS-4T; and Neosepta CL-25T[b] and AVS-4T have higher energy consumption (375 and 280 kWh/t NaCl) and lower perm-selectivity $((Na^+ + K^+)/Cl^- = 72.5$ and 90) in Table 13.

The reasons to select Type 2 membrane ($NaCl-CaCl_2$-cation exchange membrane (Ca^{2+} selective) and $NaCl-Na_2SO_4$-anion exchange membrane of MVP type) are: permselectivity and transport number of Ca^{2+} in type 2 membrane are high ($P_{Na}^{Ca} = 2.04$; $t_{Ca2+,m} = 0.401$) and those values of SO_4^{2-} are very low ($P_{Cl}^{S\circ 4} = 0.12$; $t_{SO4,m} = 0.013$). Similarly, the reasons to select Type 3 membrane ($NaCl-MgCl_2$-cation exchange membrane (Mg^{2+} selective) and $NaCl-Na_2SO_4$-anion exchange membrane of MVP type) are: permselectivity and transport number of Mg^{2+} in type 3 membrane are high ($P_{Na}^{Mg} = 1.40$; $t_{Mg2+,m} = 0.313$) and those values of SO_4^{2-} are very low ($P_{Cl}^{S\circ 4} = 0.12$; $t_{SO4,m} = 0.013$). The transport number and perm-selectivity of SO_4^{2-} of $NaCl-Na_2SO_4$-anion exchange membrane of MVP type are lower than those values from $NaCl-Na_2SO_4$-anion exchange membrane of St type in Table 14.

The reasons to select Type 4 membrane (Neosepta AMX and CMX) are: literature shows AMX reduces SO_4^{2-} 58.5% and HCO_3^- 80.5%; CMX removes Ca^{2+} and Mg^{2+} at 81.5% and 85.1%. Type 4 membrane pair (Neosepta AMX and CMX) de-mineralizes 75.2% of TDS in Table 12.

3.4.3.2. Design II: percentage of de-mineralization in each type of perm-selectivity membrane

The demineralization rates of each ion, Fig. 4, in Type 1 membane (Neosepta CIMS, Neosepta ACS) and type 4 membrane (Neosepta CMX and Neosepta AMX) are referred from literature values that are summarized in Table 12.

Table 12. Comparison of performance in different types of membranes

	Ion in feed Seawater [c]	Ion in concentrate of ED Conventional[a]	Monovalent [b] permselective	Demin[d]	Feed 3-Cl mix test NaCl-Na_2SO_4 test	Dilute	Demin[d]	Feed Bolko water	Conc[e] 0.19mm	Demin[d] spacer
Membrane		Neosepta CIM	Neosepta **CIMS**		Duration 25 min. **ACS** & CMX			Neosepta **AMX** & CMX		
	N (%)	N (%)	N (%)	%	mg/L	mg/L	%	mg/L	mg/L	%
Cl^-+Br^-	0.523(90.6)	3.650(99.9)	3.710(99.9)	99.9	15186	888	94.2	359	2791	85.2
SO_4^{2-}	0.054(9.4)	0.003(0.1)	0.003(0.1)	1.9	596	420	29.5	1066	7383	58.5
HCO_3-								297	1051	80.5
Na^+	0.443(76.8)	2.580(70.6)	3.520(94.8)	99.9	10626	240	97.7			
Mg^{2+}	0.108(18.7)	0.760(20.8)	0.067(1.8)	1.9	590	459	22.2	130	614	85.1
Ca^{2+}	0.020(3.4)	0.210(5.7)	0.031(0.8)	12.5	272	211	22.4	284	2307	81.5
K^+	0.010(1.6)	0.110(3.6)	0.100(2.6)	99.9						
TDS								2136	14146	75.2
R										0.90
	All are Neosepta type menbrane from the same supplier, Tokuyama Soda,Tokuyama Corp.,									
Reference	[62]	[62]	[62]	Calculc	[68]	[68]	[68]	[69]	[69]	[20;69]

a Cation exchange membrane, Neosepta CIM manufactured by Tokuyama Corp.,

bMonovalent cation permselective membrane, Neosepta CIMS manufactured by Tokuyama Corp.,

c Concentration and composition of seawater in Tokuyama Bay.

d Demineralization.

e concentrate. [68] Sata, 2004. [69] Quemeneur et al., 2002. [20] Turek et al., 2007. [70] Turek and Dydo, 2003.

Table 13
Progress in ion exchange membrane of electrodialysis

Referred from [64] Mizutani (1990). Year Ion exchange membrane			Electric power electrodialysis kWh /t NaCl	Permselectivity in concentrate, % ($Na^+ + K^+$)/Cl^-
1965	Cation Anion	Neosepta CL-25T[a] Neosepta AVS-4T	370 to 380	70 to 75
1965	Cation Anion	Neosepta CLS-25T[b] Neosepta AVS-4T	270 to 290	90
1987	Cation Anion	Neosepta CIMS[c] Neosepta ACS-2	155	93-96

[a] Unmodified cation exchange membrane.

[b] Modified cation membrane.

[c] Permanently modified cation exchange membrane.

Neosepta AVS-4T and ACS are preferentially permselective for Cl^-.

3.4.4. *Desalting power, membrane area, and cost comparisons between designs I and II*

The desalting power, membrane area, and cost comparisons between designs I and II are done in Table 9. Table 9 shows design I requires more desalting power and less membrane area; however, design II which needs less desalting energy and more membrane area. The purpose of design II area to eliminate the $CaSO_4$ and $CaCO_3$ fouling in membrane by separating the Ca^{2+}, CO_3^{2-}, and SO_4^{2-} in different concentrate line for enhancing the membrane life. By separating the Ca^{2+} from CO_3^{2-} and from SO_4^{2-}, the concentration differential between concentrate and dilute decreases, and subsequently reduces the tendency of permeated ion diffuses back from concentrate into dilute [70]. Permselective membrane also help to reduce this back ion diffusion due to the higher concentration differential [70]. Due to these two and other effects, design II is more economical than design I by 7% without considering the longer life of membrane and without considering the potential for increasing the water recovery rate in design II. If the long life of membrane is considered, the saving increases from 11 (1 year more longer) to 18% (3 years more longer) with the same water recovery rate and the same de-mineralization rate.

3.4.5. *Checking in the number hydraulic stages requirement*

For the design checking [3] recommends that a single hydraulic stage removes 50% of TDS, two hydraulic stages provide 75% of TDS removal, and third-stage gives 90% of TDS removal. For the ILEDR, more than 95% of TDS removal is required in Table 12, and four hydraulic stages are necessary which are shown in Fig. 3. This four hydraulic stage requirement is also met with the requirement from [10] which recommended that each hydraulic removes half of the TDS. Our design also coincidentally calculates four hydraulic stages are required in Table 9. From the design, 108 cell pairs are required for this project; cell pair configuration for our project is arranged to be //27//27//27//27// as four hydraulic stages in series and four sets of electrodes. The TDS removal in the first-, second-, third-, and fourth-stage are 40%, 60%, 80%, and 93.5%.

4. Conclusion

A ILEDR is designed by using the available information from the literature review. The design includes oxidizing with 2% solution of potassium permanganate followed by manganese greensand filtration as

Table 14
Relative transport numbers of divalent selective membranes

	Transport number					
Referred from [63]	Cation	Anion	SO_4^{2-}	Ca^{2+}	Mg^{2+}	Permselectivity P
Cation exchange memb						
NaCl-$CaCl_2$-	0.987	–	–	0.4	–	$P_{Na}^{Ca} = 2.04$
NaCl-$MgCl_2$-	0.985	–	–	–	0.31	$P_{Na}^{Mg} = 1.4$
Anion exchange memb						
NaCl-Na_2SO_4 (St type)	–	0.971	0.017	–	–	$P_{Cl}^{SO_4} = 0.16$
NaCl-Na_2SO_4 (MVP type)	–	0.986	0.013	–	–	$P_{Cl}^{SO_4} = 0.12$

pretreatment, four different types of membranes (Neosepta CIMS and ACS-2 monovalent selective membrane; NaCl-$CaCl_2$-cation and NaCl-Na_2SO_4-anion exchange membrane of MVP type; NaCl-$MgCl_2$-cation and NaCl-Na_2SO_4-anion exchange membrane of MVP type; and Neosepta AMX and CMX, membranes) are used to remove and recover the individual ions separately. This individual membrane design based on the specific targeted ion leads to eliminate the membrane scaling from $CaSO_4$ and $MgSO_4$. The polar reversal interval is 15 min, and sheet-flow stack spacer is selected in design (spacer' thickness 0.19 mm), and dead-end volume is eliminated in inside of EDR. EDR will have four hydraulic stages and four sets of electrodes that has cell pair configuration //27//27//27//27// as EDR. The cost 7–18% saving is found from permselected membrane design with the same water recovery rate and the same de-mineralization rate.

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