Evaluation of sealed-cell electrodialysis for industrial effluent treatment

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Abstract

Sealed-cell electrodialysis (SCED) was evaluated for desalination/concentration of industrial effluents. It was found that relatively dilute (500 to 3 000 mg/ℓ) salt solutions (NaCl, NH_4NO_3, Na_2SO_4, $NaNO_3$, $CaCl_2$) could be effectively desalinated (< 300 mg/ℓ)/concentrated (up to 15,0%). Electrical energy consumption of 0,34 to 5,9 kW·h/m^3 product water was obtained. Brine volume comprised approximately 2% only of the treated feed. However, SCED became less efficient in the 5 000 to 10 000 mg/ℓ feed concentration range due to high electrical energy consumption (3,86 to 13,06 kW·h/m^3).

It was demonstrated that a relatively dilute ammonium nitrate type effluent (3 600 mg/ℓ TDS) could be treated with SCED for water and chemical recovery. Brine volume comprised 2,8% of the treated water and the electrical energy consumption was determined at 2,67 kW·h/m^3. It was shown that it would be difficult to treat a concentrated "ammonium nitrate" effluent (123 700 mg/ℓ TDS) with SCED. However, it would be much easier to desalinate/concentrate a much more dilute effluent (16 557 mg/ℓ TDS). This effluent could be desalinated to 88 mg/ℓ TDS (17,9 kW·h/m^3). Brine volume comprised 8,4% of the treated feed. Nitrate and ammonium removals of greater than 99% were obtained. It was also shown that it would not be possible to desalinate/concentrate an effluent saturated with calcium sulphate with SCED. However, the effluent could be desalinated/concentrated after sulphate removal by chemical precipitation. The effluent was desalinated from 4 401 mg/ℓ TDS to 299 mg/ℓ (%,5 kW·h/m^3). Brine volume comprised 5,3% of the treated feed.

Scale-forming ions (Ca^{2+}; SO_4^{2-}) affect the SCED process adversely due to membrane scaling and these ions should be removed by nanofiltration, chemical precipitation or ion exchange prior to SCED treatment. Sealed-cell ED has potential for treatment of relatively low TDS (< 3 000 mg/ℓ) non-scaling waters for chemical and water recovery. However, higher TDS (up to 16 000 mg/ℓ) waters should also be treated effectively depending on the value of the products that can be recovered.

Introduction

Effluent produced by the mining, chemical manufacturing and chemical process industry and brine derived from demineralisation processes are the main sources of inorganic pollution in South Africa. Inorganic pollution results from tannery operations (Schoeman and Steyn, 1991), electroplating operations (Schoeman and Vorster, 1990), metal pickling operations (Cilliers, 1990), ferrochrome manufacture (Schoeman et al., 1990a), fertiliser manufacture (Schoeman et al., 1988), etc. Some of these effluents, however, contain valuable chemicals (nickel, ammonium nitrate, acid, etc.) that can be recovered for reuse. Scarce water can also be recovered for reuse. Therefore, treatment of these effluents with suitable technologies can turn pollution into profit.

Electrodialysis (ED), reverse osmosis (RO), ion exchange (IX) and evaporation are desalination/concentration processes potentially suitable for water and chemical recovery from industrial effluents (USAID Desalination Manual, 1980; Nishiwaki, 1972; Porter, 1990). Of these processes, ED appears most suitable when large volumes have to be treated, high brine concentrations (small volumes) are required and waters with relatively low membrane fouling and scaling characteristics are encountered (Schoeman, 1985).

Many examples are given in the literature to demonstrate how ED can be used for desalination/concentration of model waters and industrial effluents (Millman and Heller, 1982; Schoeman, 1985; Smirnova et al., 1983; Nishiwaki, 1972). The filter press ED stack with sheet or tortuous flow paths is mainly used for these demonstrations. Brine is circulated through the brine compartments and relatively high brine concentrations have been reported (up to 20%). Water recoveries in excess of 80% can be obtained.

Electrodialysis stacks have been operating smoothly and efficiently for approximately 20 years now for brackish water desalination. However, there is a feeling among certain membrane experts that ED is more complicated and more difficult to operate than RO. Therefore, a very simple ED membrane stack, called sealed-cell ED (SCED), was developed by Kedem et al. in 1978. In this design, the membranes are sealed together at their edges to form membrane bags. Many of these bags can be placed between a pair of electrodes in a very simple membrane stack. No brine is circulated through the brine compartments. Therefore, higher brine concentrations should be obtained with this technique than with conventional ED.

The objectives of this paper will be to describe the performance of a SCED unit for industrial effluent treatment. The SCED unit was evaluated firstly for treatment of synthetic salt solutions normally encountered in industrial effluents and secondly for concentration/desalination of selected industrial effluents.

Bench-scale ED stack

A simplified diagram of the SCED membrane stack is shown in Fig. 1. The brine sealed cells with outlets are arrayed in an open vessel, separated by spacers (0,3 mm). The dialysate enters through a suitable port at the bottom of the vessel and runs out through an overflow. Direct current is applied through carbon suspension electrodes (Kedem et al., 1978). The external dimensions of the sealed brine cells are 60 x 180 mm, giving an effective membrane area of 100 cm^2 per cell pair (cp).

The potential per cell pair, V_{cp} (in volt), at a given current density (i = I/A, mA/cm^2), is the sum of several terms (Kedem and Cohen, 1983):

$$V_{cp} = V_n + i\,(R_m + R_p + R_d + R_b) \qquad (1)$$

where V_n is the concentration potential, a counter driving force built up by the concentration process. Its magnitude depends on the concentration ratio between the brine and dialysate and the

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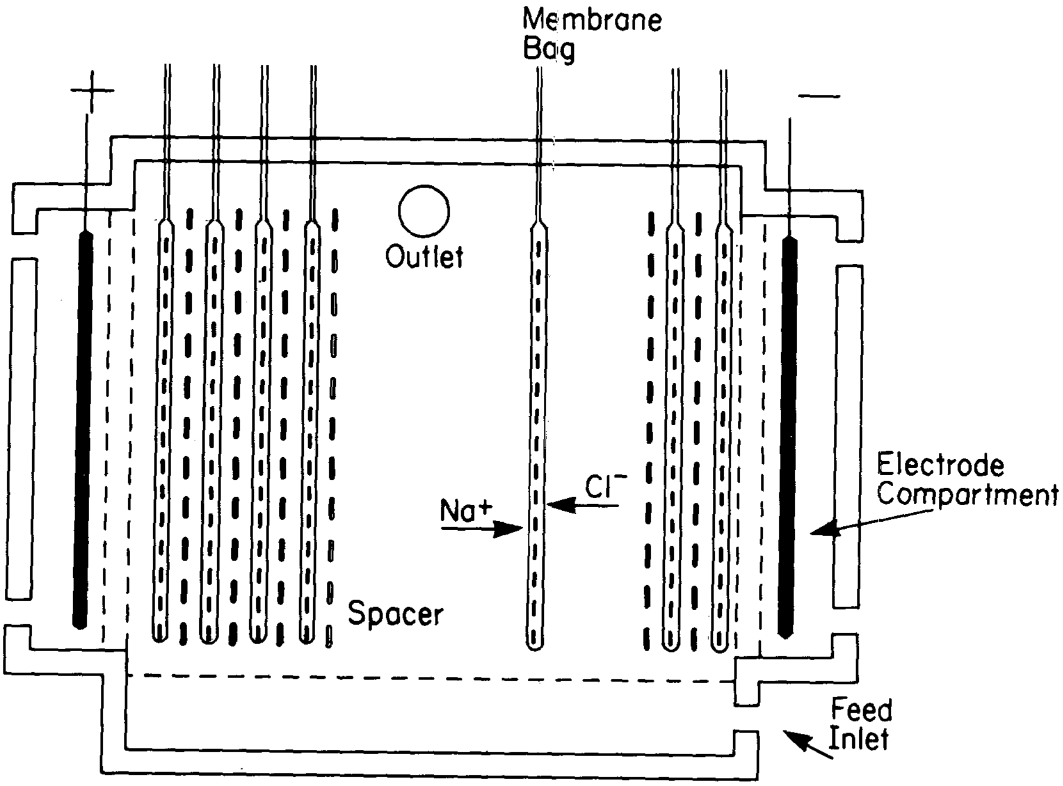

Figure 1
Simplified diagram of SCED membrane stack

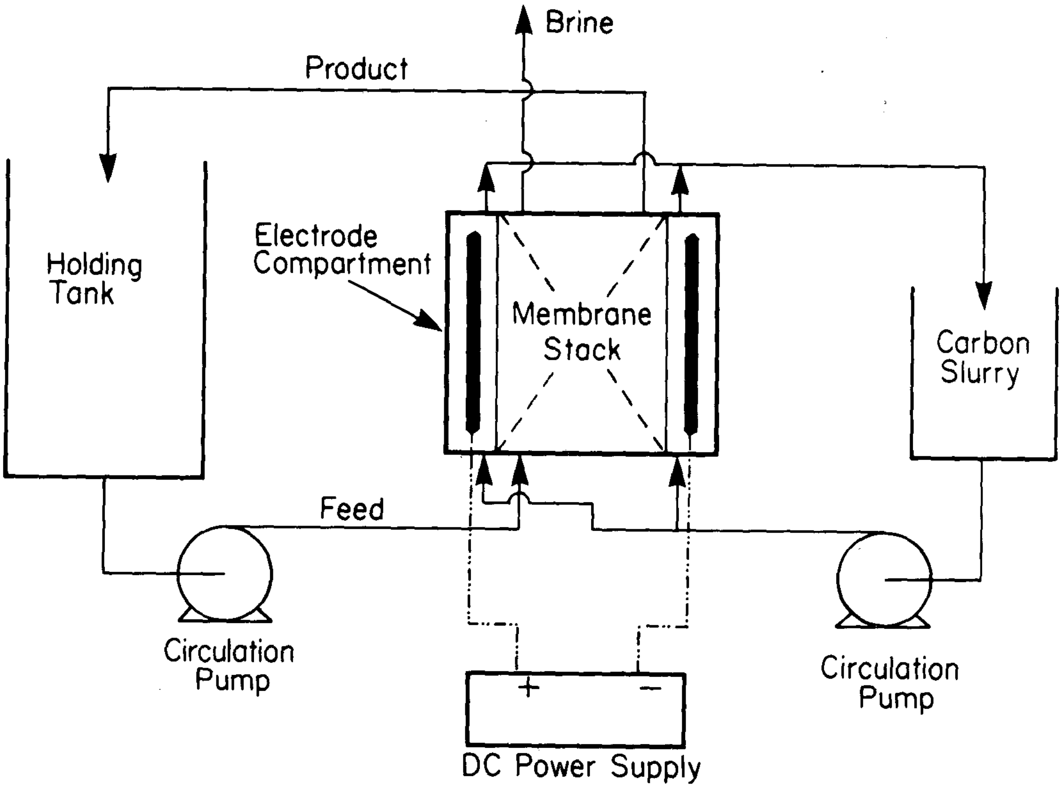

Figure 2
Simplified diagram of SCED system

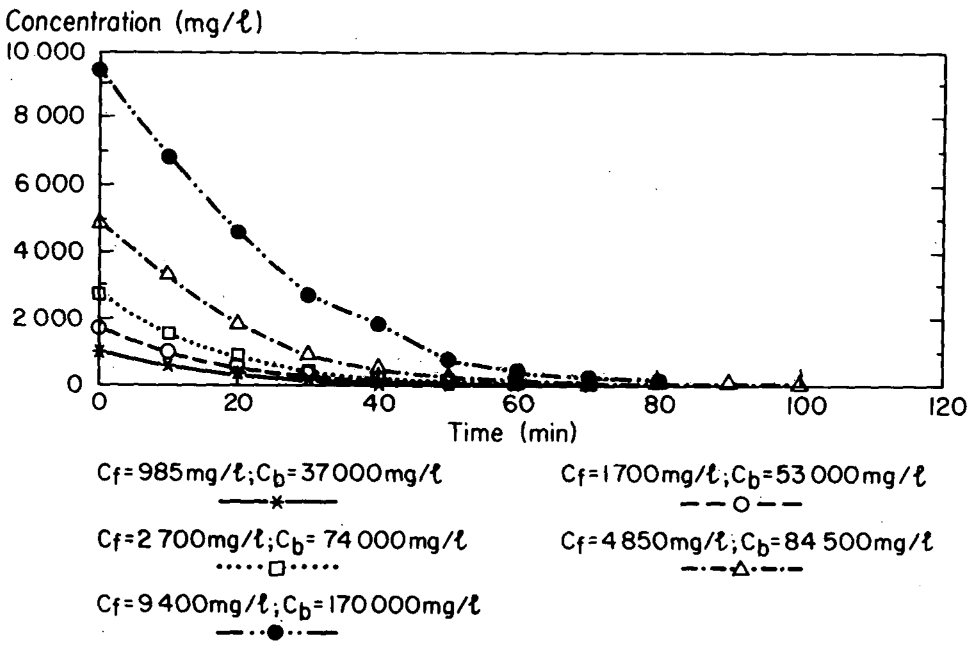

Figure 3
Desalination/concentration of different sodium chloride feed concentrations at 1,76 V/cp

permselectivity of the membrane at the given conditions. V_n is measured during interruption of the current for a few seconds — long enough to disperse concentration gradients near the membranes, short enough to avoid changes of bulk concentration.

$\frac{V_{cp}-V_n}{i}$ is the resistance of the cell pair; R_m membrane resistance; R_b brine compartment resistance; R_d dialysate compartment resistance; and R_p the ohmic resistance and additional counter potential due to polarisation layers adjacent to the membrane surface facing the dialysate. In this system, R_b is negligible, since the brine is always more concentrated than the dialysate. For the simplest characterisation of the system, it can be written (Kedem and Cohen, 1983):

$$R_{cp} = \frac{V_{cp}-V_n}{i} = R_m + \rho d_{eff} \qquad (2)$$

where ρ is the specific resistance of the dialysate solution, and d_{eff} is the effective thickness of the dialysate compartment. In this simple representation the shadow effect of the spacer, polarisation layers and any other ~~possible~~ disturbances are lumped into d_{eff}.

The measurement of voltage and current during desalination at a given circulating flow velocity gives the stack resistance as a function of concentration. If desalination is carried out at constant voltage, straight lines are obtained for a plot of cell pair resistance (R_{cp}) as a function of the specific resistance of the bulk dialysate solution (ρ) in a wide range of currents and concentrations (C). This is due to nearly constant i/C, which determines, at given bulk flow, the polarisation. Straight lines show not only that R_d, but also that the contribution of polarisation, is an approximately linear function of bulk dialysate resistance.

In the unit cell flow regime the brine is not circulated but evolved in the sealed-cell. Brine enters the cell as electro-osmotic and osmotic water. Brine leaves the stack by electro-osmotic pumping. This usually leads to a very high concentration factor (CF) and thus high recovery of product and small volume of brine to be disposed of. This high concentration factor, however, can cause enhanced scale formation. Therefore, a pre-softening step by ion exchange may be necessary under certain circumstances.

Membranes

The ion-exchange membranes used for the investigation were developed at the Weizmann Institute of Science in Israel (Kedem, 1986). The membranes were made from microbeads of styrene-divinyl benzene copolymer which were modified to cation- and anion-exchange particles. The cation-exchange particles were formed by chlorosulphonation with chlorosulphonic acid followed by hydrolysis to yield the sulphonated product. The anion-exchange particles were formed by chloromethylation followed by amination with triethylamine to yield the anion-exchange particles.

The ion-exchange membranes were formed by casting a suspension of the particles on a fabric. The suspension was evaporated to dryness to yield the dry membrane. The cation- and anion-exchange membranes were then heat-sealed to give the membrane bags. The membrane bags were tightly packed between the electrode compartments with spacers (0,3 mm) between them.

Experimental

Solutions of sodium chloride, ammonium nitrate, sodium sulphate, sodium nitrate and calcium chloride (in deionised water) of different initial concentrations were desalinated/concentrated at different cell pair voltages in the SCED unit. Feed (C_f), product (C_p) and brine (C_b) concentrations were determined from conduc-

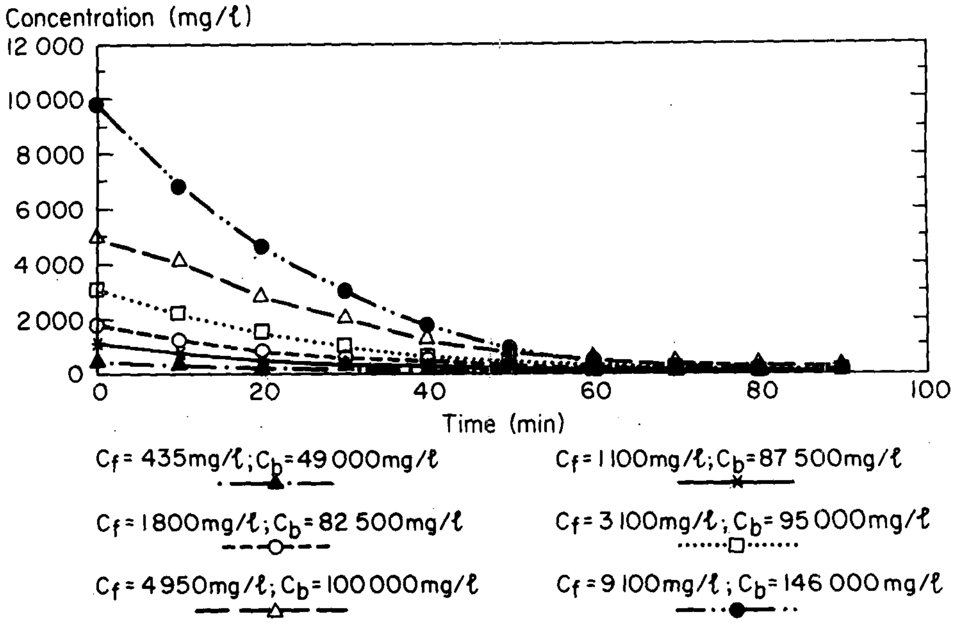

Figure 4
Desalination/concentration of different ammonium nitrate feed concentrations at 1,18 V/cp

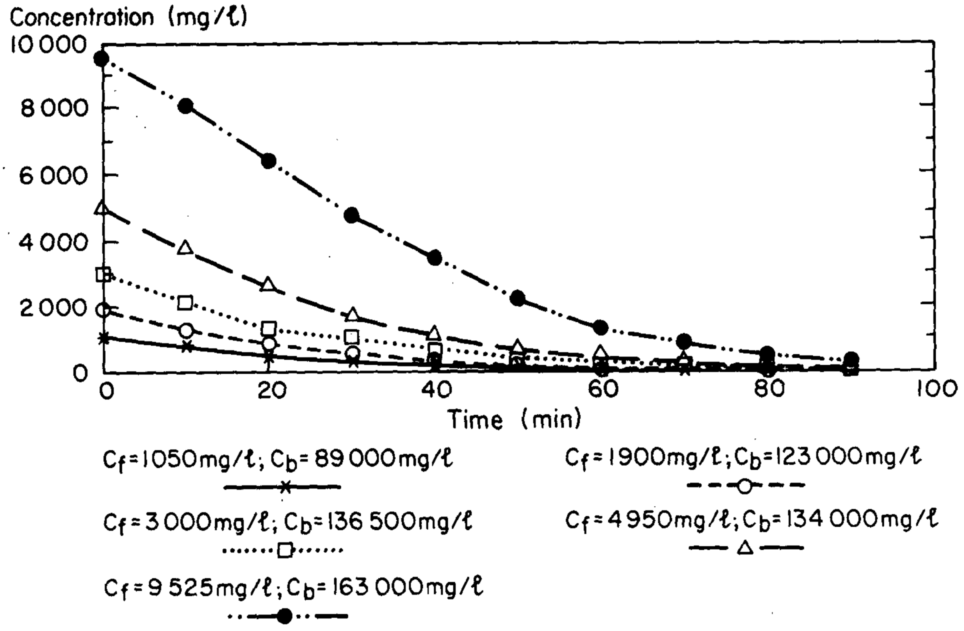

Figure 5
Desalination/concentration of different sodium sulphate feed concentrations at 1,76 V/cp

tivity measurements.

Feed solution (15 ℓ) was circulated at a linear flow velocity of 15 cm/s through the dialysate compartments (Fig. 2). The electrode solution consisted of 2 ℓ of a 2% carbon slurry in 1 mol/ℓ sodium chloride solution. The pH of the solution was adjusted to approximately 5 and circulated through the electrode compartments.

Electrodialysis was started by applying a DC voltage of approximately 0,5 V per cell pair across 17 membrane bags. Voltage between the membrane bags was measured with calomel electrodes connected to a salt bridge. Current was recorded at 10 or 20 min intervals during ED and V_n was determined during interruption of the current for a short period. The final brine volume, concentration of the desalinated feed (product water) and brine were determined at the end of the runs.

Current efficiency (CE), water recovery (WR), brine volume (BV), electrical energy consumption (EEC), concentration factor (CF), output (OP) (water yield), effective thickness of diasylate compartment (d_{eff}), and membrane resistance (R_{cp}) were determined from the experimental data. Graphs were plotted of feed water concentration, brine concentration, current efficiency and electrical energy consumption as a function of time and of cell pair voltage as a function of the specific resistances (ρ) of the dialysate.

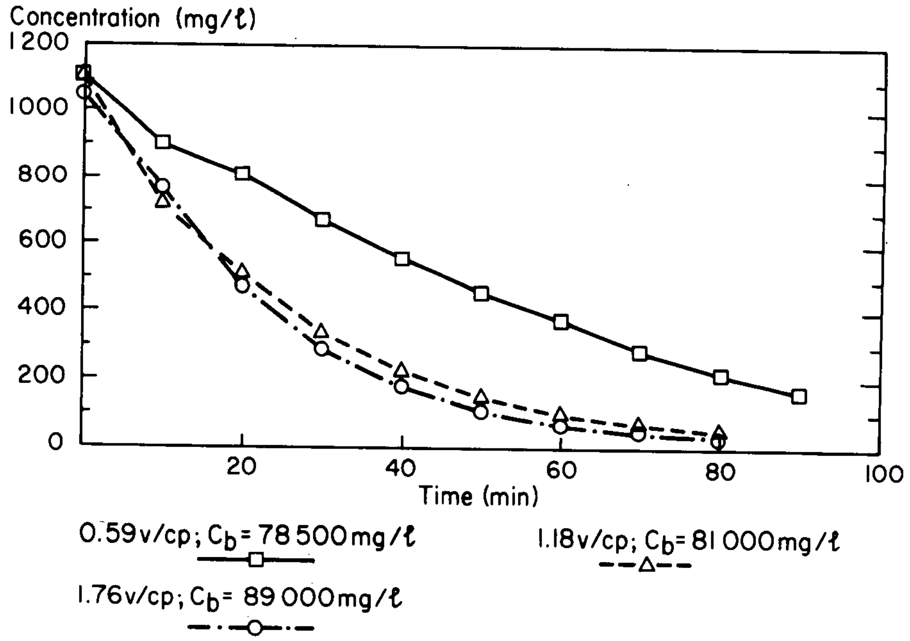

Figure 6
Desalination/concentration of sodium sulphate solutions at different cell pair voltages

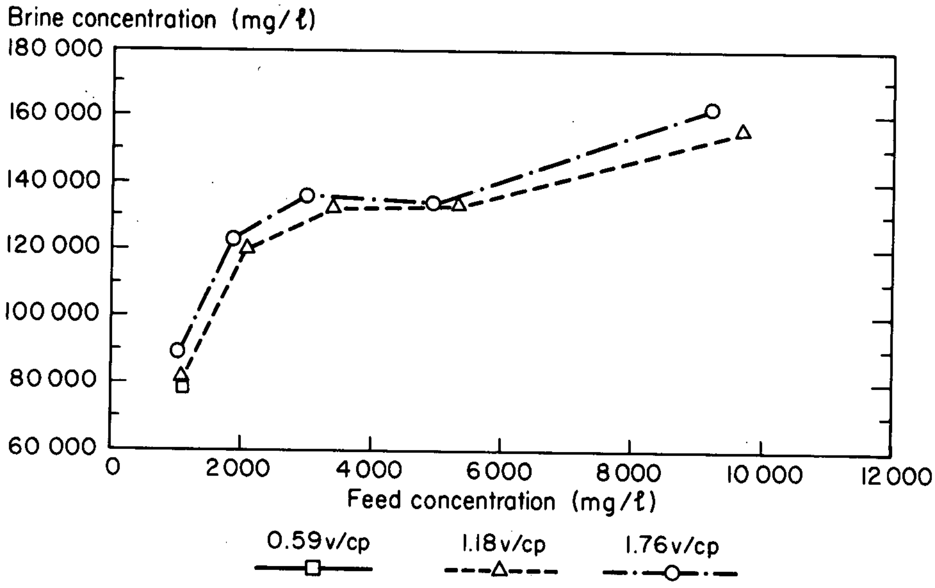

Figure 7
Brine concentration as a function of sodium sulphate feed concentration and cell pair voltage

Results and discussion

Desalination/concentration of salt solutions with SCED

Desalination rate, product and brine concentration

A summary of the desalination/concentration results of the different salt solutions is shown in Tables 1 to 5.

Examples of the desalination/concentration of sodium chloride, ammonium nitrate and sodium sulphate solutions as a function of time at constant cell pair voltage are shown in Figs. 3 to 5. The effect of increasing cell pair voltage on desalination/concentration of an approximately 1 000 mg/ℓ sodium sulphate solution is shown in Fig. 6.

Desalination rate decreased with decreasing feed concentration (Figs. 3 to 5) and decreasing cell pair voltage (Fig. 6). However, approximately the same initial desalination rate was obtained at 1,18 and 1,76 V/cp (Fig. 6). The optimum cell pair voltage for desalination regarding polarisation and electrical energy consumption should be determined for each feed concentration, because this information is required to operate an ED stack under optimum conditions. This, however, was not the main purpose of this investigation. The main purpose of this investigation was to evaluate the performance of the SCED unit for desalination/concentration of

TABLE 1
DESALINATION/CONCENTRATION RESULTS OF SODIUM CHLORIDE SOLUTIONS AT DIFFERENT CELL PAIR VOLTAGES

Vcp	Cf mg/ℓ	Cp mg/ℓ	Cb mg/ℓ	CF	CE %	WR %	BV %	EEC kWh/m³	OP m³/m².d	d_{eff} mm	Rcp ohm-cm²
0,59	1010	282	22 450	22,20	72,20	96,00	4,00	0,34	1,22	0,95	38,8
1,18	950	35	31 000	35,40	66,70	96,30	3,70	0,77	1,53	1,01	39,2
	1 900	40	53 500	28,10	73,70	96,50	3,50	1,41	1,36	0,91	35,1
	3 400	125	72 000	21,20	56,40	96,40	3,60	3,26	1,36	0,97	21,3
	5 400	65	82 000	15,20	78,60	94,80	5,20	3,86	1,20	0,88	18,2
	10 200	195	161 000	15,80	67,90	93,50	6,50	8,04	1,19	0,87	14,4
1,76	985	25	37 000	37,70	63,90	96,70	3,30	1,25	1,75	1,09	46,5
	1 700	25	53 500	31,10	67,80	96,40	3,60	2,07	1,53	1,08	32,9
	2 700	48	72 000	27,40	55,20	96,50	3,50	3,74	1,75	1,05	26,7
	4 850	25	82 000	17,40	69,60	94,60	5,40	5,82	1,20	0,90	21,8
	9 400	120	161 000	18,10	71,90	94,00	6,00	11,10	1,49	0,95	15,1

Vcp = cell pair voltage
Cf = feed concentration
Cp = product concentration
Cb = brine concentration
CF = concentration factor
CE = current efficiency

WR = water recovery
BV = brine volume
EEC = electrical energy consumption
OP = output
d_{eff} = thickness of dialysate compartment
Rcp = cell pair resistance

TABLE 2
DESALINATION/CONCENTRATION RESULTS OF AMMONIA NITRATE SOLUTIONS AT DIFFERENT CELL PAIR VOLTAGES

Vcp	Cf mg/ℓ	Cp mg/ℓ	Cb mg/ℓ	CF	CE %	WR %	BV %	EEC kWh/m³	OP m³/m².d	deff mm	Rcp ohm·cm²
0,59	580	240	58 000	100,00	29,70	99,30	0,70	0,23	1,58	0,97	25,6
	1 010	230	80 000	79,60	43,50	98,90	1,10	0,35	1,26	0,97	24,6
1,18	435	50	49 000	112,60	28 70	99,30	0,70	0,54	1,58	0,67	68,2
	1 100	55	87 500	79,60	51,80	98,80	1,20	0,80	1,39	0,84	38,6
	1 800	90	82 500	45,80	45,80	98,30	1,70	1,50	1,39	0,80	38,2
	3 100	125	117 630	30,70	48,20	98,00	2,00	2,45	1,38	0,75	20,2
	4 950	190	100 000	20,20	47,20	97,20	2,80	4,09	1,37	0,79	14,5
	9 100	320	146 000	16,00	49,40	95,30	4,70	7,37	1,21	0,85	14,7
1,76	420	42	64 500	153,50	22,40	99,00	1,00	1,00	1,58	0,85	45,3
	1 300	60	78 000	60,00	36,30	98,70	1,30	2,05	1,57	1,14	35,6
	1 800	35	120 000	66,70	41,70	98,50	1,50	2,54	1,39	0,87	28,8
	2 800	35	150 000	53,60	47,20	98,10	1,90	3,55	1,24	1,02	19,0
	4 525	45	136 500	30,20	47,20	97,30	2,70	6,78	1,24	1,06	12,8
	9 800	70	130 000	13,30	46,50	94,70	5,30	13,09	1,20	0,87	11,2

TABLE 3
DESALINATION/CONCENTRATION RESULTS OF SODIUM SULPHATE SOLUTIONS AT DIFFERENT CELL PAIR VOLTAGES

Vcp	Cf mg/ℓ	Cp mg/ℓ	Cb mg/ℓ	CF	CE %	WR %	BV %	EEC kWh/m³	OP m³/m².d	d_{eff} mm	Rcp ohm·cm²
0,59	1 110	165	78 500	70,70	79,30	98,90	1,10	0,27	1,40	0,84	65,6
1,18	1 100	50	81 000	73,60	71,90	98,70	1,30	0,66	1,57	0,99	57,2
	2 100	70	120 000	57,10	71,70	98,50	1,50	1,28	1,39	0,99	47,7
	3 400	95	132 000	38,80	76,20	98,10	1,90	1,97	1,25	0,75	37,2
	5 350	445	133 000	24,80	62,30	97,50	2,50	3,59	1,24	1,02	32,3
	9 700	1 500	156 000	16,08	63,20	96,50	3,50	6,01	1,23	0,89	28,6
1,76	1 050	30	89 000	84,80	52,70	98,30	1,70	1,31	1,56	1,11	59,0
	1 900	35	123 000	64,60	63,40	98,50	1,50	1,99	1,39	1,25	42,8
	3 000	77	136 000	45,50	76,20	98,20	1,80	3,20	1,25	1,14	45,6
	4 950	65	134 000	27,10	62,30	97,50	2,50	4,75	1,24	1,25	29,9
	9 525	180	163 000	17,11	63,20	96,10	3,90	13,85	1,23	1,17	23,2

TABLE 4
DESALINATION/CONCENTRATION RESULTS OF SODIUM NITRATE SOLUTIONS AT DIFFERENT CELL PAIR VOLTAGES

Vcp	Cf mg/ℓ	Cp mg/ℓ	Cb mg/ℓ	CF	CE %	WR %	BV %	EEC kWh/m³	OP m³/m².d	deff mm	Rcp ohm·cm²
0,59	1 100	465	65 000	59,30	41,50	98,90	1,10	0,28	1,57	1,01	28,8
1,18	1 000	90	63 500	63,3	47,0	98,6	1,40	0,73	1,57	0,99	32,1
	1 950	100	71 000	36,5	65,0,	98,4	1,60	1,07	1,39	1,01	30,4
	2 800	100	82 000	29,3	63,2	98,1	1,90	1,61	1,38	0,83	29,7
	5 000	140	102 000	20,5	56,67	97,3	2,70	3,29	1,24	0,86	19,3
	10 100	530	123 000	12,2	53,1	96,0	4,00	6,98	1,22	1,02	10,2
1,76	1 000	70	60 500	60,30	40,70	98,50	1,50	1,30	1,57	1,16	33,6
	2 100	60	69 500	33,10	51,30	98,20	1,80	2,25	1,39	1,12	28,3
	2 800	50	81 000	29,00	53,80	98,00	2,00	2,90	1,38	1,06	25,3
	5 200	90	117 000	22,50	55,00	97,10	3,90	5,34	1,23	1,27	17,0
	9 800	150	125 000	12,80	51,80	95,60	4,40	10,85	1,21	1,27	10,7

TABLE 5
DESALINATION/CONCENTRATION RESULTS OF CALCIUM CHLORIDE SOLUTIONS AT DIFFERENT CELL PAIR VOLTAGES

Vcp	Cf mg/ℓ	Cp mg/ℓ	Cb mg/ℓ	CF	CE %	WR %	BV %	EEC kWh/m³	OP m³/m².d	deff. mm	Rcp ohm·cm²
0,59	1 100	310	42 000	38,20	47,80	98,70	1,30	0,48	1,57	0,93	40,3
1,18	970	50	41 200	42,50	45,70	98,50	1,50	1,17	1,56	1,05	36,4
	2 100	110	51 000	24,30	49,70	97,80	2,20	2,34	1,38	1,15	27,5
	2 950	160	57 000	19,30	46,30	97,20	2,80	3,53	1,37	1,19	19,9
	5 000	230	75 000	14,00	45,70	95,80	4,20	6,21	1,22	1,19	15,4
	10 300	940	75 000	7,30	44,30	92,70	7,30	13,06	1,18	1,12	9,6
1,76	840	20	38 500	45,80	36,50	98,50	1,50	1,94	1,56	1,18	34,7
	2 000	35	45 500	22,80	48,10	97,80	2,20	3,57	1,38	1,32	28,2
	3 000	85	54 500	18,20	43,40	97,00	3,00	5,91	1,37	1,37	22,9
	5 050	65	73 000	14,50	43,20	95,60	4,40	10,31	1,22	1,31	14,0

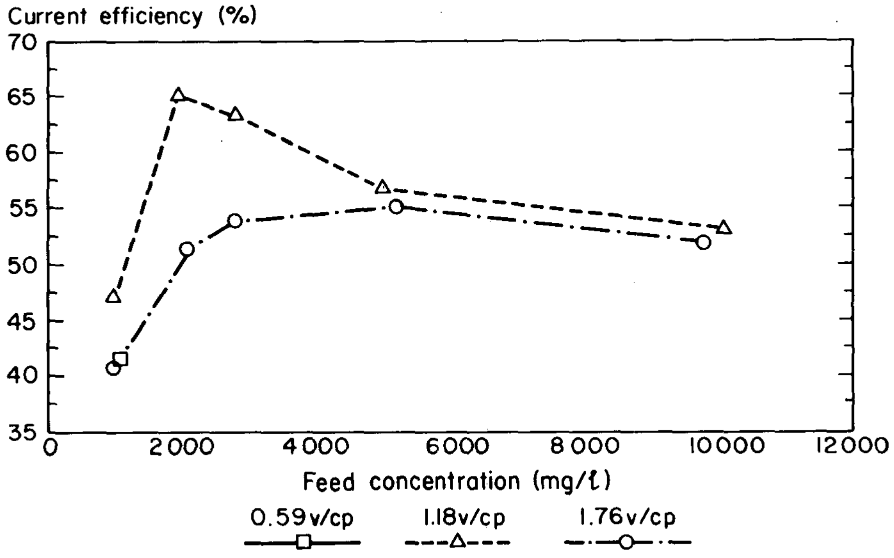

Figure 8
Current efficiency as a function of sodium nitrate feed concentration and cell pair voltage

saline solutions at cell pair voltages normally applied in ED.

All the different salt solutions could be easily desalinated from approximately 10 000 mg/ℓ to 300 mg/ℓ and less (Figs. 3 to 5 and Tables 1 to 5). Product concentrations of less than 100 mg/ℓ could be obtained with ease in some cases. Therefore, SCED appears to be effective for the production of low TDS water.

Brine concentration increased with increasing feed concentration and increasing cell pair voltage (Tables 1 to 5 and Fig. 7). Sodium chloride, ammonium nitrate, sodium sulphate, sodium nitrate and calcium chloride brine concentrations of 2,2 to 16,1%, 4,9 to 15%; 7,8 to 16,3%; 6,0 to 12,5% and 3,8 to 7,5% could be obtained respectively in the feed concentration and cell pair voltage range of 0,1 to 1% and 0,59 to 1,76 V/cp respectively. Therefore, relatively high brine concentrations could be obtained which would make the SCED technique suitable for concentration/desalination of industrial effluents. It is interesting to note that relatively low brine concentrations have been obtained with calcium chloride solutions (Table 5) in comparison with the other ions. However, the low current efficiency obtained with calcium chloride solutions explained the low brine concentrations that were obtained.

Concentration factors (brine/feed) decreased with increasing feed concentration (Tables 1 to 5). Therefore, there is a limit to the brine concentration that can be achieved. The brine concentration that can be obtained depends *inter alia* on the permselectivity of the ion-exchange membranes, feed concentration and current density used (Nishiwaki, 1972; Schoeman et al., 1990b). Ion-exchange membranes tend to loose their permselectivity at high concentration due to backdiffusion with the result that there is a limit to the brine concentration that can be achieved.

Current efficiency

Current efficiency increased with increasing feed concentration and decreasing cell pair voltage (Tables 1 to 5 and Fig. 8). Current efficiency, however, decreased slightly at higher feed concentrations due to the lower permselectivity of the ion-exchange membranes at high feed concentration. Increasing current efficiency with increasing feed concentration can be ascribed to a higher flow of electro-osmotic water through the membranes at increasing feed concentration (Kedem and Bar-on, 1986; Schoeman et al., 1990b). The higher flow of water into the membrane bags causes an increase in the flow of counter ions into the bags. This inhibits the penetration of co-ions into the membranes with the result that a high permselectivity is maintained.

Current efficiencies of 55 to 74%; 30 to 52%; 53 to 79%; 42 to 65%; and 37 to 50% were obtained with sodium chloride, ammonium nitrate, sodium sulphate, sodium nitrate and calcium chloride solutions respectively in the concentration and cell pair voltage ranges studied. Relatively low current efficiencies were obtained with ammonium nitrate and calcium chloride solutions. This shows that the ion-exchange membranes used do not have a very high permselectivity for ammonium nitrate and calcium chloride solutions.

Water recovery and brine volume

High water recovery and low brine volume were obtained at low to moderately high feed (1 000 to 3 000 mg/ℓ) concentrations (Tables 1 to 5). Brine volumes between 3 and 4%; 1 and 2%; 1 and 2%; 1 and 2%; and 1 and 3% were obtained with sodium chloride, ammonium nitrate, sodium sulphate, sodium nitrate and calcium chloride solutions respectively. Higher brine volumes (3 to 7%), however, were obtained at higher feed concentrations (5 000 to 10 000 mg/ℓ). Therefore, very low brine volumes could be obtained with SCED. This low brine volume that is produced with SCED can reduce brine disposal cost significantly especially where brine is to be trucked away for disposal.

Excellent water recoveries were obtained. Water recoveries of approximately 96% were obtained in the feed concentration range of 1 000 to 3 000 mg/ℓ and of approximately 94% in the feed concentration range from 5 000 to 10 000 mg/ℓ. These high water recoveries and low brine volumes are significantly better than water recoveries of approximately 80% which is normally obtained with conventional electrodialysis.

Electrical energy consumption

Electrical energy consumption increases with increasing feed concentration and cell pair voltage (Fig. 9 and Tables 1 to 5). Very low

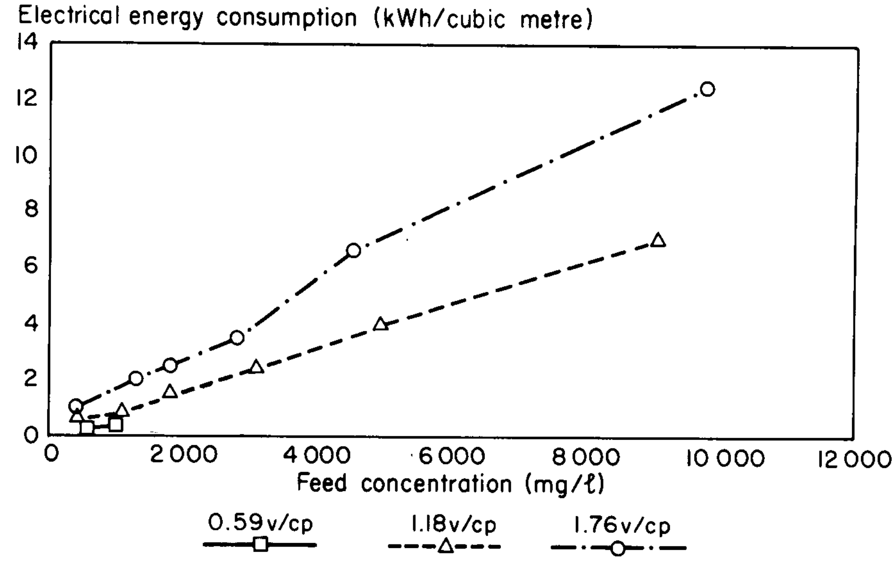

Figure 9
Electrical energy consumption as a function of ammonium nitrate feed concentration and cell pair voltage

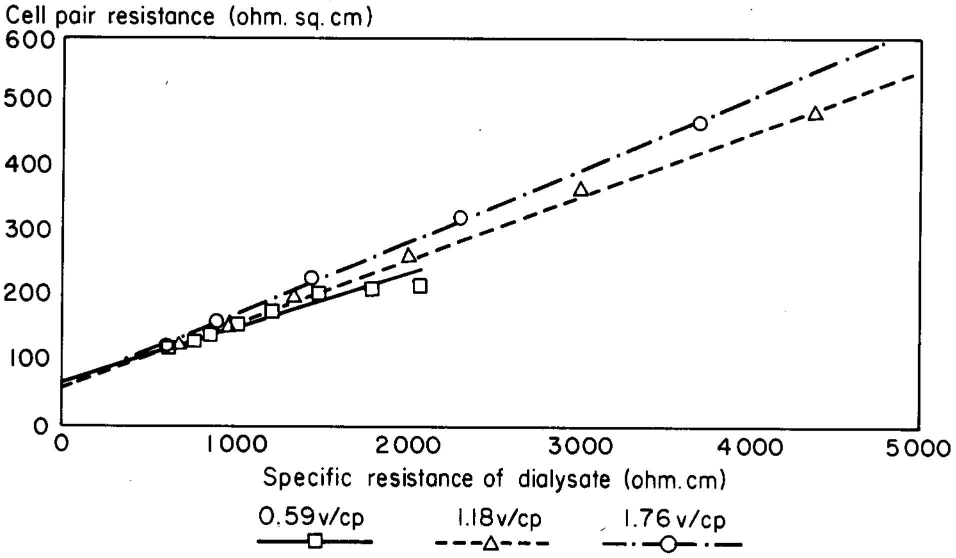

Figure 10
Cell pair resistance as a function of specific resistance of dialysate at different cell pair voltages. Feed: 1 000 mg/ℓ Na_2SO_4

electrical energy consumptions (0,27 to 0,48 kW·h/m³ product water) were obtained at a cell pair voltage of 0,59 in the 1 000 mg/ℓ feed concentration range. Electrical energy consumptions of 0,66 to 5,91 kW·h/m³ were obtained in the feed concentration range of 1 000 to 3 000 mg/ℓ (1,18 to 1,76 V/cp range). Higher electrical energy consumption (3,29 to 13,06 kW·h/m³) was encountered in the feed concentration range from 5 000 to 10 000 mg/ℓ.

Electrical energy consumption was determined for ion transport only. The voltage drop across the electrode compartments was not taken into consideration because it is usually insignificant in a large membrane stack containing many membrane pairs (300 membrane pairs or more). The electrical energy consumption obtained during SCED usage would give a good indication of the operational cost that could be expected with SCED applications.

Product water yield

Product water yield (output) increases with increasing cell pair voltage and decreases with increasing feed concentration (Tables 1 to 5). Product water yield is a very important engineering design parameter because the membrane area required for a certain flow rate can be calculated from this figure.

TABLE 6
CONCENTRATION/DESALINATION RESULTS OF FERTILISER RUNOFF AT DIFFERENT CELL PAIR VOLTAGES

V_{cp}	C_f mS/m	C_p mS/m	C_b mS/m	% Conductivity Removal	CE %	WR %	BV %	EEC kWh/m^3	OP m^3/m^2·d
1,18	545	29,8	10 724	94,5	56,9	97,2	2,8	2,7	1,03
0,88	556	48,9	10 312	91,2	63,3	97,2	2,8	2,0	0,77
0,59	520	53,3	8 830	89,7	-	96,9	3,1	1,24	0,54

TABLE 7
CHEMICAL COMPOSITION OF FEED, PRODUCT AND BRINE

Constituent	Feed mg/ℓ	Product mg/ℓ	Brine mg/ℓ	% Removal
Sodium	111	25	3 758	77,50
Potassium	34	5	1 035	85,30
Calcium	93	24	3 404	74,20
Magnesium	64	8	2 121	87,50
Ammonium	621	30	16 638	95,20
Nitrate	1 936	73	63 783	96,30
Silica	7,70	4,60	54,40	40,30
Sulphate	299	48	8 469	83,90
Ortho-phosphate (P)	73,80	20,80	1 143	71,80
Chloride	187	14	5 371	92,50
Alkalinity ($CaCO_3$)	22	3	24	86,40
COD	219	19	587	91,30
Manganese	0,409	<0,025	18,90	42,90
Iron	<0,025	<0,025	0,91	
Fluoride	1,66	0,35	3,70	78,90
TDS (calculated)	3 602	296	108 114	91,80
pH	5,7	4,3	4,4	

TABLE 8
CONCENTRATION/DESALINATION RESULTS OF AMMONIUM SULPHATE EFFLUENT

Vcp	Cf mS/m	Cp mS/m	Cb mS/m	% Conductivity Removal	CE %	WR %	BV %	EEC kWh/m^3	OP m^3/m^2·d
0,53	13 230	8 452	26 313	36,1	43,1	84,7	15,3	23,3	0,448
0,53	8 751	2 437	18 952	72,2	-	78,8	21,2	28,9	0,318
1,18	2 424	6,2	17 416	99,8	46,9	91,6	8,4	17,9	0,282

Cell pair resistance (R_{cp}) and effective thickness (d_{eff}) of the dialysate compartment

Cell pair resistance as a function of the specific resistance of the dialysate for sodium sulphate solutions at different cell pair voltages is shown in Fig. 10. Similar graphs were obtained for the other salt solutions. The lines consist of a linear region followed by a curved region (Schoeman and Enslin, 1989). The line starts to curve when the specific resistance of the dialysate becomes very high. Linear regression through the linear region of the lines gives d_{eff} (slope) and the cell pair resistance (R_{cp}) (y-intercept). The lines show that polarisation and hence effective thickness of the dialysate compartment depends on cell pair voltage. The effective thickness of the dialysate compartment increased from 0,84 (at 0,59 V/cp), 0,99 mm (at 1,18 V/cp) to 1,11 mm (at 1,76 V/cp). Membrane resistance (R_{cp}) for the sum of the anion- and cation-exchange membranes was determined at 65,6 - (0,59 V/cp); 57,2- (1,18 V/cp) and 59,0 ohm·cm^2 (at 1,76 V/cp). It was further found that R_{cp} decreased with increasing feed concentration (Tables 1 to 5). The cell pair resistance at 1,18 V/cp and an initial ammonium nitrate feed concentration of 9 100 mg/ℓ was determined at only 14,7 ohm·cm^2 (Table 2).

The model $R_{cp} = R_m + \rho d_{eff}$ is applicable not only to sodium chloride solutions but also to ammonium nitrate, sodium sulphate, sodium nitrate and calcium chloride solutions. However, care must be taken to use the linear portion of the curve (R_{cp} v. specific resistance) in the determination of R_{cp} and d_{eff}. This is also a method that can be used for the determination of cell pair resistance. Cell pair resistance, however, depends on the initial feed concentration. Therefore, feed concentration must be specified when cell pair resistance is given.

Treatment of runoff from a fertiliser factory terrain with SCED

Runoff from an ammonium nitrate fertiliser factory terrain is presently stored in evaporation ponds. This runoff contains, amongst other ions, ammonium, nitrate and phosphate ions which have the potential to pollute the environment. Water and chemicals can also be recovered from the effluent for reuse. Sealed-cell ED was therefore investigated for treatment of this effluent (Schoeman and Enslin, 1989).

The concentration/desalination results of the relatively dilute runoff are shown in Table 6.

Excellent conductivity removals were obtained at the three cell pair voltages investigated. Conductivity removal of 94,5% was obtained at a cell pair voltage of 1,18. Conductivity removal decreased to only 89,7% at 0,59 V/cp.

Feed water conductivity was reduced from 545 mS/m to 29,8 mS/m at an electrical energy consumption of 2,7 kW·h/m^3 (1,18 V/cp). Brine volume comprised only 2,8% of the initial feed volume. Effluent volume could therefore be reduced significantly. The chemical composition of feed, product and brine is shown in Table 7.

TDS was reduced from 3 602 mg/ℓ to 296 mg/ℓ (1,18 V/cp) with ease. Therefore, a very good quality product water could be produced which might be reused at the factory. Very good ammonium (95,2%) and nitrate (96,3%) removals were obtained. Ammonium and nitrate were reduced from 621 and 1 936 mg/ℓ in the feed to 30 and 73 mg/ℓ in the product respectively.

The brine had a TDS of 10,8%. Brine volume comprised only about 3% of the initial feed volume. Therefore, brine volume could be reduced significantly which means that smaller evaporation ponds would be required, or that the present ponds could last much longer. Ammonium and nitrate values may also be recovered from the brine for reuse. Potential pollution problems will therefore be reduced significantly.

The ion-exchange membranes used in the SCED unit performed well for treatment of the fertiliser runoff. However, membrane fouling or scaling in the long term may affect the process adversely. Therefore, membrane fouling and cleaning studies over an extended time period will be necessary to determine the effectiveness of SCED for this application.

Treatment of a concentrated ammonium nitrate type effluent with SCED

The treatment of a more concentrated ammonium nitrate type effluent from a fertiliser manufacturing plant was also investigated with SCED. The pH of the effluent was approximately 11 and the effluent was neutralised with sulphuric acid prior to SCED treatment (Schoeman and Enslin, 1989). Concentration/desalination of the ammonium sulphate effluent was conducted in stages because of the high concentration of the effluent (13 230 mS/m or 123 700 mg/ℓ). The product water after the first desalination stage was used as feed for the next concentration/desalination stage. The concentration/desalination results are shown in Table 8.

Feed (13 230 mS/m) was first desalinated to 8 452 mS/m. Desalination rate was low due to the low cell pair voltage (0,53 V/cp) that could be applied as a result of excessive current that was drawn by the high conductivity of the feed solution (Schoeman and Enslin, 1989). It was only at the third desalination stage that a higher cell pair voltage could be applied.

The chemical composition of the feed, product and brine after the third desalination/concentration stage is shown in Table 9.

Very good ion removals were obtained. TDS was reduced from 16 557 mg/ℓ to 88 mg/ℓ, a 99,5% removal. Ammonium and nitrate removals were both approximately 99%. Brine with a TDS of 19,1% was obtained. Brine volume comprised 8,4% of the initial feed volume. Electrical energy consumption was determined at 17,9 kW·h/m^3 in this case. This energy consumption is high. However, an excellent quality product water was obtained which could be reused. This demonstrates that SCED should be effective for the treatment of relatively high TDS waters although the electrical energy consumption is high.

Treatment of an effluent saturated with calcium sulphate with SCED

Hydrochloric acid is used for extraction of calcium from activated carbon which is used for gold extraction by a gold recovery company. The effluent contains at times high concentrations of calcium (3 800 mg/ℓ), chloride (7 000 mg/ℓ) and sulphate (600 mg/ℓ). Sealed-cell ED was attempted for treatment of this high concentration calcium sulphate effluent (TDS 23 000 mg/ℓ) for chloride recovery (Schoeman and Enslin, 1989). However, a white precipitate of calcium sulphate formed in the membrane bags shortly after the experiment was started. Therefore, calcium sulphate should be reduced to low levels to prevent calcium sulphate scaling during SCED treatment. This was done by treating another effluent sample (TDS 4 500 mg/ℓ) with barium carbonate. Sulphate was reduced from 339 mg/ℓ to 5 mg/ℓ. The concentration/desalination results are summarised in Table 10.

Concentration/desalination was conducted in two stages. Conductivity was first reduced from 1 182 mS/m to 362 mS/m and then from 362 mS/m to 51 mS/m. Very low current efficiencies were obtained for the first (32,5%) and second (28,8%) desalination stages. Brine volume comprised approximately 3% (1st stage)

TABLE 9
CHEMICAL COMPOSITION OF FEED, PRODUCT AND BRINE (3RD STAGE DESALINATION)

Constituent	Feed mg/ℓ	Product mg/ℓ	Brine mg/ℓ	% Removal
Sodium	268	12	2 787	95,52
Potassium	3	1	17	66,67
Calcium	7	1	60	85,71
Magnesium	1	4	13	
Kjeldahl-N	3 340	17	38 199	99,49
Ammonium	4 179	10	48 214	99,76
Nitrate	2 215	17	25 473	99,23
Silica	9,50	3,90	40,10	58,95
Sulphate	9 762	10	113 184	99,90
Total phosphate (P)	3,20	0,20	28,20	93,75
Chloride	103	28	1 167	72,82
COD	41	19	163	53,66
TDS (Calculated)	16 557	88	191 208	99,47
pH	3,6	4,9	2,9	

TABLE 10
CONCENTRATION/DESALINATION RESULTS OF CALCIUM CHLORIDE EFFLUENT

V_{cp}	C_f mS/m	C_p mS/m	C_b mS/m	% Conductivity Rejection	CE %	WR %	BV %	EEC kWh/m^3
1,18	1 182	362	13 548	69,4	32,5	97,0	3	6,4
1,18	383	51	9 609	86,7	28,8	97,7	2,3	3,1

TABLE 11
CHEMICAL COMPOSITION OF FEED, PRODUCT AND BRINE (2ND STAGE DESALINATION)

Constituent	Feed mg/ℓ	Product mg/ℓ	Brine mg/ℓ	% Removal
pH	8,1	8,1	6,7	
Conductivity (mS/m)	383	51	9 609	86,7
Sodium	191	77	4 862	59,7
Potassium	9	3	162	66,7
Calcium	278	10	17 045	96,4
Magnesium	5	4	7	20,0
Ammonium	27	7	447	274,1
Nitrate	4	2	241	50,0
Sulphate	3	4	3	-
Chloride	783	113	46 412	85,6
Alkalinity ($CaCO_3$)	139	65	338	53,2
TDS (calculated)	1 469	299	102 180	79,6

and 2,3% (2nd stage) of the feed water volume and electrical energy consumption was determined at 6,4 and 3,1 kW·h/m^3 for the first and second desalination stages respectively.

The chemical composition of the feed, product and brine for the second desalination stage is shown in Table 11.

A very good quality product water was obtained after the second desalination stage. TDS was reduced from 1 469 mg/ℓ to 299 mg/ℓ at an electrical energy consumption of 3,1 kW·h/m^3.

Chloride was effectively concentrated. The chloride concentration in the brine was 4,6%. This chloride may be converted into hydrochloric acid in an electrochemical cell. The recovered hydrochloric acid can then be used for removal of calcium from the spent activated carbon. This matter, however, warrants further investigation.

The high calcium concentration in the brine may cause scaling problems. However, no sign of scaling was detected during the laboratory tests. Membrane fouling and cleaning tests, however, should be conducted over an extended period of time to determine the practical feasibility of the process.

Advantages and disadvantages of SCED

The capital cost of SCED equipment should be less than that of a conventional plate-and-frame ED stack because of the simpler construction of the SCED stack. The membrane utilisation factor in the membrane bags is 95% compared to approximately 80% for membranes in conventional ED stacks. Higher current densities can be used in SCED than with conventional sheet flow ED because higher linear flow velocities can be obtained. The higher current densities will result in higher water production rates. Brine volumes produced by SCED are smaller than obtained with conventional ED. Therefore, the brine disposal problem will be reduced.

More electrical energy per unit of product water produced is used in the SCED stack due to the higher current densities used. However, the increased cost for electrical energy should be offset by a decrease in capital cost. Scale may form more readily in the membrane bags because the SCED stack does not have a built-in self cleaning device such as encountered in the electrodialysis reversal system (USAID Desalination Manual, 1980). It will be difficult to remove scale from the membrane bags once it has formed because the bags cannot be opened. Therefore, scale-forming chemicals should be removed by ion-exchange or nanofiltration prior to SCED treatment. This will affect the economics of the process adversely especially if large volumetric flows are involved.

Scale-up of a laboratory size SCED unit (100 cm^2/cp) to a pilot or full-scale plant should be possible. It should be possible to manufacture large-scale membrane bags commercially and the bags would be robust. An advantage of the membranes is that they can be stored dry. This is usually not the case with ion-exchange membranes normally used in conventional ED. The successful application of this technology will depend on the need to apply this technology in preference to conventional ED for specific applications where high brine concentrations and small brine volumes are required.

Conclusions

- Sealed-cell ED could be effectively applied for desalination/concentration of relatively dilute (500 to 3 000 mg/ℓ TDS) non-scaling salt solutions. Product water with a TDS of less than 300 mg/ℓ could be produced in the feed concentration range from 500 to 10 000 mg/ℓ TDS. Electrical energy consumption of 0,27 to 5,9 kW·h/m^3 was obtained (500 to 3 000 mg/ℓ feed range). Brine volume comprised approximately 2% of the initial feed volume. Therefore, brine disposal costs should be significantly reduced with this technology.
- Sealed-cell ED became less efficient in the 5 000 to 10 000 mg/ℓ TDS feed concentration range due to high electrical energy consumption (3,29 to 13,0 kW·h/m^3). However, SCED could be applied in this TDS range depending on the value of the products that would be recovered.
- Treatment of scale-forming waters will affect the process adversely because scale will precipitate in the membrane bags which cannot be opened for cleaning. Membrane scaling should be removed by current reversal or with cleaning solutions. However, this matter needs investigation. Scale-forming waters, however, should be avoided or treated with ion exchange or nanofiltration prior to SCED.
- It was demonstrated that a relatively dilute ammonium nitrate effluent (TDS 3 600 mg/ℓ) could be successfully treated in the laboratory with SCED. Brine volume comprised only 2,8% of the treated water volume. Electrical energy consumption was determined at 2,7 kW·h/m^3. Both the brine and the treated water should be reused. Membrane fouling or scaling, however, may affect the process adversely and this matter needs further investigation.
- It was difficult to desalinate/concentrate a concentrated ammonium sulphate effluent (approximately 13 200 mS/m or 123 700 mg/ℓ TDS) with SCED. Desalination/concentration was conducted in stages. Nevertheless, it was possible to desalinate the effluent to 6,2 mS/m (88 mg/ℓ TDS). However, electical energy consumption was high (59 kW·h/m^3). Brine volume comprised 45% of the treated volume. A very high brine concentration (approximately 26 300 mS/m or 332 300 mg/ℓ TDS) could be obtained after the first desalination stage. However, a more dilute (16 557 mg/ℓ TDS) ammonium sulphate effluent (3rd stage) could be more easily desalinated/concentrated to 88 and 191 208 mg/ℓ TDS product water and brine respectively at a water recovery and electrical energy consumption of 91,6% and 17,9 kW·h/m^3 respectively. Therefore, SCED could also be effectively applied for the desalination/concentration of relatively high TDS waters.
- It was not possible to desalinate/concentrate an effluent saturated with calcium sulphate with SCED due to membrane scaling which took place. However, it was possible to concentrate/desalinate the effluent effectively after sulphate removal by chemical precipitation. It was possible to desalinate/concentrate the effluent from 1 182 mS/m (4 461 mg/ℓ TDS) to 51 mS/m (299 mg/ℓ TDS) at an energy consumption of 9,5 kW·h/m^3. Brine volume comprised only 5,3% of the treated feed. The procedures need to be evaluated cost-wise.
- The ion-exchange membranes used in the SCED stack performed very well for ammonium and nitrate removal. Ammonium and nitrate ions were removed from 4 179 and 2 215 mg/ℓ in one case to 10 mg/ℓ (99,8% removal) and 17 mg/ℓ (99,2% removal) respectively.
- Capital cost of SCED equipment should be less than that of conventional ED due to the simpler design of the SCED stack. The membrane utilisation factor of 95% is much higher than in conventional ED (80%).
- Sealed-cell ED has potential for treatment of relatively dilute (< 3 000 mg/ℓ TDS) non-scaling waters for water and chemical recovery for reuse. However, high TDS (up to 16 000 mg/ℓ waters) could also be treated depending on the value of the products that can be recovered.

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