

# The status of electrodialysis technology for brackish and industrial water treatment

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## Abstract

The electrodialysis (ED) process and some of its applications are reviewed. Fouling of ED membranes may be a serious problem unless the necessary precautions are taken. Seawater desalination by ED is limited at present, owing to high cost. The electrodialysis reversal (EDR) process is used mainly for the desalination of brackish waters. Waters saturated with calcium sulphate have been treated successfully. The treatment of brackish waters with a combination of EDR and ion exchange (IX) for industrial purposes appears to be more economical than IX alone at TDS levels of approximately 300 mg/l and higher. Treatment of nickel-plating washwaters for water and nickel recovery is an established industrial process in Japan. The treatment of cooling tower blowdown for water recovery and to minimize brine disposal appear attractive. The concentration of dilute chemical effluents for chemical and water recovery, the recovery of acids and bases from IX regeneration wastes, and the concentration of sodium sulphate and its subsequent conversion into caustic soda and sulphuric acid, have been studied and their technical feasibility demonstrated.

## Introduction

Electrodialysis (ED) technology has progressed significantly during the past 35 years since the introduction of synthetic ion exchange membranes in 1949 (Leitz, 1976). The first two decades of this period saw the development of classical or unidirectional standard electrodialysis. However, during the past decade, the main feature has been the development of the polarity reversal process – the so-called electrodialysis reversal (EDR) (Katz, 1977). This form of electrodialysis desalination has virtually displaced unidirectional ED for most brackish water applications and is slowly gaining a significant share of this market (Katz, 1977).

EDR is at present mainly used for the desalination of brackish waters to produce fresh potable and industrial water. Unidirectional ED is used on a large scale in Japan for concentrating seawater to produce brine for salt production (Leitz and Eisenmann, 1981) and is also used on a small scale for seawater desalination (Tani, 1981) and for brackish water desalination (Urano, 1977).

Outside the water desalination field, ED is also being used on a large and increasing scale in North America and Europe to de-ash cheese whey to produce a nutritious high quality protein food supplement (Leitz, 1976). It is also finding application in the treatment of industrial waste waters for water recovery, reuse and effluent volume reduction (Wirth and Westbrook, 1977; Itoi, 1979; Melzer and van Deventer, 1983).

The purpose of this survey is to –

- define the standard ED and EDR processes;
- consider fouling and pretreatment;
- review standard ED and EDR in the desalination market;
- give a brief review of seawater desalination by standard ED;
- consider ED for desalination of scaling waters; and
- consider ED for desalination of industrial waste waters with emphasis on water reuse, product recovery, effluent volume reduction and future prospects.

## Principles of operation

### The standard ED process

In the ED process, water flows between alternately placed cation- and anion-permeable membranes (Fig. 1), which are built into a so-called electrodialysis stack. Direct current (DC) provides the

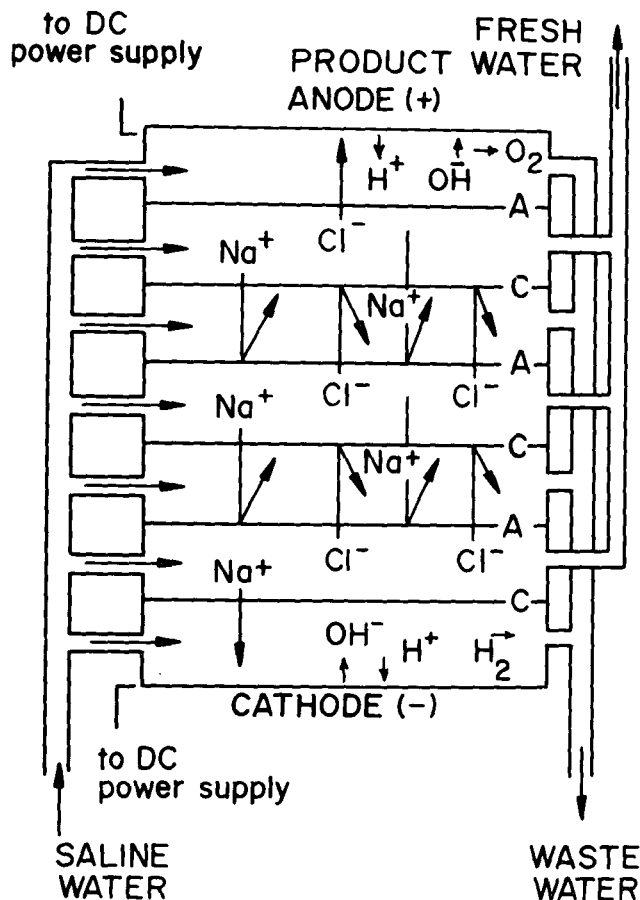


Figure 1  
A = anion-permeable membrane C = cation-permeable membrane  
The ED process.

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motive force for ion migration through the membranes and the ions are removed or concentrated in the alternate water passages by means of permselective membranes.

The standard ED process often requires the addition of acid and/or polyphosphate to the brine stream to inhibit the precipitation of sparsely soluble salts (such as  $\text{CaCO}_3$  and  $\text{CaSO}_4$ ) in the stacks. To maintain performance, the membrane stacks need to be cleaned periodically to remove scale and other surface fouling matter. This is done in two ways (*USAID Desalination Manual*, 1980) by cleaning-in-place (CIP); and stack disassembly.

Special cleaning solutions (dilute acids or alkaline brine) are circulated through the membrane stacks for in-place cleaning, but at regular intervals the stacks need to be disassembled and mechanically cleaned to remove scale and other surface-fouling matter. Regular stack disassembly is a time-consuming operation and is a disadvantage of the standard ED process.

### The EDR process

The EDR process operates on the same basic principles as the standard ED process. In the EDR process, the polarity of the electrodes is automatically reversed periodically (about three to four times per hour) and, by means of motor operated valves, the 'fresh product water' and 'waste water' outlets from the membrane stack are interchanged. The ions are thus transferred in opposite directions across the membranes. This aids in breaking up and flushing out scale, slime and other deposits from the cells. The product water emerging from the previous brine cells is usually discharged to waste for a period of one to two minutes until the desired water quality is restored.

The automatic cleaning action of the EDR process usually eliminates the need to dose acid and/or polyphosphate, and scale formation in the electrode compartments is minimized due to the continuous change from basic to acidic conditions. Essentially, therefore, three methods of removing scale and other surface-fouling matter are used in the EDR process (*USAID Desalination Manual*, 1980) viz. cleaning-in-place; stack disassembly as used in standard ED; and reversal of flow and polarity in the stacks. The polarity reversal system greatly extends the intervals between the rather time-consuming task of stack disassembly and reassembly, with an overall reduction in maintenance time.

The capability of EDR to control scale precipitation more effectively than standard ED is a major advantage of this process, especially for applications requiring high water recoveries. However, the more complicated operation and maintenance requirements of EDR equipment necessitate more labour and a greater skill level than reverse osmosis (RO) and may be a disadvantage of the process (Fraivillig, 1983).

### Fouling

Fouling of the ED membranes by dissolved organic and inorganic compounds may be a serious problem in practical electro dialysis (Korngold *et al.*, 1970; Lacey and Loeb, 1972; Van Duin, 1973) unless the necessary precautions (pretreatment) are taken. Organic fouling is caused by the precipitation of large negatively charged anions on the anion-permeable membranes in the dialysate compartments while inorganic fouling is caused by the precipitation (scaling) of slightly soluble inorganic compounds (such as  $\text{CaSO}_4$  and  $\text{CaCO}_3$ ) in the brine compartments and the fixation of multivalent cations (such as Fe and Mn) on the cation-

permeable membranes. Organic anions or multivalent cations can neutralize or even reverse the fixed charge of the membranes, with a significant reduction in efficiency. Fouling also causes an increase in membrane stack resistance which, in turn, increases electrical consumption and adversely affects the economics of the process.

The following constituents are, to a greater or lesser extent, responsible for membrane fouling (Katz, 1982):

- Traces of heavy metals such as Fe, Mn and Cu.
- Dissolved gases such as  $\text{O}_2$ ,  $\text{CO}_2$  and  $\text{H}_2\text{S}$ .
- Silica in diverse polymeric and chemical forms.
- Organic and inorganic colloids.
- Fine particulates of a wide range of sizes and composition.
- Alkaline earths such as Ca, Ba and Sr.
- Dissolved organic materials of both natural and man-made origin in a wide variety of molecular weights and compositions (Korngold, 1976).
- Biological materials – viruses, fungi, algae, bacteria – all in varying stages of reproduction and life cycles.

Many of these foulants may be controlled by pretreatment steps which usually stabilize the ED process. However, according to Katz (1982), the development of the EDR process has helped to solve the pretreatment problem more readily in that it provides self-cleaning of the vital membrane surfaces as an integral part of the desalting process.

### Pretreatment

Pretreatment techniques for ED are similar to those used for RO (*USAID Desalination Manual*, 1980). Suspended solids are removed by sand and cartridge filters ahead of the membranes. Suspended solids, however, must be reduced to a much lower level for RO than for ED. The precipitation of slightly soluble salts in the standard ED process may be minimized by ion exchange softening and/or reducing the pH of the brine through acid addition and/or the addition of an inhibiting agent.

Organics are removed by carbon filters, and hydrogen sulphide by oxidation and filtration. Biological growths are prevented by a chlorination-dechlorination step. The dechlorination step is necessary to protect the membranes from oxidation. Iron and manganese are removed by green sand filters, aeration, or other standard water treatment methods. It has been suggested that multivalent metal and organic ions, and hydrogen sulphide, however, must be reduced to a lower level for EDR than for RO (Fraivillig, 1983).

The overall requirements for pretreatment in ED, may be somewhat less rigorous than for RO due to the nature of the salt separation and the larger passages provided (*USAID Desalination Manual*, 1980). In ED, the ions (impurities) move through the membranes, while in RO the water moves under a high pressure through the membranes while the salts are rejected. Salts with a low solubility can, therefore, more readily precipitate on spiral and hollow fine fibre RO membranes to cause fouling and to block the small water passages. Suspended solids can also more readily form a deposit. However, this might not be the case with tubular RO membranes. With the EDR process, precipitated salts in the brine compartments can be more readily dissolved and flushed out of the system using polarity reversal without the need for chemical pretreatment.

However, high removals of suspended solids, iron, manganese, organics and hydrogen sulphide are still critical to

avoid fouling and suppliers of EDR equipment recommend pretreatment of the feed water (*USAID Desalination Manual*, 1980) if it contains the following ions: Fe > 0,3 mg/l; Mn > 0,1 mg/l; H<sub>2</sub>S > 0,3 mg/l; free chlorine and turbidity > 2 NTU. In every case, of course, a careful examination of the prospective water would be necessary to determine suitability and pretreatment.

A certain degree of fouling is, however, unavoidable. Membranes should, therefore, be regularly washed with dilute acid and alkali solutions to restore performance.

### Post-treatment

The EDR product water is usually less aggressive than the RO product because acid is usually not added in EDR for scale control (Fraivillig, 1983). Post-pH adjustment may, therefore, not be required as with RO. Non-ionic matter in the feed such as silica, particulates, bacteria, viruses, pyrogens and organics will not be removed by the ED process and must, if necessary, be dealt with during post-treatment.

### ED and EDR in the desalination market

Figure 2 shows the increase in water production capacity throughout the world of water desalination installations of more than 94,6 m<sup>3</sup>/d capacity for the years 1968 to mid-1980 (Katz, 1982).

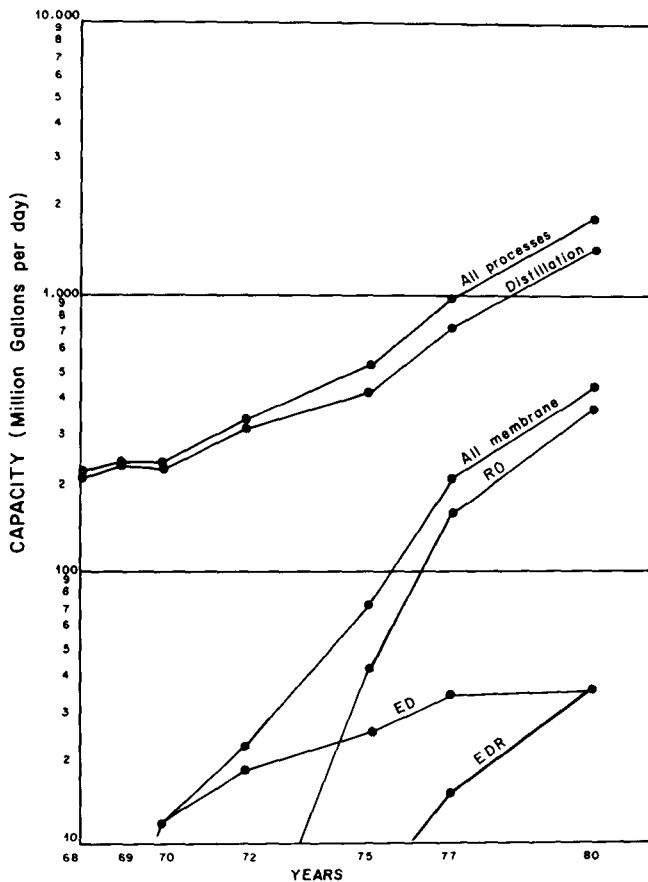


Figure 2  
Capacity (million gallons per day) of water desalting installations over 25 000 gpd (94,6 m<sup>3</sup>/d).

The aggregate water production capacity of all types of desalting plants over this 13 year period has increased almost ten-fold – from a little over 757 000 m<sup>3</sup>/d in 1968 to just under 7 570 000 m<sup>3</sup>/d in mid-1980. The membrane processes have grown from almost nothing in 1968 to account for about 25% of the capacity in mid-1980. Standard ED and EDR combined accounted for almost 15% of the membrane capacity in mid-1980. Standard ED entered the field prior to 1970 – followed by diminished growth and virtual obsolescence as a desalting process by the mid-1970's. The reasons for this were process shortcomings and the introduction of the EDR process in about 1973. The process reached a capacity of 37 850 m<sup>3</sup>/d in 1976, doubling in output about every three years. Recent data on EDR growth (Katz, 1982) show that the indicated growth rates were maintained in 1981 and EDR capacity was expected to be well over 189 250 m<sup>3</sup>/d by the end of 1982. The most recent data on EDR growth indicated that the EDR capacity was expected to be about 272 520 m<sup>3</sup>/d by the end of 1983 (Katz, 1983).

### Seawater desalination

There is limited application of ED for seawater desalination because of high costs (*USAID Desalination Manual*, 1980). A small batch system (120 m<sup>3</sup>/d) has been in operation in Japan to produce water of potable quality at a power consumption of 16,2 kWh/m<sup>3</sup> product water (Miva, 1977). A 200 m<sup>3</sup>/d seawater EDR unit was recently evaluated in China (Shi and Chen, 1983). This unit operated at 31°C; its performance was stable; total electric power consumption was 18,1 kWh/m<sup>3</sup> product water and the product water quality of 500 mg/l TDS met all the requirements for potable water. When the stacks were disassembled for inspection, there were no signs of scale formation.

With the commercial ED units currently available, the energy usage for seawater desalination is relatively high compared with that of RO. However, fairly recent work under the Office of Water Research and Technology (OWRT) programmes has indicated that high-temperature ED may possibly be competitive with RO (Parsi, Prato and O'Donoghue, 1980). Results have shown that the power consumption can be reduced to the levels required for seawater RO (8 kWh/m<sup>3</sup>) and that a 50% water recovery can probably be attained.

### Brackish water desalination for drinking-water purposes

A considerable number of standard ED plants for the production of potable water from brackish water are in operation (Urano, 1977; Kusakari *et al.* 1977; *USAID Desalination Manual*, 1980). These plants are operating successfully. However, after the introduction of the reversal process in the early 1970's, Ionics Incorporated shifted almost all their production to this process (Katz, 1982).

The major application of the EDR process is for the desalination of brackish water. The power consumption and, to some degree, the cost of equipment required is directly proportional to the TDS to be removed from the feed water (*USAID Desalination Manual*, 1980). Thus, as the feedwater TDS increases, the desalination costs also increase. In the case of the RO process, a cost: TDS removal relationship also exists, but it is not as pronounced. Often the variation in the scaling potential of the feed

water and its effect on the percentage of product water recovery can be more important than the cost: TDS relationship.

Thus, for applications requiring low TDS removals, ED is often the most energy-efficient method, whereas with highly saline feed waters RO may be expected to use less energy and is preferred. The economic crossover point between ED and RO based on operating costs is, however, difficult to define precisely and needs to be determined on a site-specific basis. Apart from local power costs, other factors must also be considered in determining the overall economics. Among these, to the advantage of ED, are the high recoveries possible (up to 90%), the elimination of chemical dosing (with EDR), and the reliability of performance that is characteristic of the ED process.

### Energy consumption

The energy consumption of a typical EDR plant is as follows (USAID Desalination Manual, 1980):

|                |  |
|----------------|--|
| Pump           | : 0,5 to 1,1 kWh/m <sup>3</sup> product water                  |
| Membrane stack | : 0,7 kWh/m <sup>3</sup> product water/1 000 mg of TDS removed |
| Power losses   | : 5% of total energy usage                                     |

The major energy requirement, therefore, is for pumping the water through the ED unit and for the transport of the ions through the membranes.

### Treatment of a high scaling, high TDS water with EDR

The successful performance of EDR on high calcium sulphate waters has been reported (Katz, 1977). Brown (1981) has described the performance of an EDR plant treating 300 m<sup>3</sup>/d of a high calcium sulphate water with a TDS of 9 700 mg/l. The only pretreatment applied was iron removal on green sand. The quality of the feed, product and brine is shown in Table 1.

The water recovery and energy consumption were 40% and 7,7 kWh/m<sup>3</sup> of product water, respectively. No attempt was made to optimize water recovery. The stack resistance increased by only 3% after one year of operation, which clearly indicates the successful operation of the EDR unit in spite of the supersaturated condition of the brine with respect to calcium sulphate. Membrane lifetimes are estimated to be 10 years.

### Recent developments

The main developments in EDR during the past few years have been the following:

- EDR has achieved CaSO<sub>4</sub> saturation in the brine stream of up to 440% without performance decline on tests of several hundred hours duration (Elyanow, Sieveka and Mahoney, 1981). Longer term tests are in progress.
- EDR has desalted a hard (Ca<sup>2+</sup> approx. 150 mg/l) brackish water of 4 000 mg/l TDS at water recoveries of up to 93% without cumbersome and expensive pre-softening (Elyanow, Parent and Mahoney, 1980).
- An EDR test unit has achieved 95% or greater recovery of a limited 4 000 mg/l TDS brackish water resource by substituting a more abundant 14 000 mg/l saline water in the brine stream (Elyanow, Parent and Mahoney, 1980). The substitution of sea water in the brine stream would be freely

available in coastal or island locations with limited high quality brackish water resources.

- The development, extensive field testing and subsequent large-scale commercial usage of a new family of thick (0,5 mm), rugged anti-fouling anion-permeable membranes in the USA with much higher current efficiencies and chlorine resistance than those formerly available (Elyanow, Parent and Mahoney, 1980).

TABLE 1  
WATER QUALITY BEFORE AND AFTER EDR TREATMENT

| Constituent                   | Feed (mg/l) | Product (mg/l) | Brine (mg/l) |
|-------------------------------|-------------|----------------|--------------|
| Na <sup>+</sup>               | 2 090       | 79             | 3 694        |
| Ca <sup>++</sup>              | 652         | 4              | 1 390        |
| Mg <sup>++</sup>              | 464         | 4              | 964          |
| Cl <sup>-</sup>               | 3 687       | 111            | 7 084        |
| HCO <sub>3</sub> <sup>-</sup> | 134         | 25             | 175          |
| SO <sub>4</sub> <sup>-</sup>  | 2 672       | 19             | 5 000        |
| TDS                           | 9 727       | 242            | 18 307       |
| pH                            | 7,0         | 6,8            | 7,2          |

### Brackish water desalination for industrial purposes

In the past most ED plants treated brackish waters of 1 000 to 10 000 mg/l TDS and produced general purpose industrial product water of 200 to 500 mg/l TDS. However, ED capital and construction costs have declined during recent years to the point where it is already feasible to treat water containing 200 to 1 000 mg/l TDS and produce product water containing as little as 3 to 5 mg/l TDS (Katz, 1971). These low TDS levels are achieved by multistaging. The systems, which often employ ion exchange (IX) units as 'polishers', are usually referred to as ED/IX systems.

### ED/IX system

New and existing ion exchange facilities can be converted to ED/IX systems by addition of ED units upstream of the ion exchange units. The ED unit reduces chemical consumption, waste, service interruptions and resin replacement of the ion exchanger in proportion to the degree of prior mineral removal achieved (Katz, 1971). For small capacity systems (2 to 200 m<sup>3</sup>/d) the optimum ED demineralization will usually be 90% or greater; for larger installations, and particularly those where adequate ion exchange capacity is already provided, the optimum demineralization via ED is more likely to be in the 60 to 80% range.

It must, however, be stressed that RO may also be used for the abovementioned application. RO may function better than ED because it removes silica and organic material better than ED. However, the choice of the treatment method (ED or RO) would be determined by the specific requirements and costs for a specific situation.

Honeywell in the USA, which manufactures printed circuit boards and does zinc plating and anodizing, used IX for the treatment of their process waters before they changed over to an ED/IX system (Highfield, 1980). ED was chosen instead of RO because of lower membrane replacement costs. Process waters of varying degrees of purity are required, dissolved solids being the primary concern. Water with a TDS of about 50 mg/l is suitable for zinc plating and anodizing and water with a TDS with a

minimum specific resistance of 100 000 ohms is satisfactory for circuit board fabrication operations (Highfield, 1980). The purity of the treated water (raw water TDS – 350 to 500 mg/ℓ) after treatment with the ED/IX system was better than expected. Service runs have been up to ten times longer than before.

### Costs

A comprehensive cost study was carried out to compare the costs of owning and operating the old IX and new ED/IX systems (Highfield, 1980). The cost of removing the TDS from the water, the prime purpose of the process water plant, was shown to be reduced by about 85% (US \$1,94 vs. \$0,30 per 3,785 m<sup>3</sup>) for the ED/IX system. The total operating cost (water cost, electrical power, chemicals, labour) was reduced from 2,25 to US \$0,65 per 3,785 m<sup>3</sup>, a reduction of slightly more than 70%.

The economic crossover point between an ED/IX system and an IX system only is not stated for the above application. However, according to Jones (1984) this economic crossover point depends on the specific circumstances (capital structure, etc) for each situation. The ED/IX system becomes more economical than the IX system alone at a TDS of approximately 300 mg/ℓ and higher.

### Future prospects

The decreased requirements for regeneration of the deionizers has reduced resin deterioration and loss and has significantly reduced maintenance problems and associated downtime. Less waste for disposal was also produced and from a pollution abatement point of view it is foreseen that ED/IX will be more extensively used in future.

### Industrial wastewater desalination for water reuse, chemical recovery and effluent volume reduction

Large volumes of water containing varying amounts of salt, which are generated by washing and regenerating processes, blowdown from cooling towers, disposal of dilute chemical effluents, to name a few, present significant problems, particularly when zero effluent discharge is required. The problem is one of too much water carrying comparatively little salt, but still having a TDS content too great for acceptance to a receiving stream. Many industries face this problem today and have to consider the application of processes for concentrating salts or desalting water. The ED system for water recovery and brine concentration may be one of those best suited to alleviate the problem.

Some typical examples are given to illustrate this principle.

### Electrodialysis of nickel plating solutions

During many plating operations, a substantial amount of bath solution adheres to plated work pieces as they leave the plating tank. In this manner valuable materials are lost as 'drag-out' into the subsequent rinse tank. This contaminated rinse solution can be passed through an ED system where these valuable materials can be recovered and returned to the plating tank.

One such opportunity of significant industrial importance is provided by nickel electroplating operations (Itoi, 1979). Earlier work by Trivedi and Prober (1972) demonstrated the successful application of ED to nickel solutions. Later, Eisenmann (1977) and Itoi (1979) reported the use of ED to recover nickel from electroplating rinse waters.

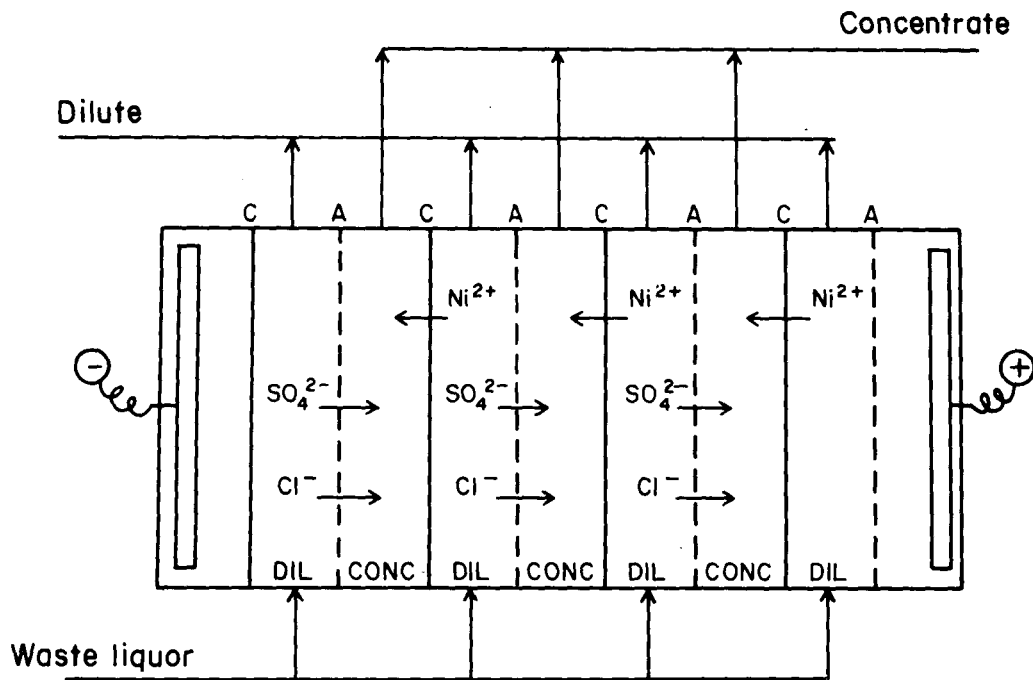


Figure 3  
Electrodialysis of the washwater from a nickel galvanizing operation.

### Principle of operation

The wash water from a nickel galvanizing line is treated by ED as shown in Figure 3 (Itoi, 1979).

### Results

The results achieved in an existing facility are given in Table 2.

The concentration ratio of the concentrated solution to the dilute solution is greater than 100. The concentrated solution is reused in the plating bath while the dilute solution is reused as wash water. The recovery of nickel discharged from the wash tank is approximately 90% or greater.

If organic electrolytes are present in the additives used in the galvanization bath, they must be removed prior to ED treatment to prevent organic fouling of the ED membranes.

| Constituent       | Effluent (g/l) | Concentrate (g/l) | Diluant (g/l) |
|-------------------|----------------|-------------------|---------------|
| NiSO <sub>4</sub> | 12,47          | 133,4             | 1,27          |
| NiCl <sub>2</sub> | 1,81           | 29,1              | 0,039         |

### Costs

For a plant recovering about 3 500 kg nickel sulphate per month, the net profit was shown to be over US \$4 000 (Itoi, 1979).

### Future prospects

Nickel recovery with ED is already carried out on an industrial scale in Japan. This technique will probably find wider application in future for the recovery and reuse of effluents from other galvanization processes.

### Treatment of cooling tower blowdown for water recovery and effluent volume reduction

The range of TDS levels encountered in cooling tower blowdown waters usually varies from about 1 500 to 4 000 mg/l and higher levels at about 4 000 to 12 000 mg/l have also been reported (Jordan, Bearden and McIlhenny, 1975). The disposal of large volumes of this saline effluent can be a serious problem. The application of ED for the treatment of blowdown streams to recover good quality water for reuse and produce a small volume of concentrate promises to be the best prospective system available (Jordan, McIlhenny and Westbrook, 1976; Westbrook and Wirth, 1976).

### Side stream desalting

Blowdown waters from cooling towers can be concentrated ten-fold or more using ED, while recovering and recycling the desalted water to the cooling tower at one-half its original concentration (Wirth and Westbrook, 1977). To accomplish this, blowdown is pretreated, filtered and passed through the ED system.

By recirculation of the brine, it is possible to concentrate the salts into a small stream, while allowing for recovery of about 90% of the water.

The concentration of cooling blowdown waters in an EDR pilot plant at one of Escom's power stations was recently evaluated (Melzer and van Deventer, 1983). Pretreatment of the blowdown water with lime softening, clarification, pH reduction, filtration and chlorination was found to be a basic precondition for successful operation. The operating experience on the EDR pilot plant was sufficiently positive to warrant full-scale application.

Detailed design studies and cost estimates for ED and several other alternative blowdown recovery/concentration systems have been reported (Wirth and Westbrook, 1977). The side stream process design which utilizes ED results in the lowest capital costs for the conditions specified. According to Wirth and Westbrook (1977), it is expected that if the cost comparison were made on overall annual operating costs, the same results would occur.

### Future prospects

It is anticipated that this technique for cooling tower blowdown treatment will find wider application in future, particularly where water reuse and zero effluent discharge is necessary.

### Other possible industrial applications

#### Concentration of sodium sulphate and its conversion into caustic soda and sulphuric acid

A pilot study has demonstrated the feasibility of the concentration of a sodium sulphate solution with ED in a first stage and the subsequent conversion into caustic soda and sulphuric acid in a second stage (Smirnova *et al.*, 1983). The sodium sulphate solution (20 to 40 g/l) was treated in a multi-compartment electro-dialyzer to yield a brine (260-320 g/l, 10% of feed volume) and a product (2 g/l, 90% of feed volume) which could be used as reclaimed water.

The brine was treated further in a three-compartment electro-dialyzer to produce caustic soda and sulphuric acid at a concentration of 17 to 19% by mass and a power consumption of approximately 3,1 to 3,3 kWh/kg sodium sulphate decomposed. The sodium sulphate content of both products was about 1%.

A commercial-scale unit for the conversion of sodium sulphate into sodium hydroxide and sulphuric acid has recently been installed in Japan (undated brochure, Ionics Incorporated).

#### Recovery of acid and caustic soda from ion exchange regeneration wastes

Laboratory results of an electro-dialytic process for acid and caustic recovery from ion exchange regenerant wastes have been described (Nott, 1981). The object of the study was to minimize the discharge of dissolved salts from a water treatment plant producing boiler feed water while recovering some of the pollution abatement process costs from the savings in regenerant chemical costs.

It was shown that the electro-dialytic process for recovery of sulphuric acid and sodium hydroxide from ion exchange regenerant wastes, and substantially reducing the amount of salt discharged to drain, is technically feasible. The nett costs for acid and caustic waste treatment was estimated at US \$4,20 and \$3,00/m<sup>3</sup> waste treated, respectively.

## Concentration of dilute chemical effluents

Laboratory investigations carried out at the NIWR have shown that dilute (approx. 2%) solutions of  $\text{NH}_4\text{NO}_3$ ,  $\text{Na}_2\text{SO}_4$ ,  $\text{NaNO}_3$  and  $\text{NaCl}$  can be concentrated to approximately 20% by ED at an energy consumption of about 1 kWh/kg salt (Schoeman and Botha, 1981). The concentrate volumes were less than 10% of the original volume.

## Conclusions

Membrane fouling can be a serious problem in practical ED unless the necessary pretreatment prescribed by ED suppliers is applied. The overall requirements for pretreatment may be less for standard ED and EDR than for spiral or hollow fine fibre RO membranes owing to the nature of the salt separation process and larger passages provided. However, where organics or multivalent cations are present, more stringent pretreatment may be required for EDR than for RO.

The energy required for ED is less than that of RO at low TDS levels, but this situation is reversed at high feed TDS levels. The economic crossover point between ED and RO based on operating costs is, however, difficult to define precisely and needs to be determined on a site-specific basis.

Seawater desalination by ED is limited at present by high cost. However, recent results have shown that the power consumption can be reduced to levels approaching those for RO seawater desalination, and it is expected that, in future, ED will compete more favourably with RO in this field.

The major use of the EDR process is currently for the desalination of brackish waters. Water saturated with calcium sulphate has been treated successfully without significant fouling problems. The capability of the EDR process to treat serious scale-forming water successfully with little or no chemical addition and the reduction of stack maintenance time are major advantages of the process. However, the more complicated operation and maintenance of EDR equipment, which requires more labour and greater skill (than RO), may be a disadvantage.

The treatment of brackish waters at TDS levels of approximately 300 mg/l and higher with a combination of EDR and IX for industrial purposes appears to be more economic than IX alone. Service runs of IX units have been increased considerably thereby and less waste regenerant is produced. It is foreseen that this combined treatment may find increasing future application owing to more stringent pollution control.

The treatment of nickel plating washwaters for water and nickel recovery is an established industrial process in Japan. This process will probably find wide application in future to include the recovery and reuse of effluents from other galvanization processes.

The treatment of cooling tower blowdown by ED to alleviate the brine disposal problem at power stations has been shown to be the most economical method compared with other alternatives studied. Pretreatment of the blowdown water with lime softening, clarification, pH reduction, filtration and chlorination was found to be a prerequisite for successful operation. It is anticipated that this method may be used on a large scale in future to reduce the waste disposal problem at power stations.

The concentration of dilute chemical effluents for chemical and water recovery, the recovery of acids and bases from IX regeneration wastes, the concentration of sodium sulphate and its subsequent conversion to sulphuric acid and caustic soda have been studied at laboratory or pilot level, and their technical

feasibility demonstrated. However, no large-scale application, with the exception of the conversion of sodium sulphate to sulphuric acid and caustic soda, has been reported.

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