Evaluation of Electrodialysis Reversal for the Desalination of Brackish Mine Service Water

G J G Juby and W Pulles

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CHAMBER OF MINES RESEARCH ORGANIZATION

EVALUATION OF ELECTRODIALYSIS REVERSAL FOR THE DESALINATION OF BRACKISH MINE SERVICE WATER

by

G J G Juby and W Pulles

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The research in this report emanated from a formal collaboration and joint funding between COMRO, the Water Research Commission, Beatrix Gold Mine and Process Plant (Pty) Limited. The title of the project was:

EVALUATION OF ELECTRODIALYSIS REVERSAL FOR THE DESALINATION OF BRACKISH MINE SERVICE WATER.

The Steering Committee for this project consisted of the following persons:

Dr HNS Wiechers	COMRO (Chairman)
Mr JD Fitzgibbons	COMRO (Secretary)
Mr GJG Juby	COMRO
Dr CF Schutte	Water Research Commission
Mr J du Preez	Gencor
Mr RA Jones	Process Plant
Mr L Prince	Process Plant

The financing of the project by the collaborators and the contributions by the members of the Steering Committee are acknowledged.

SUMMARY

Introduction

The mining industry relies heavily on water for its mining operations, notably for cooling in deep mining and stoping.

Mine service water quality varies significantly from mine to mine, and generally this has deteriorated since the Water Boards introduced restrictions and the mines placed more emphasis on recycling and reusing the water. The latter in turn, has lead to far greater corrosion and scaling potential of the mine waters.

Although there are no exact figures on the cost of poor quality service water in mines, the latest estimate puts the figure at up to R300 million per annum for the gold mining industry as a whole. Recent developments such as the use of hydro-power, as well as future implementation of more restrictive legislation on the discharge of waste waters, have highlighted the poor quality of mine service waters currently in use in reticulating systems, and the need for improving their quality.

Of the existing desalination technologies available, evaporative desalination processes are typically very expensive, and for this reason the cheaper membrane desalination processes are considered as holding greater potential for use in treating mine service waters. Although many mine service waters in the gold mining industry contain significant concentrations of dissolved calcium sulphate, which make them more difficult to desalinate, the water under consideration in this particular study contained very little calcium sulphate but high levels of dissolved sodium and chloride. This made it a suitable water for the evaluation of the electrodialysis reversal (EDR) desalination process.

The aims of the study were five-fold, namely: to establish whether or not the EDR process was a technically viable desalination process for brackish type mine waters; to establish the product water recovery and salt rejection levels that could be achieved by the process; to

establish those pretreatment techniques which would be required to ensure good plant performance; to establish whether membrane fouling/scaling would occur whilst treating a mine water with a high sodium chloride content, and to obtain data to establish operator requirements, as well as operating and capital costs for full-scale plants.

Findings and Discussion

A 1,6 ℓ/s EDR pilot plant (Aquamite V-M) was operated on a continuous basis (24 hours per day), treating a mine water stream pumped directly from a surface storage dam at Beatrix G M Co Ltd. Pretreatment of the feed water included: the oxidation of iron and manganese using potassium permanganate; the removal of suspended solids and the products of oxidation by flocculation and then filtration through dual media (anthracite/sand) filters, and removal of any remaining suspended matter by filtration through 10 micron cartridge filters. This pretreatment produced a suitable quality of feed water for the plant. Dosing of sodium hexametaphosphate (SHMP) into the brine loop to prevent barium sulphate scale formation within the plant was also necessary. Under these conditions the plant was able to operate at up to 84 per cent product water recovery.

The mean product water quality, averaged over the 5 900 hour test period, is given in Table 1, together with the mean feed water quality.

Table 1 MEAN FEED AND PRODUCT WATER QUALITIES

DETERMINAND		MEAN FEED VALUES	MEAN PRODUCT VALUES
pH		6,42	6,14
Conductivity at 25°	(mS/m)	451	104
Turbidity	(NTU)	0,96	0,36
Total dissolved solids (TDS)	(mg/ℓ)	3184	637
Ca^{2+}	(mg/ℓ)	100	25
Na^{+}	(mg/ℓ)	1392	193
SO_4^{2-}	(mg/ℓ)	74	4,5

The average salt rejection obtained from the unit was 80 per cent (based on TDS).

As indicated by Table 1, the product water from the EDR unit was of a high quality (conductivity 104 mS/m). This quality was considerably better than that required for effluent discharge as shown in Table 2, which summarizes the major constituents of the product water and the effluent standards as specified by the Department of Water Affairs' General Effluent Standard.

Table 2 COMPARISON OF EDR PRODUCT WATER DETERMINANDS WITH STANDARD EFFLUENT DISCHARGE REQUIREMENTS

DETERMINAND	EDR PRODUCT WATER	GENERAL EFFLUENT DISCHARGE STANDARD
pH	6,14	5,5 to 9,5
Conductivity at 25°C (mS/m)	104	Not to be increased by more than 75 mS/m (determined at 25 °C) above that of the intake water i.e. 135 max.*
Sodium (Na) (mg/ℓ)	193	Not to be increased by more than 90 mg/ℓ above that of intake water i.e. 190 max.*
Manganese (Mn) (mg/ℓ)	0,16	0,4

* Intake water refers to water supplied by the Water Board.

Apart from being able to discharge the water as effluent, the mine could also reuse the water within its reticulation system, since the product water also satisfies the requirements for:

(a) use in a hydro-power system (after slight pH adjustment the quality would satisfy the hydro-power guidelines established by COMRO); and

(b) use as drinking water (based on the total salt content which is within the SABS maximum allowable limits for drinking water; health related parameters were not considered in this study).

Monitoring of chemical usage and power consumption, as well as regular analysis of membrane samples, enabled operating cost estimations to be made. Operating costs for a 46 ℓ/s (4 Mℓ/day) plant, including labour and membrane replacement, were estimated at 76 c/m^3 (June 1990). This figure is based on an electricity consumption of 2,4 kWh/m^3 (a figure obtained from the pilot plant data) and an electricity cost of 9c/kWh. The frequent membrane analyses have enabled membrane life to be estimated at four and seven years for anion and cation membranes, respectively.

Although the EDR process is a modern technology, mine operators with reduction works experience (Patterson grading B3) and a supervisor (Patterson grading C4) coped adequately with the operation of the pilot plant.

A conceptual design for a 46 ℓ/s (4 Mℓ/day) EDR plant, including brine disposal, was carried out giving plant components and plant operating staff requirements, as well as capital and operating costs.

Main Conclusions

1 The EDR process was demonstrated to be capable of successfully desalinating a mine service water containing high levels of dissolved sodium and chloride, and low levels of dissolved calcium sulphate.

2 Under the optimized conditions of feed pretreatment (i.e. potassium permanganate and flocculant dosing, followed by anthracite/sand filtration and cartridge filtration), as well as sodium hexametaphosphate addition to the brine loop, the 1,6 ℓ/s EDR pilot plant was able to achieve a product water recovery of 84 per cent, and average a salt rejection of 80 per cent for the 5 900 hour test period.

3 The product water produced by the plant was suitable for a number of uses, namely: in a hydro-power system (after slight pH adjustment); human consumption (in terms of the SABS maximum allowable limits for total salt content), and as an effluent for discharge to the environment. To be used continuously as a domestic water supply further consideration would have to be given to health related water quality parameters.

4 The membrane samples taken from the stack showed no signs of excessive membrane deterioration. Membrane life has been estimated at four and seven years for the anion and cation membranes, respectively. These figures are similar to those quoted in the literature.

5 It is felt that further pilot plant evaluation of the EDR process on similar feed waters is not required. Data obtained during this study have been sufficient to enable realistic operating cost estimates to be made for full-scale plants. For example, the operating costs for a 46 ℓ/s (4M ℓ/d) plant are estimated at 76 c/m^3 (June 1990) of product. Labour requirements were estimated at eight and two hours per day for operators and a plant supervisor, respectively.

6 The capital cost for a 46 ℓ/s (4M ℓ/d) EDR installation has been estimated at R9,1 Million, with the cost of an 8 ℓ/s brine disposal mechanical vapour recompression evaporator being R3,5 Million. With the brine evaporator operating at 90 per cent water recovery, the total water recovery of the overall desalination system will exceed 98 per cent. The remaining highly concentrated brine could be disposed of in a small crystallizer unit.

PREFACE

The gold mining industry buys large quantities of water from Water Boards for consumption within mines. Because of the ever increasing demand on South Africa's limited water resources and the gradually deteriorating water quality, the cost of high quality Water Board water is increasing steadily.

The use of water for mining results in significant increases in the salt content of the water, making it more corrosive and scale-forming, and this in turn dramatically affects operating costs of mine water reticulation systems. To counteract the build-up of salts in these systems, saline return mine water has to be disposed of, in certain instances to public water bodies. This has resulted in problems with downstream use of the public water, and consequently, the polluters have come under scrutiny from the water authorities. This problem is expected to take on serious proportions in the next 5 to 10 years.

A solution to the aforementioned problems is desalination of the mine water. No two mine waters are identical; however, they may be broadly classified into two distinct groups which are defined as those mine waters which have a scaling potential with respect to calcium sulphate, and those that do not. The latter waters are essentially sodium chloride containing, or brackish waters. COMRO has reviewed conventional desalination technologies and identified the electrodialysis reversal (EDR) process as a technique which holds considerable potential for the desalination of such mine waters.

This report describes the test work carried out on an Electrodialysis Reversal (EDR) pilot plant treating underground mine service water at the Beatrix Gold Mining Co Ltd. - a brackish water with a low calcium sulphate content. It is shown that the plant operated well on this type of mine water, producing a good quality product water. The work is considered definitive and no further test work on this type of feed water using the EDR process is required. Sufficient data are available to make realistic projections for full-scale application of this process. Cost estimates have been made for a typical full-scale plant.

D G WYMER
Director
Underground Environment

SYNOPSIS

This report should be of interest to consulting engineers, mine managers and mine personnel concerned with the problems associated with poor quality mine water, and the methods used to improve its quality.

A 1,6 ℓ/s electrodialysis reversal (EDR) pilot plant has been continuously tested on a mine service water containing a low concentration of dissolved calcium sulphate and a high concentration of sodium chloride. The study confirmed the technical feasibility of applying EDR to this type of water and generated cost data.

The pretreatment steps incorporated in the study were iron and manganese removal by chemical oxidation, and suspended solids removal by dual media and cartridge filtration. Barium sulphate precipitation in the brine stream has been prevented by the addition of an anti-scalant chemical.

Under conditions of suitable pretreatment the pilot plant operates successfully at about 80 per cent salt rejection and at a product recovery of up to 84 per cent. The product water produced by the unit during the 5 900 hours of operation had an average total dissolved salts content of 637 mg/ℓ, making it suitable for hydro-power, effluent discharge and human consumption in terms of its total salt content. Tests on membrane samples have shown no evidence of irreversible fouling, indicating that the pretreatment selected was adequate to protect the membranes against the contaminants present in the mine service water.

The EDR pilot plant study was successful, in that it established the technical feasibility of the process for treating brackish mine water; it highlighted certain areas of pretreatment that are essential to ensure satisfactory plant performance; it demonstrated that the operation of the process is relatively simple and does not require highly skilled operating personnel; and it enabled cost estimates (June 1990) for the process to be made, i.e. for an EDR installation producing 46 ℓ/s (4 Mℓ/d) of desalinated product water the operating

cost would be around 76c/m^3. In addition, by installing an 8 ℓ/s (0,7 Mℓ/d) evaporator unit to treat the brine from the EDR plant, the combined processes would produce 53 ℓ/s (4,6 Mℓ/d) of high quality product water. The total capital cost (1990) of the combined desalination installation is estimated to be R12,6 million (R9,1 million for the EDR plant, and R3,5 million for the brine evaporator), equivalent to 223c/m^3 amortized over 20 years at a 15 per cent per annum interest rate.

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1 INTRODUCTION

The gold mining industry in South Africa uses vast quantities of water. The water qualities, in terms of their dissolved salts content, vary widely amongst the gold mines. Generally, the levels of dissolved salts have increased in recent years due to the implementation of mine water recirculation after restrictions were imposed on the mining industry by the various Water Boards. In view of the current emphasis on water conservation and its effect on the quality of the service water, many reasons exist for desalinating such service water to maintain a given dissolved salts level. These include : protection of the mine service water reticulation system against corrosion and scale formation; providing water for drinking purposes rather than purchasing board water; protecting equipment used in a hydro-power mining system where the hydrostatic head of water is used to power machinery underground(1,2), and treating discharge streams from a mine circuit before they enter a public stream.

Since the water qualities vary so widely in South African gold mines, there is no single desalination process at present that can efficiently treat all service waters encountered. In the current investigation, undertaken at the Beatrix gold mine in the Orange Free State, mine water with a high sodium chloride content but with a low calcium sulphate content was treated. Electrodialysis Reversal (EDR) appeared eminently suitable as a desalination process for treating such water.

2 OBJECTIVES

The objectives of this pilot plant study were as follows:

(i) to establish whether the EDR process could successfully desalinate a mine water high in dissolved sodium chloride and low in calcium sulphate content

(ii) to establish water recovery and salt rejection levels attainable with EDR for this type of feed water;

(iii) to establish pretreatment techniques required to operate an EDR unit on the abovementioned water;

(iv) to establish whether or not membrane fouling/scaling would occur during the test period; and

(v) to obtain data that would enable estimates of level of supervision and operating costs for commercial scale plants to be made.

3 PROCESS DESCRIPTION AND TEST EQUIPMENT

3.1 Electrodialysis

Electrodialysis (ED) is a desalting process in which ions are removed from water by passing through semipermeable membranes that are impervious to water. The driving force for such movement of ions is a direct current electric field. The process is illustrated in Figures 1(a) and (b).

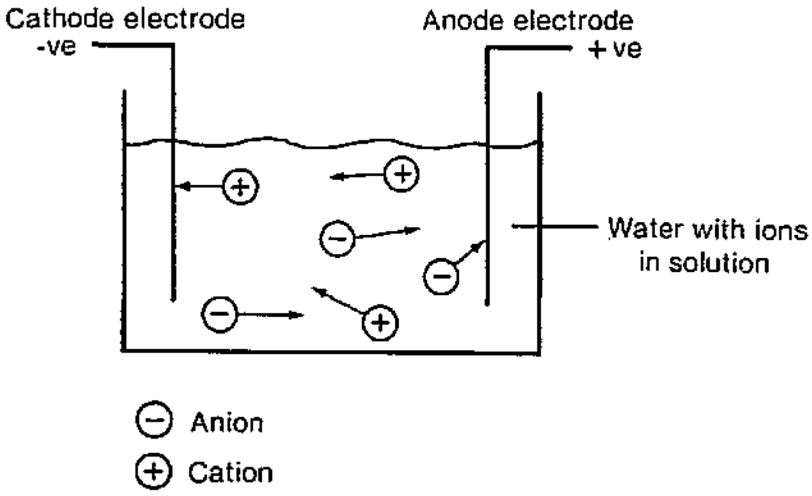

Figure 1(a) MOVEMENT OF IONS IN A DIRECT CURRENT ELECTRIC FIELD

In Figure 1(a) two electrodes are placed in a container of water, which contains ions in solution. The electrodes are connected to a DC source so that one electrode becomes positively charged (the anode) and the other negatively charged (the cathode). Being charged, the ions in solution migrate towards an electrode of opposite charge, i.e. cations to the cathode and anions to the anode. Figure 1(b) shows the same system as Figure 1(a) except that now two pairs of membranes have been included. Each pair consists of two types of membrane, one (the cation membrane) which will only allow positively charged ions to pass through it and the other (the anion membrane) which will only allow negatively charged ions to pass through it.

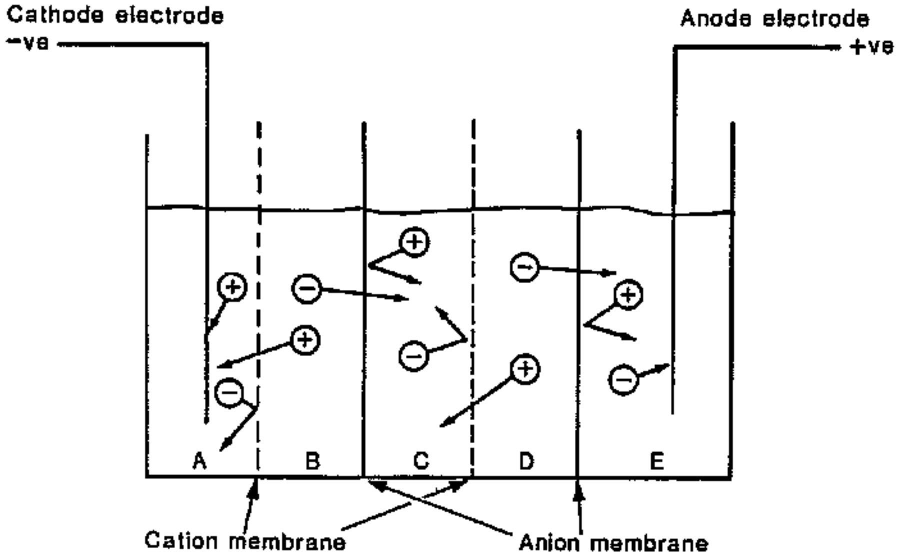

Figure 1(b) A SIMPLIFIED ELECTRODIALYSIS PROCESS

In Figure 1(b) five compartments i.e. A, B, C, D and E have been created within the container. Once a direct current (DC) potential is applied to the electrodes the ions will begin their migration, and anions will pass through the anion membranes and cations will pass through the cation membranes. Since anions cannot pass through cation membranes and vice versa the respective ions remain trapped in compartments A, C and E (Figure 1(b)). With the migration of ions from compartments B and D, compartments A, C and E become more concentrated in ions, and are termed the brine compartments. On the other hand, compartments B and D which are depleted of ions, are termed the product compartments.

Alternate compartments are created by having alternate cation and anion membranes. In practice these membranes, in the form of sheets, are placed together horizontally in a stack and the compartments are created by means of perforated plastic spacers about 0,5 mm thick.

The cation membranes are usually made of cross-linked polystyrene with sulphonate groups ($-SO_3^-$) attached to the polymer. Similarly, the anion membranes are also made from cross-linked polystyrene but in this case they contain the quaternary ammonium groups ($-NR_3^+$). All of the membranes are fabricated by applying the ion selective polymer to a fabric material which gives it structural support.

Experience with the ED process has shown that unacceptably high levels of membrane fouling occur due to certain ions and scales becoming caught in the polymer network. The level of scaling and fouling of the membranes is reduced by reversing the direction of movement of the ions by reversing the DC potential, so that the cathode electrodes become the anodes, and vice versa. This reversal in turn alters the configuration of the compartments in the stack, i.e. what were product compartments become brine compartments and vice versa. This modification of the ED process to incorporate polarity reversal and which basically improves the process is termed Electrodialysis Reversal (EDR). The reversal operation needs to be carried out frequently to be effective, and is typically set to occur about three to four times per hour.

3.2 Experimental site

The Beatrix gold mine pumps approximately 50 Mℓ/d of mine water to the surface (580 ℓ/s). This water consists of about 20 Mℓ/d of spent mine service water (cooled on surface and gravitated underground), and about 30 Mℓ/d of fissure water which enters from the workings. Thus the quality of the mine water pumped out of the mine at present, is determined largely by the quality of the fissure water entering the mine in the workings. The mine water is pumped from underground to a surface hot water well (7,5 Mℓ), from where it is distributed for cooling and use elsewhere on the mine, or pumped to evaporation dams and other mines. The feed water for the EDR pilot plant was pumped directly from the hot water well to the test site.

3.3 EDR pilot plant

The test equipment was the Aquamite V-M, manufactured by Ionics Inc., USA. This unit produces 1,6 ℓ/s (140 m^3/d) of desalted water at a product water recovery of between 75 and 90 per cent. It was not specifically designed for this pilot plant study, but was a commercially available unit considered suitable for this pilot plant investigation.

The water recovery of the unit was set manually according to the operating conditions and chemical composition of the feed. The stack consisted of 900 cell pairs (one cell pair comprises one cation membrane, one anion membrane and two spacers).

Figure 2 shows a simplified flow diagram of the unit including the feed pretreatment steps (see Section 3.4).

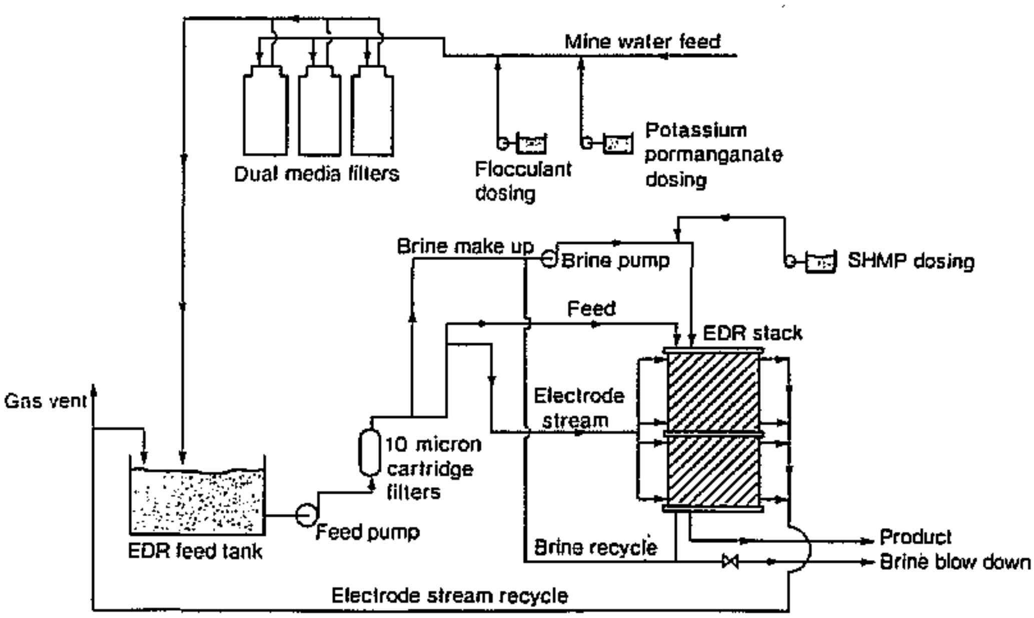

Figure 2 SIMPLIFIED FLOW DIAGRAM OF THE 1,6 ℓ/s EDR PILOT PLANT

When the electrode polarity is reversed, the product water quality deteriorates due to contamination during the switching of the product and brine compartments. During this period, usually 30 to 60 seconds, the product is recycled back to the feed tank to prevent loss of partially desalinated water. This 'off-spec' product is controlled by an in-line conductivity cell. The polarity reversal period on this unit was set at between 15 and 20 minutes.

Manual valves were used for controlling feed and product flow rates as well as setting stack pressures. A series of pressure gauges enabled detailed monitoring of the status of the plant to be carried out. The voltage and current across each of the electrical stages were monitored.

The brine pump circulates the concentrate or brine stream through the stack. This stream is separated from the feed and product or dilute streams by the membranes. The brine loop is maintained at a pressure slightly below that of the dilute stream (450 and 480 kPa respectively) so that if there is any leakage across the membranes in the stack, it will take place from the dilute stream into the brine, thus preventing pollution of the product stream. Such leakage in the stack takes place continuously and is known as cross-leakage.

Brine was continuously purged from the system and make-up was maintained continuously from the feed stream. The electrode-flush stream was also supplied from the main feed stream. This water is used to continuously flush the electrode compartments of the stack to remove any products of electrolysis (e.g. hydrogen, chlorine and oxygen) and is recycled to the feed tank to prevent water loss and to allow the gases to vent to the atmosphere.

3.4 Pretreatment of feed water to EDR pilot plant

Mine water contains constituents which are undesirable and harmful to the EDR process, and which cause fouling and/or scaling within the membrane stack.

The pretreatment processes that were carried out on the feed water during this study are shown schematically in Figure 2, and can be summarised as follows:

(i) Suspended solids removal

Flocculant addition at a level of around 1 mg/ℓ (ANIKEM 8100), followed by filtration through dual media down-flow anthracite/sand filters. The final step in the suspended solids pretreatment was the 10 μm cartridge filtration. The turbidity of the feed water was reduced from about 20 NTU to less than 1 NTU in the feed water to the plant i.e. from about 40 mg/ℓ to about 2 mg/ℓ suspended solids.

(ii) Manganese and iron removal

Initially, dissolved iron was not removed from the mine water and this resulted in iron fouling within the stack. Later this foulant was removed by dosing with potassium permanganate (to achieve a concentration of about 7 mg/ℓ $KMnO_4$) followed by dual media anthracite/sand filtration, which reduced the level of iron in the feed water to less than 1 mg/ℓ. Unfortunately the manganese levels in the mine water were found to be mostly unaffected by the $KMnO_4$ dosing (see Appendix II).

(iii) SHMP addition

Sodium hexametaphosphate (SHMP) was introduced into the brine stream (to achieve a concentration of 15 mg/ℓ) to prevent barium sulphate precipitation and scale formation within the membrane stack (see Appendix III).

More details of the pretreatment steps employed are given in Section 5.3.

4 TEST WORK PROGRAMME

The pilot plant unit at the Beatrix gold mine operated for 5 900 hours during the period May 1986 to May 1987.

The test site was manned on a 24 hour basis by an operator, with a supervisor in attendance daily during normal office hours. Data including pressures, flow rates, amperages and voltages were collected every four hours.

In addition, during the morning shift chemical analyses of the feed, product and brine streams were carried out at the site to determine the following:

Conductivity
pH
Turbidity (NTU)
Temperature
Chloride

Iron Manganese	feed only

Once a week, samples of feed, product and brine were collected and sent for chemical analysis at the Beisa Mine laboratory. The following analyses were carried out:

pH
Conductivity at 25 °C
Total dissolved solids
Calcium hardness
Chloride
Sulphate
Sodium
Iron
Manganese
Barium

By completion of the 5 900 hour operational period some 190 sets of plant performance data together with 40 sets of weekly chemical analysis data had been collected. A detailed discussion of these analyses is given in Section 5.

In order to eliminate certain foulants within the stack that were not removed sufficiently during the polarity reversals, it was necessary to carry out Clean In Place (CIP) operations after approximately every 1 000 hours of operation. This involved pumping a 5 per cent solution of hydrochloric acid through the stack for about 2 hours, after which the stack was flushed thoroughly with clean water before re-introducing the feed water.

5 FINDINGS AND DISCUSSION

5.1 Water qualities and salt rejection

Table 1 and Figure 3 present the mean values of the chemical constituents of the feed, product and brine streams, taken from both the weekly chemical analyses and the daily performance data over the total study period.

Table 1 MEAN CHEMICAL ANALYSES OF EDR PROCESS STREAMS - FEED, PRODUCT AND BRINE

Chemical constituent		Feed	Product	Brine
pH		6,42	6,14	5,42
Conductivity at 25 °C	(mS/m)	451	104	1116
Turbidity	(NTU)	0,96	0,36	0,35
TDS	(mg/ℓ)	3184	637	9135
Ca^{2+}	(mg/ℓ)	100	25	NA
Cl^-	(mg/ℓ)	1748	376	4990
SO_4^{2-}	(mg/ℓ)	74	4,5	344
Na^+	(mg/ℓ)	1392	193	3110
Iron	(mg/ℓ)	0,63	0,45	0,90
Manganese	(mg/ℓ)	0,59	0,16	1,39
Barium	(mg/ℓ)	0,59	0,27	1,01

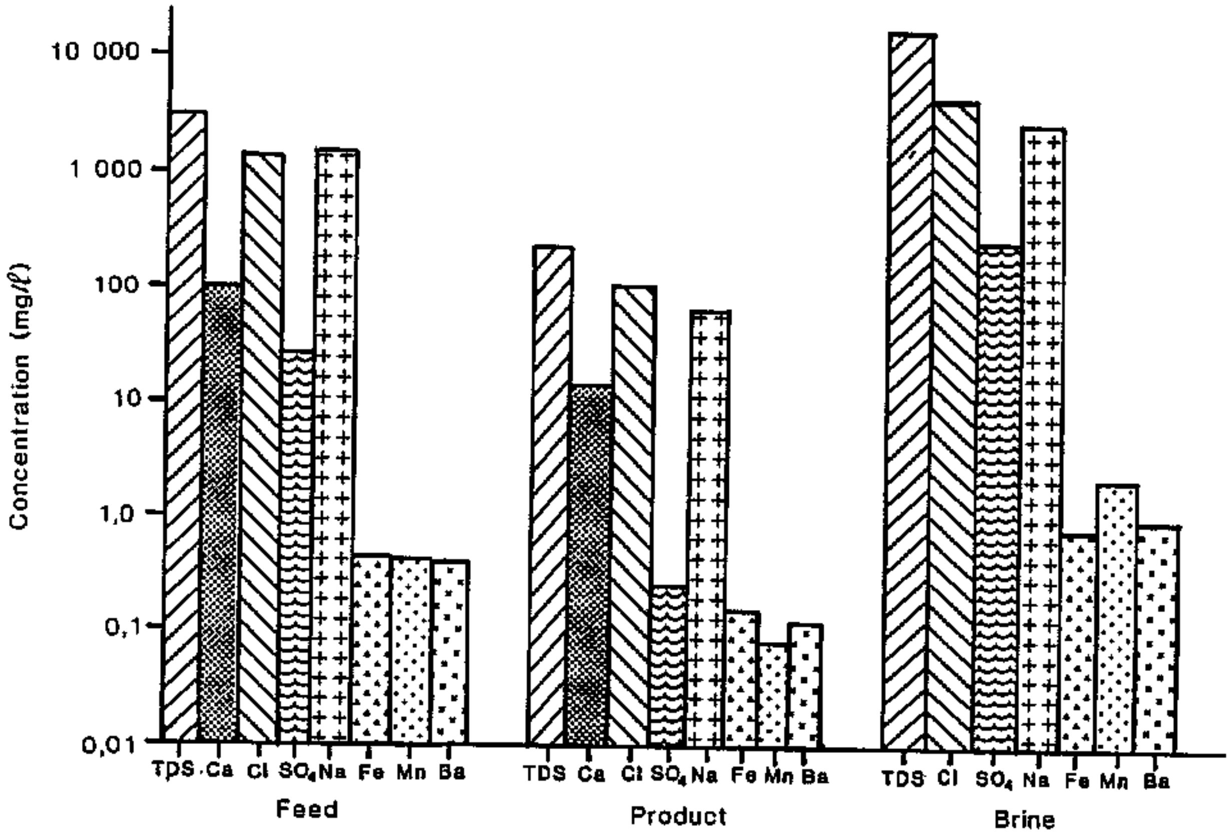

Figure 3 COMPARISON OF DISSOLVED COMPONENTS IN FEED, PRODUCT AND BRINE

The quality of the feed water in terms of TDS remained relatively constant throughout the experimental period. The iron and manganese levels of the raw feed to the pretreatment stage, however, varied significantly (see Section 5.3.2).

A problematic feature of many mine waters is their calcium content. The calcium level in the feed averaged 100 mg/ℓ (see Figure 4), an acceptably low level in terms of calcium sulphate scaling potential (less than 1 per cent saturated with respect to calcium sulphate). This is not surprising since the mine was not adding lime (calcium hydroxide) to neutralize any of the mine water underground.

Figure 5 compares the conductivity measurements of the feed and the product taken on a daily basis. As can be seen there is a vast improvement in the product water quality in terms of conductivity, with product water values consistently around 100 mS/m, and not significantly affected by the feed water conductivities which varied between about 300 mS/m and 600 mS/m.

Also shown in Figure 5 is the salt rejection of the unit, based on conductivity, defined as:

$$\%\ \text{Salt rejection} = \left(\frac{\text{Feed conductivity - Product conductivity}}{\text{Feed conductivity}}\right) \times 100 \qquad (2)$$

As can be seen, there was a certain variation in the salt rejection, but it was generally between 70 and 80 per cent.

This method of calculating salt rejection offered a quick and simple technique which could be carried out at the site just by measuring two conductivities. If the TDS values were known then TDS could be substituted for conductivity in Equation (2). The average salt rejection in terms of TDS was 80 per cent.

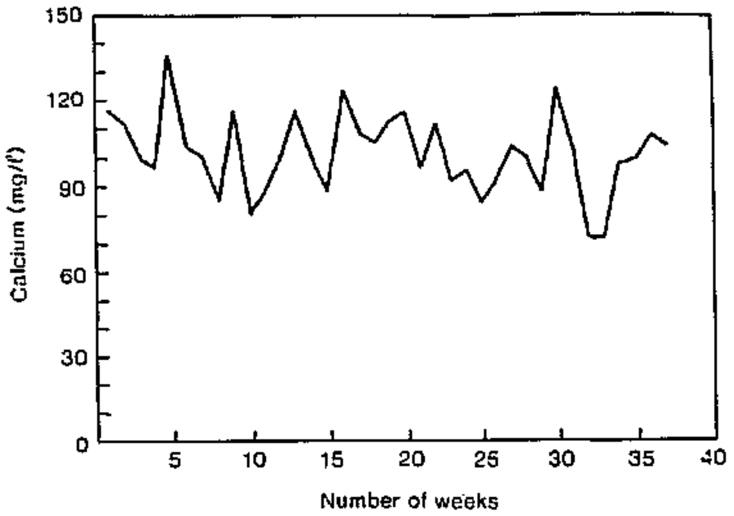

Figure 4 VARIATION IN CALCIUM CONCENTRATION IN FEED WATER

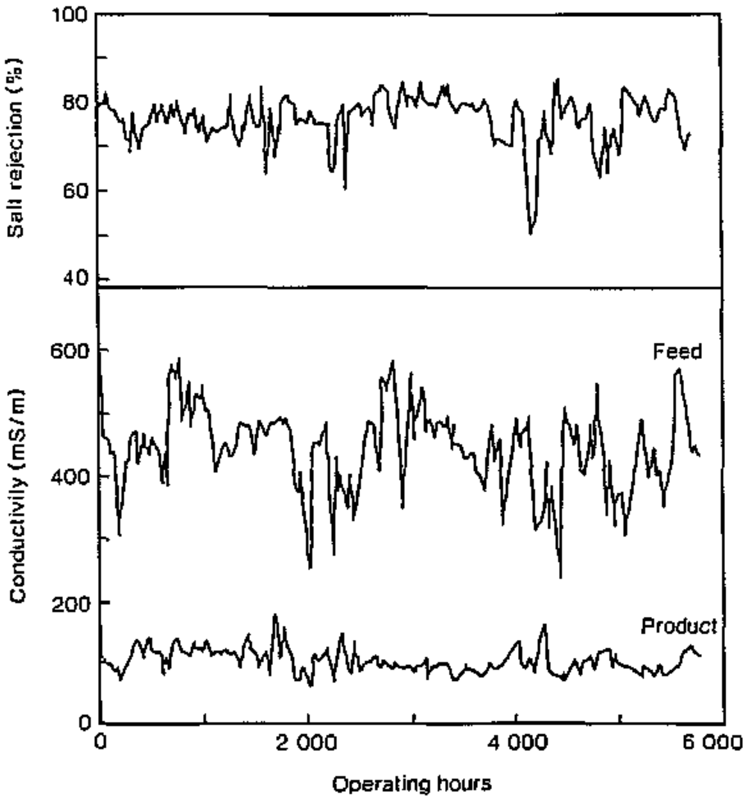

Figure 5 VARIATION IN FEED AND PRODUCT CONDUCTIVITY AND SALT REJECTION BASED ON CONDUCTIVITY

Using average values obtained over the experimental period, the salt rejected by the membranes was 83,5 per cent, or considered another way, the product water contained 16,5 per cent of the salts in the feed water.

The average rejection of sodium (Na^+) (based on Equation (2)) was 86 per cent and chloride ($C\ell^-$) was 78 per cent, whereas the calcium (Ca^{2+}) rejection was 75 per cent and the sulphate ($SO_4{}^{2-}$) rejection was 94 per cent.

The quality and quantity of the brine produced by a desalination plant are of great importance from the point of view of suitable brine disposal. It can be seen from Table 1 that the TDS of the brine averaged approximately 9 000 mg/ℓ.

5.2 Water recovery

Water recovery is an important parameter since it determines the volume of brine that will require disposal. It is defined as follows:

$$\% \text{ Water Recovery} = \frac{\text{Water recovered as product}}{\text{Total feed water to plant}} \times 100 \qquad (4)$$

Figure 6 shows that there was a certain amount of fluctuation in the water recovery over the test period. During the first 2 500 hours of the investigation the system was set to maintain a recovery of around 80 per cent. During this period the pretreatment of the feed water to the unit comprised suspended solids removal (i.e flocculation, dual media filtration and cartridge filtration), and chlorine dosing to reduce the possibility of iron fouling of the membranes. This initial pretreatment step was not totally successful since iron fouling and scale formation occurred in the stack. To prevent iron fouling the pretreatment of the feed was improved, as discussed later, and the recovery was adjusted to around 75 per cent to prevent scale formation. The unit was operated successfully at this recovery level for the next 1 000 hours.

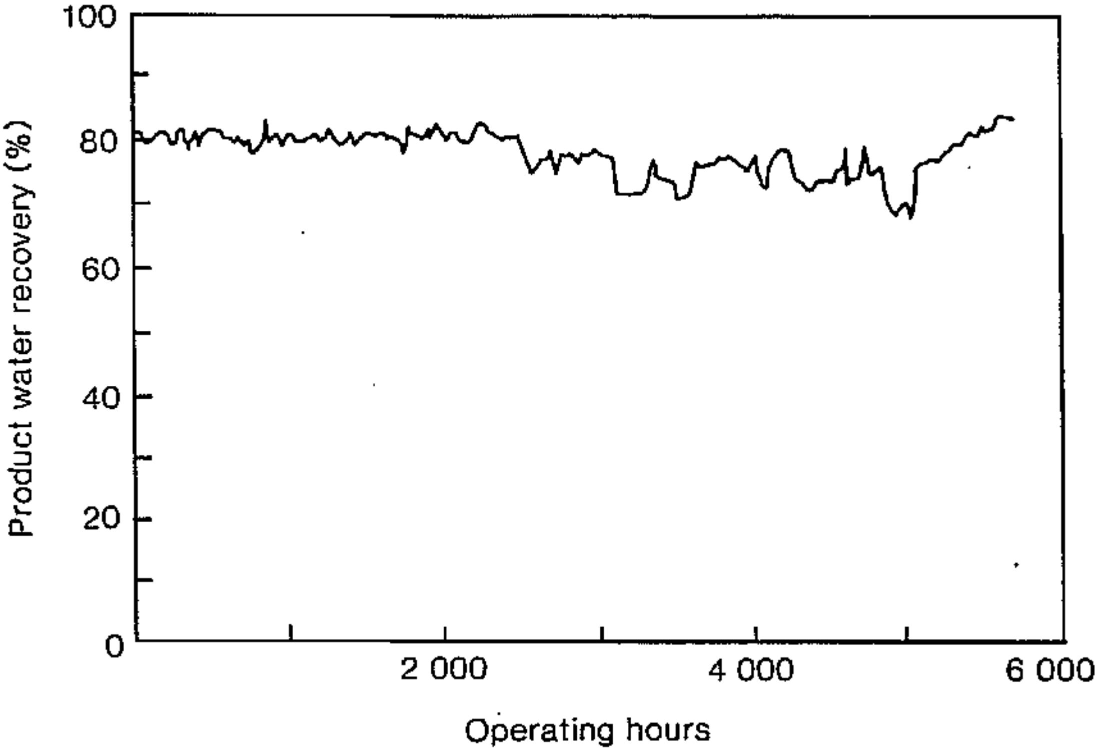

Figure 6 VARIATION OF ACTUAL PRODUCT WATER RECOVERY

For the last 900 hours of the plant operation, sodium hexametaphosphate (SHMP) was dosed into the brine loop to prevent the formation of barium sulphate scale, and the recovery was gradually increased from about 70 per cent to around 85 per cent without the recurrence of the scale.

5.3 Pretreatment

In this section more details are given concerning the pretreatment steps that were employed during the test work. Raw mine water refers to the untreated mine service water and EDR feed water refers to the pretreated mine service water.

5.3.1 Removal of suspended solids

In order to obtain a rapid assessment of the suspended solids content of the various process streams, turbidity measurements were made at the test site. The turbidity of the raw mine water being pumped to the site averaged around 20 NTU, which is equivalent to a suspended solids content of about 40 mg/ℓ. The suspended solids pretreatment steps employed were:

(i) Flocculation

Anikem 8100 liquid flocculant, a highly charged moderate molecular weight cationic polyamine coagulant, was dosed at a concentration of around 1 mg/ℓ of feed water to the plant. This was actually a secondary treatment step since the mine carried out flocculation using Magnafloc 351, a nonionic polyacrylamide, followed by settling underground.

(ii) Dual media filtration

Simple pressure dual media downflow anthracite/sand filters were used for this step. These filters were very effective, reducing the turbidity of the feed water from around 20 NTU to less than 1,0 NTU, (about 2 mg/ℓ suspended solids). The sizes of media used in the filter were: anthracite particle size - 0,8 to 1,5 mm, and the sand particle size - 0,5 to 0,7 mm.

(iii) Cartridge filtration

This was carried out using ten standard 300 mm, nominal 10 µm polypropylene cartridges, housed in a cartridge filter holder in the feed line from the sand filters. The frequency with which the cartridges had to be changed was between 3,5 and 10 days, which is a very short life. The associated cost of this frequent cartridge filter replacement is a substantial part of the operating costs of the pilot plant (see Sections 5.5 and 6.2).

The variation in the EDR feed water turbidity during the test period is shown in Figure 7. The values were well above 1,0 NTU for the first 2 000 hours of operation even though, as mentioned above, the sand filters were producing a product of less than 1,0 NTU. The reason for these relatively high turbidities in the feed water to the stack was due, in part, to the electrode stream water which was returned to the feed tank. Each electrode in the stack has an electrode compartment associated with it, which is flushed continuously by the electrode stream to ensure no build-up of electrolysis products (for example hydrogen and chlorine) in the electrode compartment (see Figure 2). This stream was found to have a turbidity exceeding 1,0 NTU and for the remainder of the test period was not returned to the feed tank. In addition, the frequency of backwashing the sand filters was increased from once per day to once per shift. From that point onwards, the turbidity of the feed water to the unit remained well below 1,0 NTU.

5.3.2 Pretreatment for iron and manganese

Figure 8 shows the levels of iron in the raw mine water being pumped to the test site. In the early stages of the testwork the levels of iron in the feed water were not envisaged as being

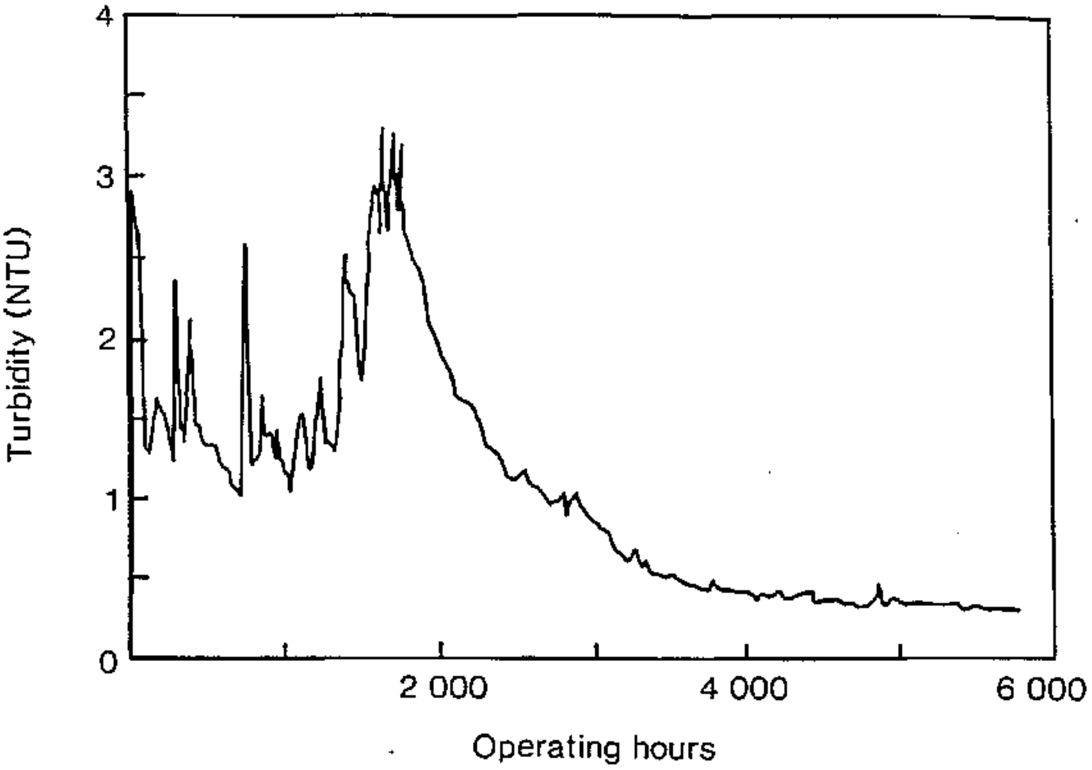

Figure 7 VARIATION IN TURBIDITY OF EDR FEED WATER

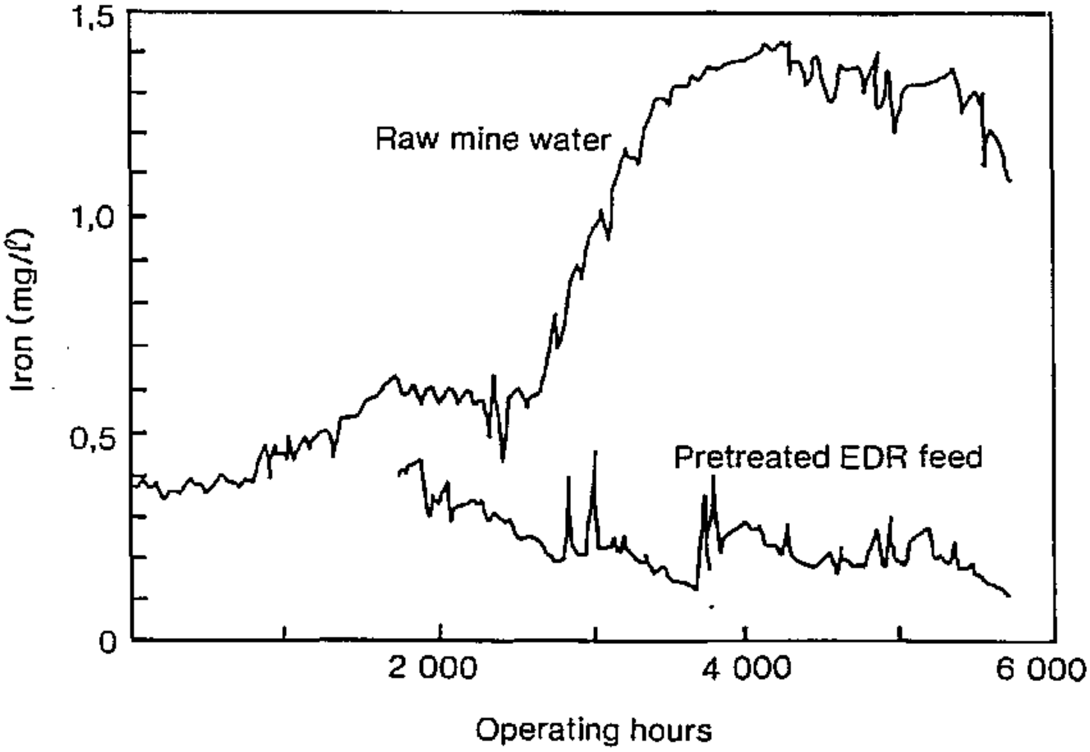

Figure 8 IRON CONCENTRATION IN RAW AND PRETREATED MINE WATER FEED

problematic. In fact manganese was thought to be the potentially more harmful ion, even though no permanent fouling of the membranes had been reported by the manufacturers who inspected membrane samples taken from the stack after 1 000 hours of operation.

Nevertheless, it was decided to operate the unit under conditions closely resembling those applicable to a commercial plant, and dosing of calcium hypochlorite began after 1 650 hours of operation. The free chlorine was expected to oxidize the manganese present in the water to form insoluble salts which would then be removed by the filters. However, the calcium hypochlorite had very little effect, and consequently sodium hypochlorite was tried as an alternative oxidizing agent after 2 300 hours. This reduced the iron levels to about 0,3 mg/ℓ, but once again there was no noticeable effect on the manganese level. Laboratory tests were carried out to investigate the effect of the hypochlorite on the turbidity of the water, and showed that it gave rise to an increase in turbidity. It was decided, therefore, not to use the hypochlorite as a chemical oxidizing agent, and after 3 400 hours of operation potassium permanganate ($KMnO_4$) dosing was introduced. It can be seen from Figure 8 that the level of iron in the raw feed water had increased to well above 1,0 mg/ℓ at this stage of the test work. The dosing of $KMnO_4$ was continued for the remainder of the test period and effectively reduced the iron concentration. However, the manganese levels were not reduced by the oxidizing agent but actually increased slightly. The most likely explanation as to why the manganese levels remained around 0,6 to 0,7 mg/ℓ is that the oxidation of Mn(II) by potassium permanganate requires about five minutes of contact time(8), and the total contact time during the study was only one minute. On a full-scale plant both the manganese and iron values would have to be controlled at a level below 0,3 mg/ℓ (i.e. total iron plus manganese less than 0,3 mg/ℓ). Methods for doing so are outlined in Section 6.1 and in Appendix VIII.

5.3.3 Treatment for preventing barium sulphate scale formation

Barium sulphate is a very insoluble salt. Its saturation point is dependent upon the other ions in solution, but it can be reached at levels as low as 2 mg/ℓ. Barium was present in the feed water to the unit, as can be seen from Table 1, with an average concentration of 0,6 mg/ℓ as barium. In fact, the feed water to the unit was saturated with respect to barium sulphate (see Appendix III).

Precipitation in the concentrated brine loop could thus be expected, and would increase as the product recovery of the plant increases. At recoveries below 70 per cent there was no evidence of $BaSO_4$ scale within the brine loop. In the last 1 000 hours of operation the recovery of the unit was increased to 85 per cent, a condition under which $BaSO_4$ precipitation occured. To overcome this problem sodium hexametaphosphate (SHMP), a chelating agent, was dosed at 15 mg/ℓ.

The figures given above, together with operational experience, suggest that when operating an EDR unit on water with a similar feed chemistry at a recovery of 75 per cent or higher, a scale inhibitor such as SHMP should be dosed into the brine loop.

5.4 <u>Operational difficulties</u>

During the 6 000 hour pilot plant study the majority of problems encountered involved the membrane stack. As mentioned previously, the inadequate pretreatment of the feed water in the early stages of the study resulted in fouling and scale formation within the membrane stack. When this occurred it was necessary to shut down the plant, strip down the membrane stack, physically clean the membranes (by lightly scrubbing them), and in some cases where the membranes had been damaged, it was necessary to replace them. In total 32 membrane pairs, two electrodes, and two motor operated valves were replaced during the study. Details of the major operational events which occurred during the study are given in Appendix V.

The availability of the unit was not easy to quantify, since by the very nature of the test programme the pilot plant was subjected to unusual and extreme operating conditions.

5.5 Pilot plant operating costs

During the pilot study the power input to the plant was monitored, as were all the costs for pretreatment chemicals, and the costs for cartridge filter replacement. The operating costs have been broken down into electricity costs, based on a figure of 9,0 c/kWh (1990), chemical costs (this includes the costs for the acid used in the "Clean-In-Place" operations), pretreatment costs (this includes the cost of chemicals for pretreatment), and cartridge filter costs. Costs for membrane replacement and labour have been excluded from this section since they would be unrealistic considering the duration of this study and the variety of operating conditions to which the plant had been subjected. Table 2 summarizes the operating cost data.

Table 2 PILOT PLANT OPERATING COSTS (1990) INCURRED EXCLUDING MEMBRANE REPLACEMENT AND LABOUR COSTS

Category	Explanation/Breakdown	Cost c/m^3
Electrical	2,4 kWh/m^3 at 9 c/kWh	21,6
Pre-treatment	Flocculant at 1 mg/ℓ-0,8 c/m^3)	3,1
	$KMnO_4$ at 3 mg/ℓ-1,3 c/m^3)	
	SHMP at 15 mg/ℓ-1,0 c/m^3)	
Clean in place	HCl acid clean every 1 000 hours	0,5
Cartridge filters	Changing every 5 days	9,0
	Total	34,2

Cost estimates for full-scale plants are discussed in Section 6.2.

5.6 Product water uses

Table 3 summarizes the product water chemical constituents and shows the SABS limits for potable water$^{(4)}$, the present provisional guidelines for hydro-power water quality as established by COMRO, as well as the General Effluent Standards$^{(10)}$.

The average quality of the product water from the EDR unit lies well within the SABS maximum allowable limit for dissolved salts in drinking water. However conductivity, chloride and sodium exceed the SABS recommended limits. Table 4 presents results on the bacterial content of a grab sample of product water which conform to the specifications for bacterial content for potable water. It should, however, be noted that regular bacterial sampling is necessary (Appendix IV), to ensure compliance with bacterial count limits.

In terms of hydro-power water quality, the EDR product water meets all the requirements apart from its slightly low pH. In Table 3 the general effluent discharge standard refers to the quality of regional 'intake water' for both conductivity and sodium. The intake water quality is taken as the quality supplied by the Orange Free State Water Board at Bothaville to the Free State Mines. Such water has an average conductivity of 88,5 mS/m and a sodium concentration of 56 mg/ℓ, thus increasing the maximum allowable limits for an effluent to 163,5 mS/m and 146 mg/ℓ for conductivity and sodium, respectively. The average quality of the product water satisfies all the requirements of the general effluent standard except for the sodium concentration. The average sodium concentration is 193 mg/ℓ and would thus have to be reduced by about 25 per cent.

Table 3 <u>COMPARISON OF PRODUCT WATER QUALITY WITH LIMITS FOR HUMAN CONSUMPTION, HYDRO-POWER AND EFFLUENT DISCHARGE</u>

Chemical constituent		Average product composition	SABS Recommended limit*	SABS Maximum allowable limit*	Hydro-power guidelines		Requirements for the purification of waste water or effluent, General standard(10)
					For stainless pipework	For carbon steel pipework	
pH		6,14	6,0 to 9,0	5,5 to 9,5	6,5 to 8,5	6,5 to 8,5	5,5 to 9,5
TDS	(mg/ℓ)	637	NS	NS	Limited by chlorides	1 500	NS
Conductivity at 25 °C	(mS/m)	104	70	300	NS	NS	Not to be increased by more than 75 mS/m (determined at 25°C) above that of the intake water. (+)
Turbidity	(NTU)	0,36	1	5	3	3	Suspended solids not to exceed 25 mg/ℓ
Chloride (as Cl)	(mg/ℓ)	376	250	600	250	limited by TDS	NS
Calcium (as Ca)	(mg/ℓ)	25	NS	NS	(CaSO₄ ionic product less than Ksp CaSO₄ at 0 °C)	(CaSO₄ ionic product less than Ksp CaSO₄ at 0 °C)	NS
Sulphate (as SO₄)	(mg/ℓ)	45	200	600			NS
							NS
Magnesium (as Mg)	(mg/ℓ)	<1	70	100	NS	NS	
Sodium (as Na)	(mg/ℓ)	193	100	400	NS	NS	Not to be increased by more than 90 mg/ℓ above that of the intake water. (+)
Nitrate (as N)	(mg/ℓ)	Nil	6	10	NS	NS	NS
Fluoride (as F)	(mg/ℓ)	1	1,0	1,5	NS	NS	1,0
Total hardness (as $CaCO_3$)	(mg/ℓ)	65	20 min. 300	650	Limited by $CaCO_3$ and $CaSO_4$ saturation	Limited by $CaCO_3$ and $CaSO_4$ saturation	NS

* Both of the limits specified as 'recommended' and 'maximum allowable' represent water fit for human consumption and for domestic purposes, but the 'recommended' limit should if possible be applied to all water supplied for domestic use(3).

NS Not specificed

\+ In take water is taken as the water supplied by the Orange Free State Water Board at Bothaville to the Free State Mines.

Table 4 BACTERIAL COUNTS ON EDR FEED AND PRODUCT WATERS

Sample	Result	
	Total plate count	Total coliform count
EDR Feed	130/mℓ	1/100 mℓ
EDR Product	50/mℓ	2/100 mℓ
Typical drinking water specification	1000/mℓ	5/100 mℓ
SABS Domestic water specification (SABS 241-1984) maximum allowable limit	Not specified	* 5/100 mℓ

* if >0 <5 then some additional considerations are necessary. See SABS 241 - 1984.

To summarize, the product water from the EDR unit is of a quality that:

(i) contains 80 per cent less chloride than the feed water;

(ii) is non-scaling with respect to calcium sulphate;

(iii) is suitable for use in a hydro-power system after slight upward pH adjustment;

(iv) is suitable for human consumption only in terms of the SABS maximum allowable limits for salt in drinking water;

(v) would meet the standard effluent discharge requirements if the sodium content was reduced to less than 146 mg/ℓ; and

(vi) would be suitable for continuous domestic use, provided that the conductivity, and concentration of chloride and sulphate could be reduced by further treatment, in order to comply with the SABS recommended limits. Furthermore, it would require disinfection for proper control over bacterial contamination.

6 CONCEPTUAL DESIGN AND COST ESTIMATES FOR A COMMERCIAL SCALE PLANT

The Aquamite V-M EDR plant was used in this investigation since it has virtually all the attributes of a full-scale plant. Owing to the standard stack components, the pilot plant test data can be extrapolated in a modular fashion to commercial scale plant(6). A full-scale application would be designed to produce a product water quality according to specification.

This section is included to give an indication of the size of a commercial installation and pretreatment facilities, as well as giving an estimate of the operating costs anticipated for a 46 ℓ/s (4 Mℓ/d) EDR plant.

6.1 Conceptual ground arrangement of a 46 ℓ/s (4 mℓ/d) EDR plant - 85 per cent recovery

At a product flow rate of 46 ℓ/s and a product recovery of 85 per cent the feed flow rate to the unit will be 54 ℓ/s (4,7 Mℓ/d) with a brine flow rate of 8 ℓ/s (0,7 Mℓ/d), (see Section 6.3). Figure 9 shows a conceptual ground arrangement for a 46 ℓ/s installation. The pretreatment section is shown, as well as some tank volumes and the approximate size of the site.

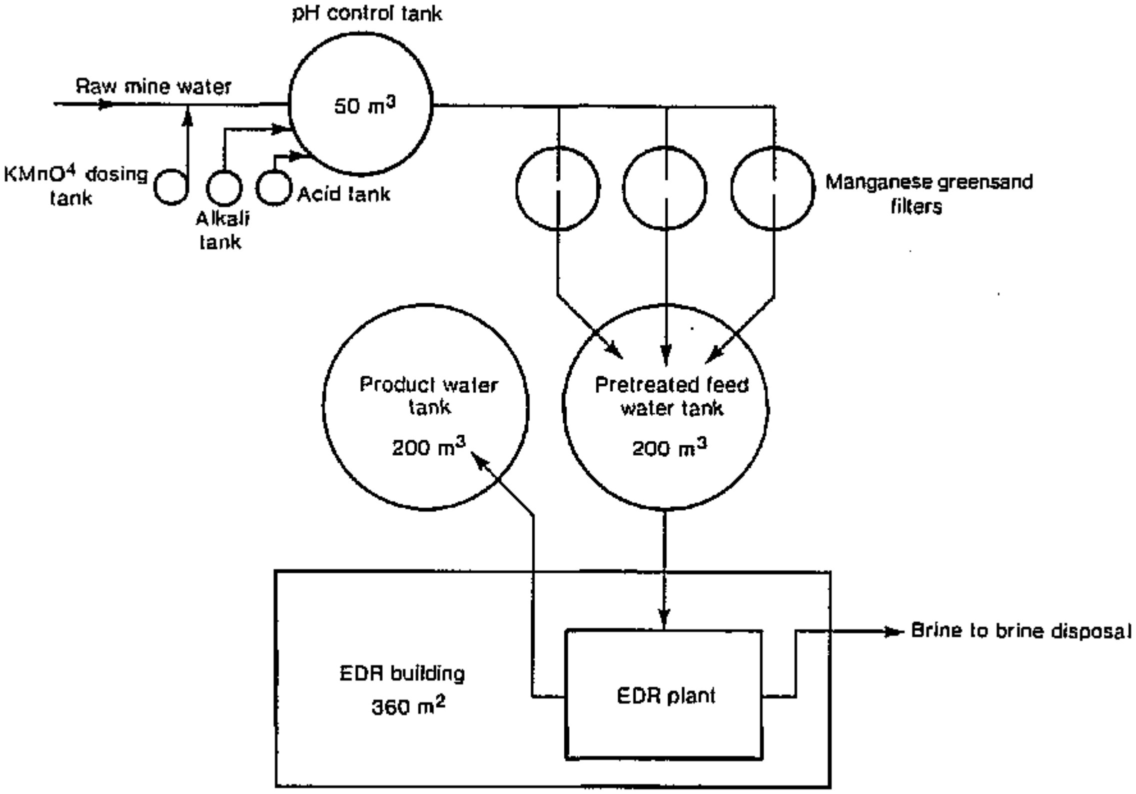

Figure 9 CONCEPTUAL GROUND ARRANGEMENT FOR A 46 ℓ/s COMMERCIAL SCALE EDR PLANT

The major components of the conceptual 46 ℓ/s EDR plant are listed in Table 5.

Table 5 SUMMARY OF MAJOR COMPONENTS OF A 46 ℓ/s EDR PLANT

Component	Details
EDR Plant (46 ℓ/s product)	* 18 membrane stacks * 360 m^2 roofed floor area required
Pretreatment	* Chemical dosing tanks, 3 x 2 000 ℓ * pH control tank, 50 000 ℓ * Clarifier, 200 000 ℓ * Manganese greensand filters, 3 x 20 000 ℓ * Pretreated feed storage tank, 200 000 ℓ * Product tank, 200 000 ℓ

Further details concerning the choice of the pretreatment steps and tank sizes are given in Appendix VI.

6.2 Cost estimation for commercial scale plant

6.2.1 Operating cost estimation

These costs are based on the figures obtained during the pilot study, which have been extended to include the predicted costs for a full-scale plant corresponding to the conceptual design shown in Figure 9. The costs include electricity, pretreatment, Cleaning in Place operations, cartridge filters, labour, maintenance, physical stack cleaning, and membrane replacement (Appendix VII). A breakdown of the total anticipated operating cost of 76,1 c/m^3 produced (June 1990), is given in Table 6.

Table 6 ESTIMATED OPERATING COSTS FOR A 46 ℓ/s COMMERCIAL SCALE EDR PLANT (JUNE 1990)

Category	Details	Cost (c/m³Produced)
Electrical	2,4 kWh/m³ at 9 c/kWh	21,6
Pretreatment	Flocculant 0,6 c/m³ $KMnO_4$/Greensand filter 1,3 c/m³ SHMP 1,1 c/m³	3,0
CIP	HCl acid wash every 1 000 hours	0,5
Cartridge filters	Based on cheaper supply cost and commercial plant operational improvements	9,0
Labour	4 operators at R2 500/month (third of time) 1 supervisor at R4 600/month (quarter of time)	3,8
Maintenance	18 membrane stacks	0,9
Stack cleaning	6 labourers for 16 hours every 5 000 hours per 3 stacks at R3,50/hour	0,1
Membrane stack replacement	R30 000/stack per year	37,2
Total operating cost		76,1

6.2.2 Capital cost estimatation

In order to estimate the capital costs realistically, a firm of consulting engineers (Stewart, Sviridov and Oliver) was employed to establish the capital cost of designing, constructing and commissioning a hypothetical commercial scale desalination installation. The extensive investigation included the pretreatment and desalination of a brackish mine service water from a total dissolved salts content of 1 800 mg/ℓ to 400 mg/ℓ. The study showed that an EDR plant producing 46 ℓ/s at 85 per cent recovery would have a total depreciable capital cost of R9,1 million (R197 x 10^3 per ℓ/s) (June 1990). This cost includes the purchase of the initial membranes for the plant, which would total approximately R3,2 million.

6.3 Brine disposal

To complete the conceptual design, consideration must be given to brine disposal. The brine flow rate from a 46 ℓ/s plant operating at 85 per cent recovery would be 8 ℓ/s (0,7 Mℓ/d).

Only two brine disposal techniques are commercially viable at present, i.e. evaporative ponds - a relatively slow process, and mechanical vapour recompression evaporation - a relatively expensive but rapid and continuous process producing a high quality product that can be returned to the system, thereby increasing the overall water recovery.

Since approximately 200 000 m^2 of evaporation pond area would be required to dispose of 8 ℓ/s of brine, it was decided to consider the use of a mechanical vapour recompression evaporator as the brine disposal technique. The capital cost for such a unit, operating at 90 per cent recovery, would be roughly R3,5 million (1990). Such a plant would have an estimated operating cost of around 260 c/m^3 of brine treated. It is possible that salt could be recovered from this unit for resale by incorporating a small crystalliser.

6.4 Cost summary

Table 7 gives a summary of the operating and capital cost estimates anticipated for a 46 ℓ/s EDR installation including the disposal of 8 ℓ/s of brine.

Table 7 COST SUMMARY FOR A 46 ℓ/s EDR INSTALLATION

Component	Cost
Capital cost	
EDR Plant (46 ℓ/s) including: 18 membrane stack plant&building Pretreatment facilities&tanks Product storage tank	R9,1 million
Brine disposal (8 ℓ/s) Mechanical vapor recompression evaporator	R3,5 million
Total capital cost	R12,6 million
Operating cost	
EDR Plant (46 ℓ/s) 85% recovery	76,2 c/m^3
Brine evaporator (8 ℓ/s) 90% recovery	260 c/m^3
Total operating cost	101 c/m^3 product
Overall water recovery (53,2 ℓ/s)	98,5 %

The total operating cost per m^3 product is calculated as follows:

$$\left(\frac{\text{EDR product flowrate}}{\text{Total product flowrate}} \times \text{EDR plant operating cost}\right)$$

$$+ \left(\frac{\text{Evaporator product flowrate}}{\text{Total product flowrate}} \times \text{Evaporator operating cost}\right)$$

where Evaporator product flowrate =
Evaporator feed flowrate x Evaporator water recovery

$$= \left(\frac{46}{53,2} \times 76,2\right) + \left(\frac{(8\text{x}0,9)}{53,2} \times 260\right) = 101 \text{ c/m}^3$$

To give an indication of the total amortized cost of the plant over a period of 20 years (capital and running) an interest rate on capital of 15 per cent per annum has been used (assumed constant over the 20 year period). Furthermore, it is assumed that the total plant (i.e. civil and mechanical structure) will have fully depreciated over this period (i.e. no salvage value). Table 8 summarises the costs, and as can be seen the total annual cost for the complete plant (i.e. EDR plant and brine evaporator) amounts to R3,7 million, which is equivalent to 223 c/m^3 of product water. Table 8 also shows that the capital cost constitutes 55 per cent of the total amortized cost of the plant.

Table 8 TOTAL PLANT COST AMORTIZED OVER 20 YEARS ASSUMING A CONSTANT INTEREST RATE ON CAPITAL OF 15 PER CENT AND TOTAL PRODUCT FLOW RATE 53,2 ℓ/s

Component	Annual cost R000	Cost c/m³ produced	% of Total cost
Capital (EDR and brine evaporator)	2 013	122	55
Operating (EDR and brine evaporator)	1 671	101	45
Total	3 684	223	

7 CONCLUSIONS

7.1 The pilot plant study was a success in that it established the technical feasibility of the process; it served to establish the operating costs of a reasonably sized pilot plant; it highlighted areas of importance that must be considered when treating mine service waters; and it showed that the EDR process can, with the necessary precautionary measures, adequately desalinate low calcium sulphate/high sodium chloride mine water without fouling the membranes. Further evaluation of the EDR process on similar feed water is not considered necessary. Sufficient data were obtained during this study to enable a realistic operating cost estimate and conceptual design for a commercial scale plant to be made. Operating costs for a 46 ℓ/s (4 Mℓ/d) plant were estimated at 76,2 c/m³ of product produced (June 1990), with labour requirements for continuous operation being 8 and 2 hours per day for plant operators and a supervisor, respectively.

7.2 Under the optimized conditions of pretreatment (i.e. suspended solids removal to a turbidity of less than 1,0 NTU, and removal of iron in solution to less than 0,3 mg/ℓ), and with the addition of 15 mg/ℓ of sodium hexametaphosphate to the brine loop, the 1,6 ℓ/s EDR pilot plant was able to maintain a salt rejection of 80 per cent while operating at 84 per cent product water recovery on a feed with a low calcium sulphate and a high sodium chloride content.

7.3 The product water from the EDR unit contained about 80 per cent less salt than the feed water. Furthermore; it met the COMRO requirements for use in hydro-power systems after slight pH adjustment; it was found to be suitable for human consumption in terms of the SABS maximum allowable limits for total salt content in drinking water; and it satisfied the requirements for discharge to the environment as an effluent.

7.4 The membrane samples taken from the stack at the end of the 6 000 hour experimental period showed no signs of excessive membrane deterioration. Projected membrane life has been calculated at four years and seven years for the anion and cation membranes, respectively. These figures are consistent with those quoted in the literature.

7.5 The study highlighted the cartridge filters as a major contributing factor to the overall operating costs of the plant. It is envisaged that this cost for a commercial plant will be considerably less than that of the pilot plant because of the reduced supply cost, and the better feed water quality control. Nevertheless, this may be a problem strictly related to mine water composition, and needs consideration.

7.6 The pilot study showed that, contrary to the commonly held view within the desalination industry, the relatively high level of feed pretreatment would result in a cost of only about 12 c/m^3 (including the cartridge filter cost at 9 c/m^3), or about 19 per cent of the anticipated overall operating cost of a 46 ℓ/s (4 Mℓ/d) plant. The highest contributory factor to this cost would be the membrane replacement cost (48 per cent).

7.7 The study highlighted that this process is simple to operate and does not require highly skilled labour - operators from the mine's Gold Plant coped well with the operation of the plant after a short training period.

7.8 The mine water feed to the plant was supersaturated with respect to barium sulphate. Nevertheless barium sulphate scale was not noticeable in the brine loop at recoveries below 80 per cent. There are three possible reasons for this: firstly that the mass of barium sulphate potentially available for precipitation was relatively small (3,6 mg/ℓ at 75 per cent recovery); secondly that the polarity reversals prevented build-up of the precipitate; and thirdly that ion-pairing within the brine increased the solubility of the barium sulphate. Notwithstanding these observations, it is considered prudent to dose sodium hexametaphosphate (SHMP) into the brine loop at recoveries greater than 75 per cent to prevent any operational problems.

7.9 The level of manganese in the feed water could not be reduced by chemical oxidation methods. This potential problem can be overcome by the use of potassium permanganate-manganese 'green-sand' filters and/or separate electrode streams utilizing manganese-free water. Notwithstanding the elevated levels of manganese in the feed water during the pilot study, examination of membrane samples taken from the stack showed very little deterioration, indicating that fouling due to manganese was not significant.

7.10 The degree of water recovery from any desalination unit has to be optimized since it determines the flow rate of the concentrated brine from the unit, and thus the size and cost of the downstream brine treatment step. The capital cost of a 46 ℓ/s (4 Mℓ/d) EDR installation has been estimated at R9,1 million (1990), and that of an 8 ℓ/s (0,7 Mℓ/d) evaporator to dispose of the brine, at R3,5 million (1990). Together, a system such as this would recover over 98 per cent of the feed water as a high quality product (53,2 ℓ/s) at an amortized cost of 223 c/m^3, over 20 years at an interest rate of 15 per cent.

7.11 Due cognizance must be taken of feed water quality changes, particularly the calcium content. If this concentration is expected to increase significantly over the life of an installed full-scale EDR plant, then considerations for future calcium pretreatment in the form of softening would have to be taken into account.

(iv) Chlorine in the feed water

The EDR membranes can tolerate free chlorine levels of about 0,3 mg/ℓ continuously. They can withstand shock doses of up to 10 mg/ℓ, but higher concentrations will cause degradation of the membranes. Excess chlorine can be removed by an activated carbon absorber or by the addition of $NaHSO_3$ (sodium bisulphite), according to the following equation:

$$NaHSO_3 + HOCl \rightarrow NaHSO_4 + HCl \qquad (I.1)$$

(v) Suspended solids

Due to the nature of the process and the very small volumes in the flow compartments of the stack, it is important to maintain a high degree of suspended solids removal to ensure a practically solids-free feed water. This is achieved by means of dual media filtration followed by 10 μm cartridge filtration. High suspended solids levels in the feed can lead to localized blockages within the stack, which, if undetected, can lead to membrane destruction.

(vi) Biological fouling

Feed water should be clear of biological matter which could give rise to fouling in the stack.

APPENDIX II

METHODS FOR THE CHEMICAL OXIDATION OF IRON AND MANGANESE

1 CHLORINATION

This is a simple process requiring a chemical dosing system, a pH adjustment system and a retention tank. This process is normally only applied to iron-bearing waters since long contact times are required for the oxidation of manganese. Chlorine in the form of the sodium or calcium hypochlorite can be used and is required at a concentration of 0,62 mg chlorine per 1,0 mg of iron. A filtration process is required downstream of the chlorination step to remove the insoluble oxidation products. If the oxidation of manganese is to be achieved with this process then the pH needs to be maintained at approximately 8,5, and about 1,3 mg of chlorine will be required per 1,0 mg of manganese(7). Wong(7) recommends the use of $Ca(OH)_2$ as the pH adjustment chemical since the calcium assists in the destabalization of colloidal MnO_2. This is essential to achieve proper coagulation and flocculation of the colloidal particles, normally too small to be retained by the filter media. In the pilot study chlorination was initially used as the oxidation technique, but it failed because the pH was too low (approximately 6) and the retention time too short (1-2 minutes).

2 POTASSIUM PERMANGANATE

Pretreatment with $KMnO_4$ also requires simple process equipment, i.e. a chemical dosing system and a retention tank. Stoichiometrically 0,92 mg of $KMnO_4$ is required to oxidize 1 mg of iron, and the pilot plant study showed that this oxidation was very effective and completed within 1,5 minutes. According to Knocke et al.,(8) the oxidation of Mn(II) is also very effective, but requires about 5 minutes of contact time. This could explain why this method was not successful in reducing the manganese concentrations in the pilot plant study. Stoichiometrically

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APPENDIX I

CONVENTIONAL PRETREATMENT REQUIREMENTS FOR EDR

The problematic constituents of mine water feed to EDR plants and the methods typically used to overcome them include:

(i) Iron in the feed water

If the dissolved iron concentration is greater than 0,3 mg/ℓ pretreatment is required, usually in the form of oxidation to form an insoluble iron precipitate, followed by filtration. In most cases, iron fouling can be removed through a conventional 5 per cent hydrochloric acid 'cleaning in place' (CIP). However at pH values above 8,5 hydroxide films on the membrane surface could cause polarization which could lead to irreversible damage(6).

(ii) Manganese in the feed water

If the level of dissolved manganese is greater than 0,3 mg/ℓ, pretreatment by oxidation is also required. Although manganese fouling of the membranes is more difficult to remove than iron, it can be achieved by using a higher strength acid CIP i.e. 10 per cent hydrochloric acid(6). However, manganese will probably manifest itself in the electrode compartments of the stack where the formation of manganese dioxide can lead to the deterioration of the electrodes.

(iii) Barium and strontium

These ions are not usually found in high concentrations in mine waters, but nevertheless, may be problematic in the brine loop of the system where the saturation levels of $BaSO_4$ and $SrSO_4$ can be exceeded causing scale formation on the membranes. Normally, an antiscalant such as sodium hexametaphosphate (SHMP) is added to the brine loop to prevent scaling. Polarity reversal alone, without the use of SHMP, has been found to prevent scaling within the EDR stacks even at supersaturated salt levels(6).

1,92 mg of $KMnO_4$ is required per mg of managnese. Once again dual media anthracite/sand filters are required downstream to remove the oxidation products.

3 POTASSIUM PERMANGANATE - MANGANESE GREENSAND FILTRATION

Equipment for this process is similar to that required for chlorination except that $KMnO_4$ is the primary oxidizing agent, and the filter medium is different. Manganese greensand is a mineral capable of exchanging electrons, thereby oxidizing iron and manganese to their insoluble filterable states. Apart from being an oxidizing medium, the greensand also acts as a filter. Normally a 1 to 4 per cent solution of $KMnO_4$ is dosed continuously into the raw feed line prior to the filters. This reduces the amount of soluble iron and manganese going into the filters.

The greensand's oxidative capacity is limited, and needs to be regenerated with $KMnO_4$ after backwashing. The process has the advantage that it can act as a buffer - oxidizing any remaining soluble iron and manganese on the one hand, and utilizing any excess $KMnO_4$ to regenerate itself on the other. The stoichiometric amounts of $KMnO_4$ required are the same as those mentioned above. It is important with this process to maintain the pH between 7,5 and 9,0. A pH below 7,1 will lead to a deterioration of the filter medium(7). Under normal operation, replacement of the manganese greensand would be required periodically.

Manganese, because of its slower oxidation rate(7), is more difficult to remove from water than iron. This, coupled with the observations of Knocke <u>et al</u>.(8), that five minutes of contact is required for the oxidation of Mn (II), could explain why no changes in the concentrations of dissolved manganese were observed during the pilot study. The stoichiometry of the iron and manganese oxidation reactions are given by the following equations:

$$5Fe^{2+} + MnO_4^- + 8H^+ \rightarrow Mn^{2+} + 5Fe^{3+} + 4H_2O \qquad (II.1)$$

$$3Mn^{2+} + 2MnO_4^- + 2H_2O \rightarrow 5MnO_2\ (s) + 4H^+ \qquad (II.2)$$

APPENDIX III

BARIUM SULPHATE SATURATION LEVELS

Taking the brine ionic strength as being 0,16, calculated from an analysis carried out when the unit was operating at 84 per cent recovery (i.e. the brine ionic strength should not be more than this value at a lower recovery), the appropriate solubility product for barium sulphate can be obtained from a plot of Ksp($BaSO_4$) vs. ionic strength (Figure III.1), i.e. Ksp=$1,6 \times 10^{-9}$ mol^2/ℓ^2. Du Pont recommends that to avoid precipitation a plant should operate at an ionic product of barium sulphate of less than or equal to 0,8 of the Ksp value, i.e. $1,28 \times 10^{-9}$ mol^2/ℓ^2 in this specific instance.

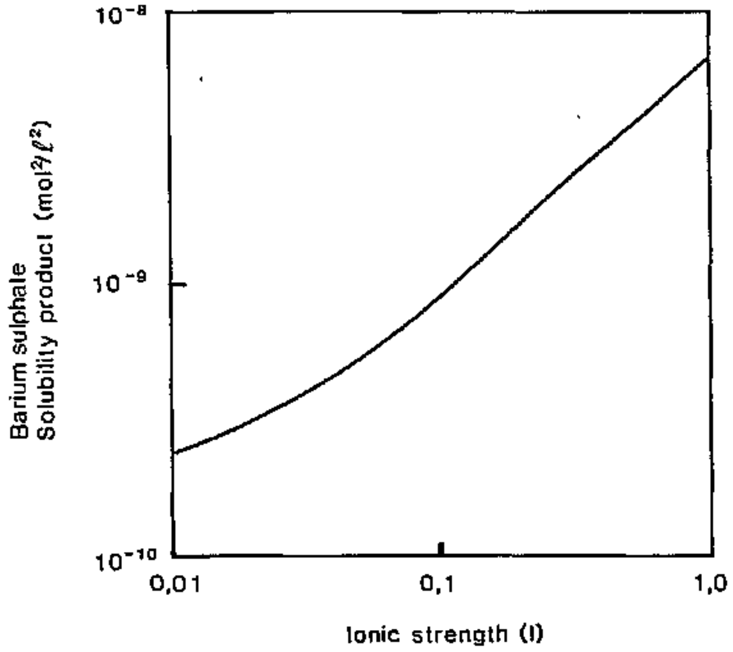

Figure III.1: SOLUBILITY PRODUCT FOR BARIUM SULPHATE VERSUS IONIC STRENGTH

Without taking into account the activities of the various species the degree of saturation of barium sulphate is defined as follows:

$$\text{Degree of saturation} = \frac{[Ba][SO_4]}{Ksp(BaSO_4)} \qquad (III.1)$$

where [Ba] = molar concentration of barium (mol/ℓ)
[SO_4] = molar concentration of sulphate (mol/ℓ)

Table III.1 shows the ionic product and the degree of saturation of barium sulphate in the feed water and in the brine loop at various water recoveries. The Concentration Factor (CF) is defined as the number of times the feed water is concentrated within the process resulting from product recovery. This can be written as an equation in the following form:

$$CF = \frac{1}{(1 - \frac{\text{Recovery (\%)}}{100})} \qquad \text{(III.2)}$$

From Table III.1 it is clear that the degree of saturation of barium sulphate in the brine stream increased as the water recovery increased. In fact, the feed water to the unit was already saturated with barium sulphate. At a recovery level of 71 per cent the brine loop had a degree of saturation of 20,6. Clearly at these concentrations the potential for barium sulphate precipitation must have been very high. The operational observations showed no evidence of barium sulphate precipitate in the brine loop or the stack until the recovery level was around 80 per cent. There are three possible reasons for this. Firstly, that the mass of barium sulphate able to precipitate is actually very small (about 100 g $BaSO_4$ per day in 40 m^3, at 75 per cent recovery); secondly, that precipitation did take place but that the polarity reversal prevented irreversible scaling until levels of about 80 per cent recovery were reached (a possible 150 g $BaSO_4$ per day in 33 m^3), and thirdly, that an ion-pairing effect in the solution increased the solubility of the salt.

Considering the ionic product alone, the Du Pont recommended limit of 0,8 of the Ksp for barium sulphate is exceeded at all levels of recovery. This recommended limit does not take into account the effect of the polarity reversal, and for the case of EDR as applied to Beatrix mine water, can be relaxed to a factor of about 35 times the Ksp for barium sulphate.

Table III.1 DEGREE OF SATURATION OF BARIUM SULPHATE IN EDR SYSTEM

Recovery	Concentration factor	Ba feed mg/ℓ	SO_4 feed mg/ℓ	*Ba brine (Cal.) mg/ℓ	*SO_4 brine (Cal.) mg/ℓ	Ion product $BaSO_4$ from * x 10^8 $mol^2/ℓ^2$	Degree of saturation	Potential of $BaSO_4$ precipitation mg/ℓ
0	-	0,5	76	-	-	0,3	1,8	0,3
71	3,4	0,5	76	1,7	258	3,3	20,6	2,4
75	4,0	0,6	83	2,4	332	6,1	38,1	3,6
80	5,0	0,6	78	3,0	390	8,9	55,6	4,6
84	6,3	0,6	65	3,8	410	11,7	73,1	6,0

* Calculated using the Concentration factor

APPENDIX IV

BACTERIAL COUNTS OF FEED AND PRODUCT WATER

Bacterial counts were made of the raw mine water feed and the product water from the EDR unit. Table IV.1 shows the results along with the typical limits for drinking water. The Total Plate count gives an indication of the overall bacterial content of the water since it considers all bacteria that are able to grow from a culture of the water sample. The Total Coliform count is much more specific and only considers the coliform bacteria (from humans). The latter figure can give an indication of sewage contamination of the water; however, the Faecal Coliform count (intestinal bacteria) is normally carried out specifically for determining sewage contamination.

Table IV.1 BACTERIAL COUNTS ON EDR FEED AND PRODUCT WATERS

Sample	Result	
	Total plate count	Total coliform count
EDR Feed	130/mℓ	1/100 mℓ
EDR Product	50/mℓ	2/100 mℓ
Typical drinking water specification	1000/mℓ	5/100 mℓ
SABS Domestic water specification (SABS 241-1984), maximum allowable limit	Not specified	* 5/100 mℓ

* if >0 <5 then some additional considerations are necessary. See SABS 241 - 1984.

It is evident from Table IV.1 that the Total Plate and Total Coliform counts are well within the limits for drinking water, but that the presence of coliforms makes regular sampling necessary, and if the water is intended for potable purposes it will have to be properly disinfected. The mine does chlorinate its service water underground by using calcium hypochlorite, but this dosing is not controlled well enough to ensure Total Coliform counts which are consistently zero.

APPENDIX V

SUMMARY OF MAJOR EVENTS AND PROCESS DIFFICULTIES DURING THE TEST PERIOD

By monitoring the electrical current across the EDR stack it is possible to detect a problem occurring within the stack. For example, as the resistance increases due to membrane fouling the current will decrease.

When membrane fouling was suspected within a stack, a 'stack probe' was carried out to isolate the area of high resistance within the stack. A stack probe involves measuring the potential across a number of membranes on opposite sides of the stack. Areas of high resistance (known as 'hot-spots') may be isolated. Stripping and cleaning of the affected section of the stack would normally be required, or in certain instances a Cleaning in Place (CIP) would have the required cleaning effect.

Figure V.1 shows the variation in 1st and 2nd stage current as well as the variation in product conductivity. Points (A) through (T) indicate times in the test work when a problem was encountered or a change was made to the operating conditions, and the description of each of these is given below.

(A)	458 hours	Sludge was accidentally pumped from the surface hot well to the pilot plant site, resulting in a blockage in the stack. The stack was opened and the 1st and 2nd hydraulic stages were cleaned. Two pairs of membranes were removed as samples.
(B)	1 200 hours	CIP carried out with 5 per cent hydrochloric acid for one hour.
(C)	1 470 hours	Stack spacer in 3rd hydraulic stage displaced due to excess stack pressure; replaced. Two anion membranes were taken as samples.
(D)	1 640 hours	Stack spacer in 3rd hydraulic stage displaced again;

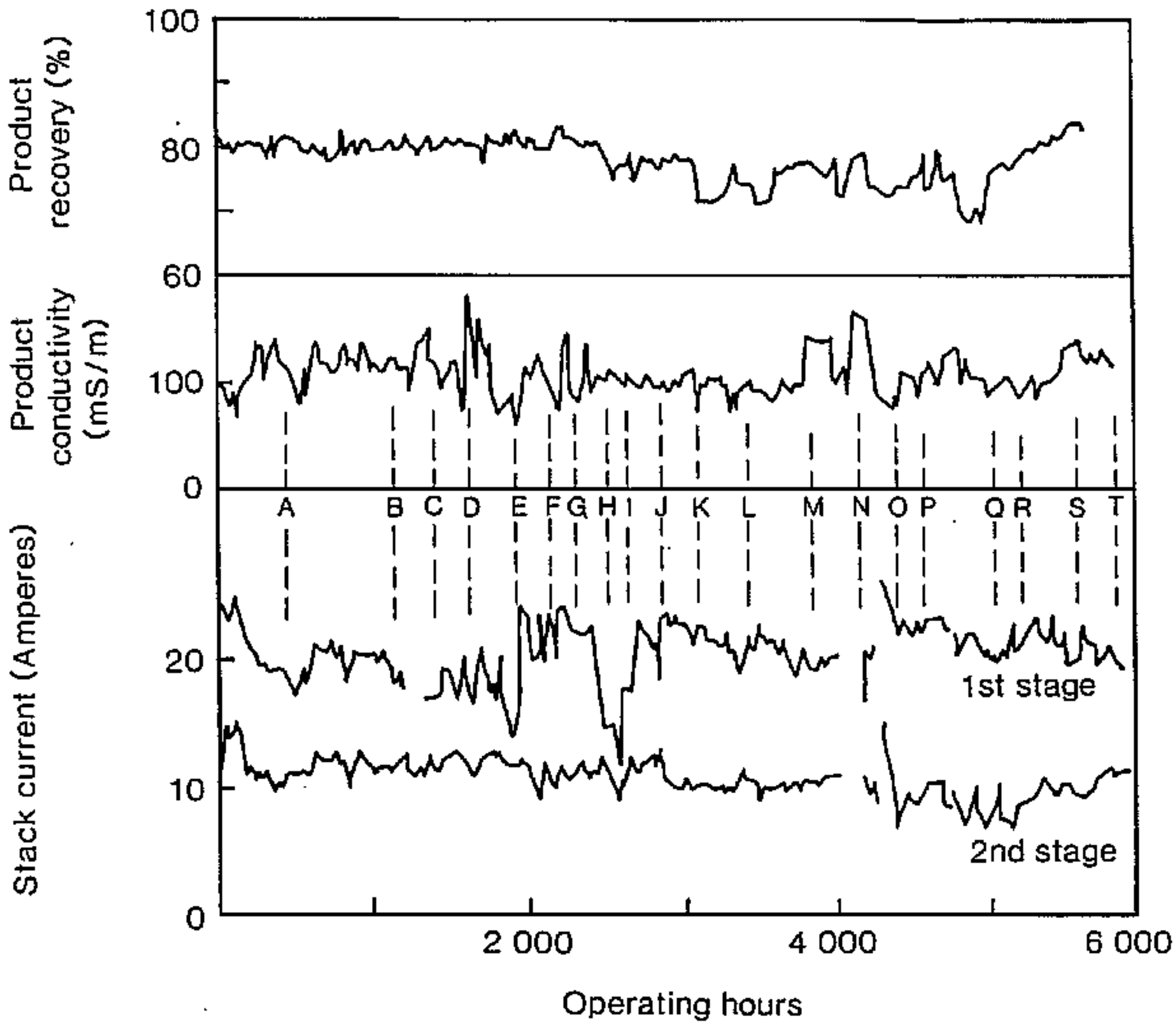

Figure V.I: VARIATION OF PRODUCT CONDUCTIVITY, STACK AMPERAGE AND PRODUCT RECOVERY FOR THE WHOLE EXPERIMENTAL PERIOD

	repaired. Calcium hypochlorite dosing commenced.
(E) 1 950 hours	Fall in current of first electrical stage. Variation in current during a cycle observed. Stack probe indicated small 'hot-spots'. Stack was opened and a sample of anion and cation membranes were taken from 1st hydraulic stage. CIP carried out. Hot-spots reduced in size.
(F) 2 200 hours	Two CIP's carried out. More sensitive analytical method for determining dissolved manganese in feed water introduced.
(G) 2 335 hours	Stopped calcium hypochlorite dosing and started sodium hypochlorite dosing.
(H) 2 520 hours	Suspected trouble with the stack because of the low current through the 1st stage. Located 'hot-spots' in 1st and 2nd electrical stages of the stack. Stripped the stack completely and replaced 9 membrane pairs in the 4th hydraulic stage, 7 pairs in the 3rd, 3 pairs in the 2nd, and 10 pairs in the 1st hydraulic stage. The spacers contained a lot of scale (which was analysed to be predominantly - 44 per cent - $BaSO_4$, with smaller amounts of silica and strontium). After reassembly of the cleaned stack, a CIP was carried out.
(I) 2 800 hours	Drifting noticed in the current and product water quality during a cycle, indicating that fouling of the membranes was occurring.
(J) 2 900 hours	CIP was carried out.
(K) 3 230 hours	CIP was carried out and the recovery of the unit reduced.

(L) 3 400 hours	A stack probe revealed a 'hot-spot' at the bottom of the 1st hydraulic stage. The stack was stripped and two cation membranes were replaced. A certain amount of $BaSO_4$ scale was noticeable. Potassium permanganate dosing was introduced to the feed stream to oxidise the iron and manganese present. Sodium hypochlorite dosing stopped.
(M) 3 880 hours	White flakes of $BaSO_4$ evident in the brine stream, - recovery reduced slightly, from 75 to 70 per cent.
(N) 4 080 hours	CIP carried out. Flocculant dosage increased by 50 per cent, from 1 mg/ℓ to 1,5 mg/ℓ.
(O) 4 200 hours	Stack stripped. Manganese dioxide presence evident in the electrode compartments. Electrodes replaced, heavy cation membranes and spacers replaced on top of stack. Motor operated valves removed for maintenance and then replaced.
(P) 4 600 hours	CIP carried out.
(Q) 5 060 hours	'Hot-spot' located in stack at bottom of 2nd hydraulic stage. Stack stripped and hot-spot removed - the scale had the appearance of $BaSO_4$ - six membranes cleaned, none replaced.
(R) 5 350 hours	Sodium hexametaphosphate (SHMP) dosing into the brine loop introduced. Recovery gradually increased from this point.
(S) 5 720 hours	CIP carried out. No sign of scale formation. Recovery increased further to 85 per cent where it was held for the last 200 hours of operation.

(T) 5 908 hours End of test work. Stack opened. No sign of scale formation. Membrane samples removed for final analysis.

As Figure V.1 shows, even with the fairly large number of events associated with the plant, the product water quality only deteriorated significantly between (C) and (J), and (N) and (O). Both of these problem phases resulted from problems in the stack: firstly, the fouling of the membranes by iron present in the feed water and secondly, because of barium sulphate scale formation in the stack as well as the formation of manganese dioxide in the electrode compartments. The latter resulted in the deterioration of two electrodes which had to be replaced (4 200 hours).

Figure V.I also shows the product recovery at which the plant was being operated. It is interesting to note that with all the pretreatment systems in operation, including the dosing of SHMP, the increased recovery of the unit in the last 900 hours to around 85 per cent did not affect the quality of the product water. It was anticipated that had the testwork continued there would have been recurring problems of electrode deterioration due to the manganese still present in the feed. However, scaling and fouling of the stack would not have been expected since the feed stream had been suitably treated for both iron and barium. Appropriate means of dealing with the presence of manganese in the feed water is dealt with in Section 6.1.

APPENDIX VI

DETAILS OF CONCEPTUAL GROUND ARRANGEMENT FOR A 46 ℓ/s EDR INSTALLATION

The following conceptual design is based on a recovery of 85 per cent; if this figure were to be lowered, the overall throughput of the plant would be greater and the dosing and feed storage vessels would need to be increased in size. The ground arrangement is shown in Figure 9 in the report.

PRETREATMENT

Removal of iron and manganese and bulk of suspended solids

The problems associated with iron and manganese concentrations in the feed water of more than 0,3 mg/ℓ have been dealt with earlier in this report i.e. iron can cause fouling in the stack and manganese can form manganese dioxide in the electrode compartments, which causes irreparable damage to the electrodes themselves. There are two possible approaches to ensure the acceptable limiting concentrations of these important ions:

(a) chemical removal of both the iron and manganese, or

(b) chemical removal of the iron only, and the use of a manganese-free water stream as the electrode flush water.

For the purpose of this conceptual design the former approach has been chosen. Once again there are various methods that could be used to accomplish the chemical removal of iron and manganese (Appendix II). The one that has been considered here is the potassium permanganate oxidation - manganese greensand filtration method. This can be carried out in three stages:

(i) dosing of potassium permanganate into the feed line,

(ii) pH adjustment to maintain the pH between 7,5 and 9,0, and

(iii) filtration through manganese greensand filters where the soluble manganese will be oxidized, and the products of oxidation and the bulk of the suspended solids will be removed.

Assuming an iron content in the raw feed water of 2 mg/ℓ, then about 2 mg/ℓ of $KMnO_4$ would be required(7). To accomplish this the $KMnO_4$ dosing system would require a 2 m^3 tank (1,2 m diameter x 2 m height), a dosing solution of 5 g $KMnO_4$/ℓ, a dosing rate of about 0,02 ℓ/s, and would require refilling once a day.

pH control would be carried out immediately after the $KMnO_4$ dosing step in an agitated vessel to ensure the oxidation reaction occurs under optimal pH conditions. Both acid and caustic dosing facilities would be required to cope with the varying raw feed water pH and it is envisaged that a 50 m^3 tank (5 m diameter x 3 m high) would be required.

Three manganese greensand filters each with a capacity of around 23 ℓ/s (2 000 m^3/d) would be required to satisfy the manganese oxidation and suspended solids removal step requirements. Each of these filters would be 3 m in diameter and about 3 m in height.

The pretreated feed would be stored in a one-hour storage tank with a volume of about 200 m^3 (10 m diameter x 3 m height).

CARTRIDGE FILTRATION

Cartridge filtration would take place immediately prior to the feed water entering the EDR stacks. The feed water would have to be pumped from the pretreated feed tank to the cartridge filter housing which would be under cover in the main building. The nominal size of the cartridge filters used depends on the type of suspended matter encountered in the mine water e.g. quartz particles, clays etc. Typically about a 10 μm nominal sized filter cartridge would be required to reduce the turbidity of the feed water to below 1 NTU.

EDR PLANT

For a product water capacity of 46 ℓ/s eighteen membrane stacks would be required. These, together with the connecting pipework would be housed in a building with a floor area of about 360 m^2, including a stack stripping and cleaning area. An overhead crane facility would be required to facilitate easy transfer of the stacks to and from the cleaning area.

The sodium hexametaphosphate (SHMP) dosing system would require a 2 m^3 dosing tank (1,2 m diameter x 2 m height), a dosing solution of 10 g SHMP/ℓ, and a dosing rate of about 0,02 ℓ/s to achieve a concentration of 20 mg/ℓ in the brine. This tank would require refilling each day and would be housed in the main EDR building.

PERSONNEL REQUIREMENTS

The pilot plant study showed that although the EDR process is a modern technology, the plant could operate with the minimum of supervision. In the pilot study four operators with reduction works experience (Patterson grading B3) were trained to operate and take readings on the EDR plant; they coped adequately with this task.

It is envisaged that a commercial scale plant would only require about eight hours of operator attention per day i.e. 2,7 hours per eight hour shift. This would involve checking over the entire plant, recording a few critical parameters, and possibly taking water samples for analysis. A supervisor would be required to check on the performance of the plant from the operator readings on a daily basis and initiate any maintenance work.

POST TREATMENT OF EDR PRODUCT WATER

It is envisaged that a one hour intermediary storage vessel for the product water would be required; this would need a capacity of about 175 m^3 (10 m diameter x 3 m height). At this point it might be necessary to increase the pH of the water as well as its alkalinity to make it suitable for down stream uses. Both could be accomplished by the use of lime and carbon dioxide which will result in a less corrosive water, i.e. a water with a Langlier Saturation Index of zero or slightly positive.

APPENDIX VII

DETAILS OF OPERATING COST ESTIMATE (JUNE 1990) FOR A 46 ℓ/s EDR INSTALLATION

The overall operating costs are shown in Table VII.1, and the break-down of each cost is given below.

Table VII.1 ESTIMATED OPERATING COSTS (JUNE 1990) FOR A 46 ℓ/s (4 Mℓ/d) COMMERCIAL SCALE EDR PLANT

Category	Details	Cost (c/m^{3}produced)
Electrical	2,4 kWh/m^3 at 9 c/kWh	21,6
Pretreatment	Flocculant 0,8 c/m^3 $KMnO_4$/Greensand filter 1,3 c/m^3 SHMP 1,0 c/m^3	3,1
CIP	HCl acid wash every 1 000 hours	0,5
Cartridge filters	Based on cheaper supply cost and commercial plant operational improvements	9,0
Labour	4 operators at R2 500/month (third of time) 1 supervisor at R4 600/month (quarter of time)	3,8
Maintenance	18 membrane stacks	0,9
Stack cleaning	6 labourers for 16 hours every 5 000 hours per 3 stacks at R3,50/hour	0,1
Membrane stack replacement	R30 000/stack per year	37,2
Total operating cost		76,2

ELECTRICITY COST

This cost is based on a power consumption rate of 2,4 kWh per m^3 of product water, as obtained during the pilot study, at a cost of 9,0 cents per kilowatthour. Table VII.2 indicates the anticipated relative electrical consumption when operating a commercial scale plant under different conditions of feed water TDS, to produce the same quality product water(6). Note that the consumption figures are lower than the 2,4 kWh/m^3 measured during the pilot plant study, making the estimated operating costs in Table VII.1 somewhat conservative.

Table VII.2 ELECTRICITY CONSUMPTION AT VARIOUS FEED CONCENTRATIONS

Feed TDS (mg/ℓ)	Electricity consumption (kWh/m^3) To produce product 500 mg/ℓ TDS	To produce product 1 000 mg/ℓ TDS	To produce product 1 500 mg/ℓ TDS
2 000	1,40	1,15	0,95
3 000	1,85	1,60	1,40
4 000	2,30	2,05	1,85

PRETREATMENT COST

The figures obtained during the pilot study were used as estimates for the commercial scale plant - approximately 3 cents per m^3 of product water produced. This includes the cost for flocculant addition, chemical dosing to reduce the iron and manganese concentrations to acceptable levels, and SHMP dosing in the brine loop to prevent barium sulphate scale formation.

There are various ways to accomplish the chemical oxidation of iron and manganese (see Appendix II), but regardless of which pretreatment step (or combination of pretreatment steps) is chosen, it is felt that a total cost of 3,1 c/m^3 should cover these operating costs as well as flocculant and SHMP addition.

CLEANING IN PLACE (CIP)

The hydrochloric acid CIP washes of full-scale plants were estimated to cost the same, or less, than the cost incurred in the pilot study. To keep the figure conservative though, it is estimated that a CIP would be carried out every 1 000 hours and would cost around 0,5 cents per m^3 product water.

CARTRIDGE FILTERS

It is possible that in a commercial installation this cost would be much lower than that experienced in the pilot scale operation, since bulk buying of cartridge filters would be cheaper, and better control of the feed water quality (i.e. optimized pretreatement, in-line turbidity measurement and good flow rate control) on a commercial plant would ensure longer filter runs. Information obtained from presently installed commercial scale plants puts the replacement cost of cartridge filters around 5 cents per m^3 product water (1987). Bearing in mind that these installations do not treat mine water, that the colloidal properties of the waters could be very different, and that very short fouling times were obtained during the pilot study, a cost of about 9 c/m^3 product water is considered a realistic, although conservative, estimate.

LABOUR

It was clear from the pilot study that continuous supervision of an EDR plant is not necessary (Appendix VI). Even with a large commercial installation it is envisaged that personnel need only spend one-third of their time at the plant, on a three by eight hour shift basis. Such a system would require four operators, and it is felt that a supervisor would also be required to check the results from the plant once a day (about a quarter of his time). The equivalent c/m^3 figures presented in Table VII.1 are based on an assumed operator cost of R2 500/month and a supervisor cost of R4 600/month.

MAINTENANCE

The EDR process operates at relatively low pressures, and PVC piping is used throughout. The polarity and flow reversal system requires a number of motor-operated and solenoid valves. With the reversals taking place typically every 20 minutes, these valves are subject to wear over a long period of time. Other areas that may require attention are the power distribution system (polarity switches, DC converters and stepdown transformers) as well as the pumps and instrumentation. The cost is estimated at around R2 300 per three stack train per year, since all three stacks in a train would operate off the same set of valves.

PHYSICAL STACK CLEANING

It is estimated that each stack would require complete stripping and cleaning every 5 000 hours. This regular cleaning would result in membrane replacement cost savings. Normally a three stack train would be done simultaneously, and with two labourers per stack it would take two days to complete the job of disassembly, physical cleaning and reassembly. For such a job unskilled labour is required. The figures in Table VII.1 are based on an assumed unskilled labour cost of R3,50/h for an eight hour day.

MEMBRANE STACK REPLACEMENT

Numerous membrane samples were taken during the pilot plant study and sent to Ionics, USA for analysis. The analysis showed no excessive membrane deterioration, and projected membrane life has been calculated at four years and seven years for the anion and cation membranes, respectively. These figures are typical of membrane lives as quoted in the literature.

Table VII.3 gives details concerning expected life and current replacement costs for stack components.

Table VII.3 EDR MEMBRANE STACKS - EXPECTED LIFE AND REPLACEMENT COST (JUNE 1990)

Component	Number per stack	Lifespan (years)	Current replacement cost/unit (Rand)*
Electrodes+	2	1	3 000
Anion membranes	500	4	80
Cation membranes	500	7	75
Spacers	1 000	15	96

* Based on US $1.00 = R2,65.

+ Incorporating a separate electrode flush stream utilizing manganese free water (see Appendix VI) could extend the electrode life to possibly two years, but for the purposes of this report the more conservative estimate of one year has been used.

These figures are based on manufacturers' experience as well as on published data, and suggest a replacement cost of around R30 000 per stack per year.