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Investigation into alternative water treatment technologies for the treatment of underground mine water discharged by Grootvlei Proprietary Mines Ltd into the Blesbokspruit in South Africa

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Abstract

Grootvlei Proprietary Mines Ltd is discharging between 80 and 100 Ml/d underground water into the Blesbokspruit. This water is pumped out of the mine to keep the underground water at such a level as to make mining possible. The water is of poor quality because it contains high TDS levels (2700–3800 mg/l) including high concentrations of iron, manganese, sulphate, calcium, magnesium, sodium and chloride. This water will adversely affect the water ecology in the Blesbokspruit, and it will significantly increase the TDS concentration of one of the major water resources if not treated prior to disposal into the stream. Therefore, alternative water desalination technologies were evaluated to estimate performance and the economics of the processes for treatment of the mine water. It was predicted that water of potable quality should be produced from the mine water with spiral reverse osmosis (SRO). It was demonstrated that it should be possible to reduce the TDS of the mine water (2000–2700–3400–4500 mg/l) to potable standards with SRO (85% water recovery). The capital costs (pretreatment and desalination) for a 80 Ml/d plant (worst-case water) were estimated at US$35 M. Total operating costs were estimated at 88.1c/kl. Brine disposal costs were estimated at US$18M. Therefore, the total capital costs are estimated at US$53 M. It was predicted that it should be possible to produce potable water from the worst-case feed water (80 Ml/d) with the EDR process. It was demonstrated that the TDS in the feed could be reduced from 4178 to 246 mg/l in the EDR product (65% water recovery). The capital costs (pretreatment plus desalination) to desalinate the worst-case feed water to potable quality with EDR is estimated at US$53.3M. The operational costs are estimated at 47.6 c/kl. Brine disposal costs were estimated at US$42M. Therefore, the total capital costs are estimated at US$95.3 M. It was predicted that it should be possible to produce potable water from the mine water with the GYP–CIX ion-exchange process. It was demonstrated that the feed TDS (2000–4500 mg/l) could be reduced to less than 240 mg/l (54% water recovery for the worst-case water). The capital cost for an 80 Ml/d ion-exchange plant (worst-case water) was estimated at US$26.7M (no pretreatment). Operational costs were estimated at 60.4 c/kl. Brine disposal costs were estimated at US$55.1 M. Therefore, the total desalination costs were estimated at US$81.8 M. The capital outlay for a SRO plant will be significantly less than that for either an EDR or a GYP–CIX plant. The operating costs, however, of the RO plant are significantly higher than for the other two processes. Potable water sales, however, will bring more in for the RO process than for the other two processes because a higher

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water recovery can be obtained with RO. The operating costs minus the savings in water sales were estimated at 17.2; 6.7 and US$8.6 M/y for the RO, EDR and GYP–CIX processes, respectively (worst case). Therefore, the operational costs of the EDR and GYP–CIX processes are the lowest if the sale of water is taken into consideration. This may favour the EDR and GYP–CIX processes for the desalination of the mine water.

Keywords: Underground mine water; Treatment technologies; Reverse osmosis; Electrodialysis reversal; Ion-exchange; Water quality; Brine disposal; Treatment costs

1. Introduction

Grootvlei Proprietary Mines Ltd is presently discharging between 80 and 100 megalitre (Ml) underground water per day into the Blesbokspruit. This water is pumped out of the mine to keep the underground water at such a level as to make mining possible. The water is of poor quality because it contains high concentrations of dissolved salts (2700–3800 mg/l TDS) including high concentrations of iron (150–300 mg/l), manganese (8 mg/l), sulphates (1700–2300 mg/l), calcium (340–240 mg/l), magnesium (150–200 mg/l), sodium (270–320 mg/l), and chlorides (190–240 mg/l). This water will adversely affect the water ecology of the Blesbokspruit, and it will significantly increase the TDS concentration of some of the water resources in South Africa if not treated prior to disposal into the Blesbokspruit. Therefore, the Department of Water Affairs and Forestry (DWAF) has set water discharge quality goals to the mine that must be met as a function of time to protect the water environment from adverse effects [1].

Phase 1 of the water discharge permit conditions to the mine makes provision for the discharge of mine water into the Blesbokspruit with a pH between 6.5 and 9, suspended solids (SS) and total iron of less than 60 and 20 mg/l, respectively. Phase 2 of the discharge permit conditions makes provision for the discharge of water into the Blesbokspruit with pH of the water between 6.5 and 8.4; and that the electrical conductivity, sodium absorption ration, and SS should have maximum values of 207 mS/m; 2.5; and 25 mg/l, respectively. Phase 3 of the discharge permit conditions stipulates that the water should be treated to potable standards (40 mS/m) prior to disposal. Therefore, the water must be desalinated prior to disposal.

Reverse osmosis (RO), electrodialysis (ED) and ion-exchange (IX) were identified as technologies which should have the potential to desalinate the mine water effectively [2–4]. These technologies were therefore evaluated for the desalination/concentration of the mine water. The emphasis of the study was on process performance, with special interest on the economics of the process for treatment of the mine water [5].

2. Performance and economics of desalination technology for treatment of Grootvlei mine water

2.1. Reverse osmosis

A simplified process flow diagram for RO treatment of the mine water is shown in Fig. 1. The proposed plant is seen to consist of four sections: (a) pretreatment, (b) desalination, (c) post-treatment, and (d) brine disposal.

2.1.1. Pretreatment

Aeration tank: The raw water from the mine is fed into an aeration tank. The purpose of the aeration tank is to oxidize any ferrous ions into ferric ions both by aeration and by the addition of sodium hypochlorite or chlorine gas.

Lime and soda ash addition tank: Lime and soda-ash are dosed from a bulk handling facility into an agitated reaction tank. The purpose of the

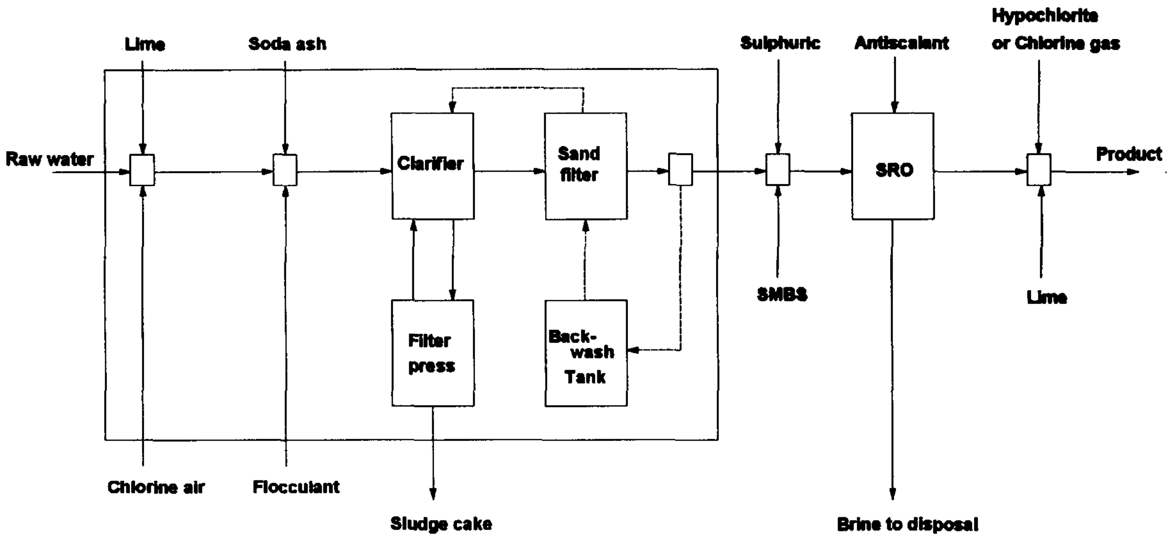

Fig. 1. Simplified process flow diagram of the RO process for desalination of the mine water.

lime and soda-ash is to precipitate calcium as calcium carbonate. The addition of soda-ash will also assist in the precipitation of iron, manganese and aluminium. Silica and barium will also be reduced by soda-ash addition.

Flocculation tank: An agitated flocculation tank was included upstream of the clarifier in order to achieve proper flocculation and particle agglomeration, so that good settling is achieved in the clarifier. At this stage the type and dosing levels of a suitable flocculant are not known and will have to be determined experimentally. A dosing rate of 20 mg/l at a typical cost has been assumed for cost estimate purposes. Lower actual dosing rates of 1 to 5 mg/l may be possible at optimum conditions, which would not significantly affect the total operating cost. It is expected that dosing rates in the order of 20 mg/l will be possible with sludge recycle from the clarifier, and should prevent the need for dosing rates in the range of 100–200 mg/l. At these levels, however, the total operating cost will increase substantially.

Clarifier: The bulk of the particles settle in the clarifier. A scraper mechanism is installed in the clarifier in order to collect the clarifier underflow efficiently in a separate underflow tank or compartment. A surface skimmer is provided in the clarifier to collect any surface scum in a collection tank. At this stage no particular design is proposed, since the final clarifier design will have a relatively small impact on the total capital cost estimate which will be absorbed within the ±30% accuracy range.

Sulphuric acid addition: A supernatant tank for pH correction after the clarifier is included so that the pH after the clarifier may be reduced by the dosing of sulphuric acid. The pH in this tank is controlled at pH 7.5–8.0 to maximize water recovery in the SRO membrane plant. This is a chemical requirement, according to the composition of the raw water, and not a function of the membrane type.

Sand filters: Sand filters are provided in order to remove the bulk of the particles which are present in the clarifier supernatant or overflow. The sand filters are back-washed using backwash water and simultaneous air scouring. Special media such as green sand may be used for the removal of residual levels of iron and manganese. It is important to recycle the backwash water for treatment in the clarifier in order to achieve the highest possible water recovery in the pretreatment section.

Sludge dewatering plant: Sludge from the aeration tank and clarifier is collected in a sludge collection tank. The sludge is dewatered using a filter press. The filter cake from the filter press is transported to a separate sludge (filter cake) disposal facility. The filtrate is returned to the aeration tank.

Membrane plant feed tank: Filtered water from the sand filters flow into a membrane plant feed tank. The purpose of this tank is to supply storage for sand filter backwash water and to allow sufficient residence time between the pretreatment plant and the SRO plant for proper control of both plants. If the quality of the filtered water is unsatisfactory for the SRO plant, or if the SRO plant is not available for a short duration, the filtered water can be recycled to the aeration tank.

2.1.2. Desalination

An SRO plant consisting of modular membrane units is included. The SRO plant is preceded by anti-scalant dosing and two-stage cartridge filtration to remove traces of SS which may break through the sand filters. A cleaning-in-place (CIP) system is also provided for chemical cleaning of the membranes.

2.1.3. Post-treatment

The permeate or product water from the SRO plant flows into a permeate storage tank. Lime (for final pH correction) and sodium hypochlorite or chlorine gas (for disinfection), if required, are dosed to the permeate storage tank.

2.1.4. Brine disposal

The brine or concentrate from the SRO flows into a brine storage tank. The brine is pumped from the brine storage tank to the evaporation ponds for disposal. Forced evaporation by thermal means has been identified as an alternative at the expense of higher capital and operating costs.

2.1.5. Water recovery

The recovery of water with RO, in the case of the probable- and worst-case waters (Table 2), is limited to approximately 50% with the use of anti-scalant, due to calcium sulphate scaling. The given chemical analyses of different lime and soda-ash softening experiments were used to determine the maximum attainable recovery with the use of anti-scalant for each case [5]. These are summarized in Table 1.

The lime to pH9 and soda-ash to pH10 softening option was adopted in subsequent work of this preliminary investigation, since this option provides the highest recovery of 85% at acceptable acid and anti-scalant consumption. As mentioned previously, the treatment option with the highest possible recovery or lowest brine volume results in the lowest total treatment cost.

2.1.6. Estimated compositions of process streams

The estimated compositions of the feed streams (best, probable, and worst cases) and the compositions of the RO permeate and brine are shown in Table 2. Water of potable quality can be produced in all three cases. It is noted that RO product water with very low TDS (approximately 100 mg/l and less) can be produced. Therefore, some degree of mixing of the RO product water with the RO feed will be possible. It is also interesting to note that brine with a relatively high TDS is produced in the probable and worst cases. The sodium concentration of the brine is high, and it will be necessary to dispose the brine safely.

Table 1
Effect of softening treatment on maximum water recovery

	pH adjustment	Max. recovery, %	Comments
Lime to pH 8	None	48	Low recovery
Lime to pH 9	None	35	Low recovery
Soda ash to pH 9	7.5	65	Low recovery
Soda ash to pH 10	6.5	85	High acid consumption
Soda ash to pH 10.5	6.5	85	Very high acid consumption
Lime to pH 8 and soda ash to pH 10	8	80	Acceptable acid consumption
Lime to pH 9 and soda ash to pH 10	8	85	High recovery, acceptable acid consumption

Table 2
Estimated composition of process streams (mg/l)

Component	Best case			Probable case			Worst case		
	RO feed	RO perm	RO brine	RO feed	RO perm	RO brine	RO feed	RO perm	RO brine
Ca	20	0.4	132	38	0.7	262	49	0.8	326
Mg	4.5	0.1	30	6.5	0.1	45	11	0.2	73
Na	650	16	4,327	1,250	35	8,756	1,450	35	10,217
K	21	1.0	135	23	1.2	154	35	1.5	227
pH	10	6.1	8.2	10	6.5	8.3	10	6.0	9
Alkalinity	5	2.1	325	254	20	3,574	178	12	2,422
Cl	178	7.0	1,317	178	15	2,750	272	14	2,838
SO_4	1,207	24	7,587	2,328	41	12,730	2,887	49	16,920
F	0.1	0	0.6	0.2	0	1.3	0.3	0	2.0
SO_2	1	0	5.5	3	0.1	20	5	0.1	20
TDS	2,100	51	13,860	4,130	113	28,292	4,920	113	32,590

2.1.7. Flow rates

The product and brine flow rates for the two capacity options are shown in Table 3. A water recovery of approximately 85% would be possible. Therefore, the brine would comprise 15% of the treated water.

Table 3
Product and brine flow rates at various plant capacities

Feed, Ml/d	Product, Ml/d	Brine, Ml/d
80	68	12
45	38.25	6.75

2.1.8. Cost estimates

Capital cost estimates: The class 1 capital cost estimates (±30%) for the pretreatment and membrane desalination plants are shown in Table 4 and Fig. 2 (1996 prices were used). The pretreatment plant comprises approximately 44%

Table 4
Capital cost estimates for pretreatment and a membrane desalination plant

Item	Cost (US$, M) Capacity, 80,000 m^3/d		Cost (US$, M): Capacity, 45,000 m^3/d		
	Probable	Worst	Best	Probable	Worst
Mechanical equipment pretreatment plant	20.7	9.6	5.6	6.2	6.9
Mechanical equipment desalination plant	11.3	12.5	7.3	8.1	8.9
Electrical and C&1 hardware	4.1	4.5	2.6	2.9	3.2
civil and structural	2.8	3.1	1.8	1.2	2.2
Engineering services (includes engineering services and design for process, mechanical, civil and structural, electrical and C&I	3.6	4.0	3.0	2.6	2.8
Preliminary and general	1.2	1.3	0.75	0.89	0.93
Total	**31.8**	**35.0**	**20.4**	**22.7**	**24.0**
Pretreatment plant turn-key	13.9 (12.2)	15.2 (13.4)	8.9	9.8	10.9
Desalination plant turn-key	17.0 (16.0)	19.8 (17.4)	11.6	12.9	14.1
Total	**31.83 (28.2)**	**35.0 (30.9)**	**20.4**	**22.7**	**24.0**

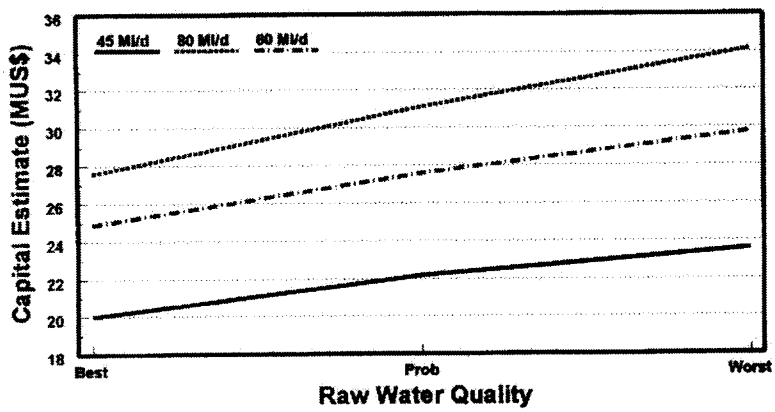

Fig. 2. Capital estimate trends for different capacities and raw water qualities (60 Ml/d curve interpolated).

of the total plant costs. The capital costs of the RO desalination plants are 19.98 (16.67) and US$19.78 M (US$17.44 M) for the probable and worst cases for 80 and 65 Ml/d plants, respectively. The RO capital costs for the best-, probable- and worst-case feed waters in the 45 Ml/d plant are 11.56, 12.87 and US$14.07 M, respectively.

Operating cost estimates: The class 1 total operating estimates (±30%) are summarized in Table 5 and Fig. 3, based on a feed capacity of 45 Ml/d and a water recovery of 85%. Since the operating estimates are expressed as a specific cost per cubic metre product, the specific costs for the 80 and 65 Ml/d feed capacity options will be the same. The total operating costs were 43.0,

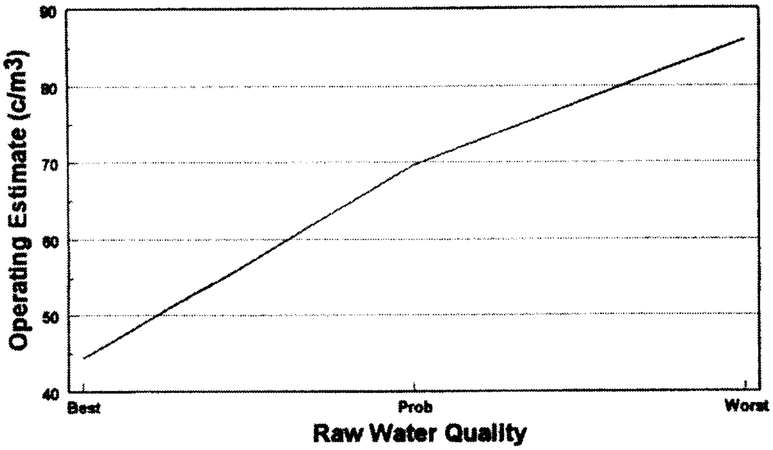

Fig. 3. Operating estimate trend for different raw water qualities.

Table 5
Total operating costs (basis: 45,000 m^3/d @ 85% recovery)

Element	Operating cost (c/m^3)		
	Best	Probable	Worst
Power and chemicals	34.5	61.2	78.4
Equipment and membrane replacement	6.2	6.3	6.4
Labour	3.3	3.3	3.3
Total	43.0	70.8	88.1

Table 6
Estimate of reverse osmosis brine disposal capital costs (US$, M)

Disposal type	Brine, Ml/d		
	6.75[a]	9.75[b]	12[c]
Evaporation ponds (unlined)[d,e]	7.6 (8.2)	11.1 (12.0)	13.6 (14.7)
Evaporation ponds (lined)[d,e]	32	46.2	57.1
Enhanced spray evaporation[f]	10.2	14.7	18.0
Ocean disposal			75.3

[a]Based on 85% recovery (45 Ml/d feed.
[b]Based on 85% recovery (80 Ml/d) feed.
[c]Based on 85% recovery (65 Ml/d) feed.
[d]Based on US$1222.2/kl brine to be evaporated per day.
[e]Based on recently determined data (November, 1995).
[f]Based on US%1500/kl brine to be evaporated per day (1US$ = R4.50).

70.8 and 88.1 c/kl for the best, probable and worst cases, respectively.

Brine disposal costs: Evaporation ponds (unlined and lined), enhanced evaporation spray and ocean disposal were considered for brine disposal. An estimate of the brine disposal capital costs are given in Table 6.

Ocean disposal (US$75.3 M) and disposal into lined evaporation ponds appear to be the most costly brine disposal options (57.11, 46.22 and US$32 M for 12, 9.75 and 6.75 Ml/d brine, respectively). However, it appears that brine disposal into enhanced evaporation spray and into unlined ponds will be less expensive. Brine disposal costs were determined at 18, 14.67 and US$10.22 M for brine flows of 12, 9.75 and 6.75 Ml/d, respectively (enhanced evaporation spray). Brine disposal costs in unlined ponds were determined at 13.56, 11.11 and US$9.56M for brine flows of 12, 9.75 and 6.75 Ml/d when data were used from a recently determined pricing exercise for brine disposal [5]. Therefore, brine disposal into the enhanced evaporation spray ponds appears to be a little more expensive than brine disposal into unlined ponds. However, the enhanced evaporation spray ponds are lined, and it will, therefore, be more suitable for storage of hazardous chemicals. The area requirements for enhanced evaporation spray ponds are also significantly less than for conventional ponds.

Operational costs for conventional evaporation ponds are estimated at approximately 0.56 c/kl. Operational costs for the enhanced evaporation spray process are estimated at approximately 18.5 c/kl (3.5% maintenance cost; energy consumption 2 kWh/kl; 2.2 c/kWh). Operational costs for ocean disposal (pumping) are estimated at approximately 26.4 c/kl (12 Ml/d brine).

It should be noted that the brine disposal costs excluded the price of land.

2.2. Electrodialysis reversal

A simplified process flow diagram for EDR treatment of the mine water is shown in Fig. 4. The proposed plant consists of four sections: (a) pretreatment, (b) desalination, (c) post-treatment and (d) brine disposal.

2.2.1. Pretreatment

The pretreatment plant is designed to prepare the water for supply to the desalination EDR plant. This pretreatment stage will remove sufficient Mg, Al, Fe, Mn, organics and SS to render the treated water suitable to use in the EDR plant. This is achieved by aeration, raising the pH by lime addition, and then finally filtration and pH adjustment using sulphuric acid. All the waste streams from the pretreatment stage are collected at a single point where, due to their solids content, they are disposed of on a slimes dam. Liquid runoff from this dump may be returned to the pretreatment plant for recycling.

2.2.2. Desalination

Under the influence of an applied DC field, ions are forced to migrate towards the appropriate electrode. This principal is used to remove ions from the feed water stream and produce a desalinated product stream. The ions that are transported out of the feed stream are collected in a brine stream. Two streams are discharged from the EDR plant, one being substantially salt free (product), the other being salt rich (brine).

For a given feed stream, the degree of purity of the product stream is dependent (up to a point) on the electric power supplied, the number of stages, the plant recovery, and the system chemistry.

The EDR system recovery is controlled by the brine make-up flow rate which determines the amount of brine blow-down discharged. The salt concentration in the brine is a function of the brine make-up and the stack power. This simple control ensures easy optimization of the desalination requirements.

The EDR plant operation efficiency increases with an increase in feed water temperature. As a result, a preheating stage has been included immediately prior to the EDR. This preheating stage will raise the feed water temperature to 35°C. In the interests of thermal efficiency and reduced operating costs, heat is recovered from the streams exiting the EDR. Plate heat exchangers are used to heat the inlet and cool the outlet streams. In order to compensate for any heat losses, a boiler plant will generate steam which will be injected into the feed stream after the exchangers and before the EDR. This steam injection will ensure that a constant feed temperature of 35°C is maintained.

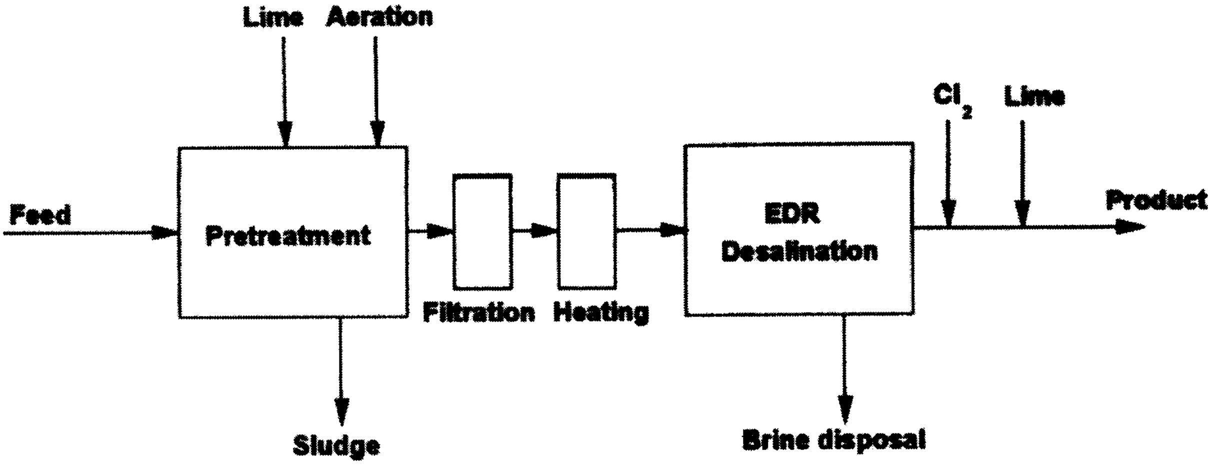

Fig. 4. Simplified process flow diagram of EDR treatment of the mine water.

2.2.3. Post-treatment

In order to render the water safe and non-corrosive, it is necessary to disinfect the water using chlorine and to stabilize it using lime.

2.2.4. Brine disposal

Three distinctly different effluents are produced within the plant battery limits, which should be disposed of separately:

- pretreatment sludge (metal hydroxides)
- brine blow-down
- ash from boilers.

Pretreatment sludge: Pretreatment sludge is formed during the aeration and lime precipitation stages, and is withdrawn from the bottom of the pretreatment clarifier as underflow. This sludge essentially contains metal hydroxides, calcium sulphate, and SS, and is usually finely divided and gelatinous. Due to the nature of this sludge, it should be disposed of in a sealed slimes dam which is lined with HDPE, or other impervious layers.

Brine blowdown: The brine blowdown from the process may be disposed of to evaporation dams or treated further for recycle to the process using a proprietary process.

Ash from boilers: The ash from the boiler plant, being a solid waste, will be transferred with a front-end loader to a hydraulic tip trailer for transfer to a suitable disposal site such as the mine spoils or on the discard dump.

2.2.5. Suggested process configurations for treatment of the mine water

Suggested EDR process configurations and plant performance data for treatment of the mine water are shown in Table 7.

A water recovery of approximately 80% should be obtained with the best-case feed water, while water recoveries of approximately 65% should be obtained with the worst-case feed waters. This means that the brine volume will comprise approximately 35% of the feed in the worst-case feed and 20% in the best-case feed. However, it should be possible to reduce the brine volume to approximately 10% of the feed water by recovering water from the EDR brine. This process, however, needs to be investigated further.

It would be possible to desalinate partially the worst- case feed water with a two-stage EDR process. The TDS would be reduced from 4178 to 1519 mg/l in this case. A four-stage process,

Table 7
EDR process configurations and plant performance data

Best case (low TDS feed):	
Feed DS, mg/l	1072
Product TDS, mg/l	126
Stages	4
Temperature, °C	35
Recovery, %	80
Power (EDR only), kWh/m^3	1.5
Worst case (high TDS feed):	
Feed DS, mg/l	4178
Product TDS, mg/l	1519
Stages	2
Temperature, °C	25
Recovery, %	65
Power (EDR only), kWh/m^3	2.3
Worst case (high TDS feed):	
Feed DS, mg/l	4178
Product TDS, mg/l	345
Stages	4
Temperature, °C	35
Recovery, %	65
Power (EDR only), kWh/m^3	2.5

however, would be required to desalinate the water to potable standards.

2.2.6. Estimated compositions of process streams

The estimated chemical compositions of the EDR feed, product and brine streams for the best and worst cases are shown in Table 8.

It should be possible to produce a very good quality product water for the best-case feed water. It is predicted that the TDS of the feed water can be reduced from 1972 to 126 mg/l in the product. In the case of partial desalination of the worst-case feed water, it should be possible to reduce the TDS of the feed (4178 mg/l) to 1519 mg/l in the product. Sulphate will be reduced from 2660 mg/l in the feed to 944 mg/l in the product. It should also be possible to produce potable water from the worst-case feed water. It is predicted that this water can be desalinated to 346 mg/l in the product.

2.2.7. Capital and operating costs

Capital cost (±20%): The estimated capital cost (desalination plus pretreatment) for the project covering the total scope of work to be undertaken on a turnkey basis is:

- 2-stage plant: US$44.4 M (80 Ml/d feed); US$36.1 M (65 Ml/d feed)
- 4-stage plant: US$53.3 M (80 Ml/d feed); US$43.3 M (65 Ml/d feed); US$30.0 M (45 Ml/d feed)

Capital costs for pretreatment will comprise approximately 50% of the above costs.

Operating costs: Based on a stand-alone plant, it is estimated that the production cost for potable water will be from US$0.37 to US$0.48/m^3 of product, based on design feed water specifications (Table 9). The acidity values given for the water have a significant influence on the cost of treatment, due primarily to the high lime consumption.

It should be noted that at the Tutuka power station, the cost of stack components amounts to US$0.04/m^3 of product compared with the cost of US$0.22/m^3 allowed for this project [5]. The latter figure is based on a 5-year warranty, although field experience demonstrates much longer component lifetimes are possible.

Based on an optimistic life for stack components, it can be shown that an operating cost saving of US$0.09/m^3 may be achieved, resulting in a production cost of US$0.39/m^3 for the four-stage plant. (NOTE: Operational cost for best-case feed is estimated at US$0.44/kl.)

2.2.8. Brine disposal

Brine disposal capital costs for the EDR process are summarized in Table 10.

Table 8
Estimated compositions of process streams

Best case					Worst case				Worst case			
Constituent[a]	Feed	Product	Brine blowdown steady state	Analysis waste average	Feed	Product	Brine blowdown steady state	Analysis waste average	Feed	Product	Brine blowdown steady state	Analysis waste average
Na	150.00	13.68	717.99	695.27	330.00	147.03	779.39	756.92	330.00	38.71	959.03	870.96
Ca	350.00	18.29	1,732.14	1,676.86	650.00	211.33	1,727.43	1,673.56	650.00	44.59	1,957.37	1,774.34
Mg	80.00	4.88	393.02	380.49	200.00	73.85	509.85	494.36	200.00	15.97	597.42	541.78
K	21.00	1.42	102.58	99.32	24.00	7.76	63.90	61.90	24.00	2.08	71.33	64.70
Cl	188.00	15.63	906.21	877.48	287.00	116.46	705.88	684.94	287.00	30.13	841.70	764.04
HCO_3	0.00	0.00	0.00	0.00	25.00	18.07	42.02	41.16	25.00	6.35	65.27	59.63
SO_4	1,182.00	72.20	5,806.17	5,621.21	2,660.00	943.54	6,875.87	6,665.08	2,660.00	207.38	7,956.37	7,214.87
NO_3	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ba	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sr	1.00	0.05	4.59	4.79	1.00	0.33	2.66	2.57	1.00	0.07	3.01	2.73
F	0.40	0.09	1.70	1.65	1.00	0.69	1.75	1.71	1.00	0.26	2.59	2.37
TDS	1,972.40	126.23	9,664.76	9,357.07	4,178.00	1,519.06	10,708.74	10,382.20	4,178.00	245.54	12,454.07	11,295.42
TH	1,204.60	65.81	5,949.58	5,759.78	2,449.00	832.58	6,419.15	6,220.64	2,449.00	177.25	7,354.78	6,667.97
µS/cm	2,499.92	521.89	10,824.16	10,592.38	4,635.54	2,008.48	10,032.11	9,778.62	4,635.54	547.61	11,461.65	10,564.73
pH	7.0	7.0	7.0	7.0	6.0	5.9	6.2	6.2	6.0	5.4	6.4	6.4
LI	−4.8	−4.6	−4.8	−4.8	−1.5	−2.2	−0.6	−0.6	−1.3	−3.5	0.0	−0.1

[a]Concentrations in mg/l unless otherwise stated.
LI, Langelier Index.

Table 9
Summary of EDR operating costs

Operating costs, US$/m^3	4-stage budget	2-stage budget
Power and chemical	0.14	0.14
Manpower	0.09	0.09
Membrane and electrode maintenance	0.22	0.11
Other maintenance	0.02	0.01
Product cost	0.48	0.37

Ocean disposal and brine disposal into lined evaporation ponds again appear to be the most expensive brine disposal options. Costs are especially high where water recovery is only 65% (worst case). However, brine disposal costs appear to be far less expensive in he cases of the enhanced evaporation spray process and unlined ponds. However, it appears that disposal costs into unlined ponds are somewhat less than in the case of the enhanced evaporation spray system, which is preferred, however, because the ponds are lined.

It should be noted that brine disposal costs could be significantly reduced when less brine needs to be treated (8 Ml/d brine). This might be achieved by treating the EDR brine with chemicals and circulating recovered water back to the plant's feed inlet. However, further investigation will be required in this regard.

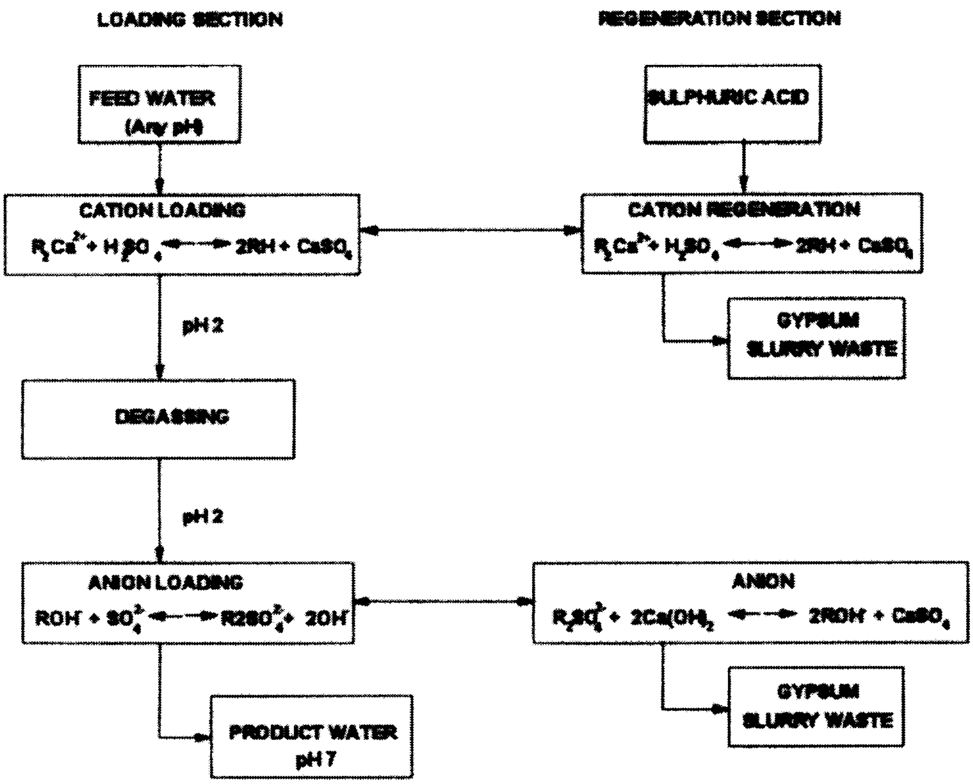

Fig. 5. Simplified process flow diagram of the GYP–CIX process for treatment of mine water.

Table 10
Estimate of EDR brine capital disposal costs

Disposal type	Disposal costs, US$, M				
	3.5 Ml/d brine[c]	9 Ml/d brine[d]	28 Ml/d brine[e]	8 Ml/d brine[f]	22.75 Ml/d brine[g]
Evaporation ponds, unlined[a]	4.0 (4.2)	10.2 (11.1)	31.8 (34.2)	9.1 (9.8)	25.8 (27.8)
Evaporation ponds, lined[a]	16.7	42.7	133.1	38.0	108.2
Enhanced evaporation spray[b]	5.3	13.6	42.0	12.0	34.2
Ocean disposal	—	—	150.7	—	—

Numbers in parenthesis are based on US$1222.2/kl brine to be evaporated per day.
[a]Based on recently determined data [5].
[b]Based on US$1500/per kl brine to be evaporated per day (1 US$ = R4.50).
[c]Based on 65% recovery (10 Ml/d feed).
[d]Based on 80% recovery (45 Ml/d feed).
[e]Based on 65% recovery (80 Ml/d feed).
[f]Based on 90% recovery (80 Ml/d feed).
[g]Based on 65% recovery (65 Ml/d feed).

Operational costs of conventional evaporation ponds are estimated at approximately 0.56c/kl. Operational costs for the enhanced evaporation spray process are estimated at approximately 18.9c/kl (3.5% maintenance cost; energy consumption 2kWh/m³; 2.2c/kWh). Operating costs for ocean disposal (pumping) are estimated at approximately 17.1c/kl (28Ml/d).

2.3. GYP–CIX process

A simplified process flow diagram for treatment of the mine water with the GYP–CIX process is shown in Fig. 5. The proposed plant consists of loading and regeneration sections.

2.3.1. Feed and loading sections

The use of fluidized beds of resin in the loading section enables raw and unfiltered feed water to be used in the process. Furthermore, it is claimed that feed water pretreatment is not a requirement, as the plant can accept feed at any pH. The flow of feed and product water is continuous and uninterrupted.

The continuous counter-current technique used for cation and anion loading is well known, commercially proven, cost-effective, and very suitable for large volumes of effluent requiring treatment. The use of the horizontal loading cascade enables easy plant maintenance, while plant control is facilitated. Maintenance of the loading cascade can easily be carried out, bypassing the particular loading contactor, without interrupting plant operation.

The raw feed water is pumped to the cation loading section where it gravitates through multiple up-flow fluidized bed contact stages. The strong acid cation resin (SAC) is airlifted between stages counter-current to the water flow. The number of stages employed depends on the concentration of salts to be removed and on the level of purity of the product water required.

The decationized water is pumped to a degassing tower where carbonate alkalinity is removed. The decarbonated water is then pumped to the anion loading section, which contains multiple fluidized stages of weak base anion resin (WBA). The mechanical operation of this section is the same as for the cation loading section.

The resulting product water is at a neutral pH, low in calcium and sulphate and other heavy metals or ions, and is non-scaling and is designed to meet effluent discharge specifications.

2.3.2. Regeneration of loaded resins

The novelty of the patented GYP–CIX process is in the resin regeneration technique and the planned production of gypsum as a solid waste product. A single-stage batch regenerator is used while regenerants are recycled to achieve maximum utilization of chemicals.

Cation resin regeneration: The fully-loaded cation resin is airlifted out of the loading section into a batch regenerator where it is put into contact with a 10% sulphuric acid solution, seeded with gypsum crystals, which is recirculated from a stirred tank. The acid strength in the regeneration solution has a conductivity controller linked to a concentrated acid dosing pump.

The solubility of calcium sulphate is low, and as soon as the solubility limit is reached, calcium sulphate precipitates as gypsum. The precipitation of gypsum is enhanced by adding gypsum seed crystals to act as precipitation nuclei to avoid the formation of supersaturated solutions.

When the resin regeneration is completed, the precipitated gypsum slurry is washed out of the resin bed by a clarified overflow from the settler. The gypsum precipitate is concentrated in the settler and discharged from the settler underflow as a thick slurry.

The washed resin is transferred to a resin rinse vessel where the regeneration solution is rinsed from the resin pores using freshly decationized water from the feed. The regenerated, rinsed and conditioned resin is then returned to the last contactor at the downstream end of the cation loading section.

Anion resin regeneration: The regeneration of the loaded anion resin is achieved with lime. To overcome the low solubility of lime, a 2% lime slurry is used, which is again seeded with gypsum crystals. This slurry is recirculated from a regenerant tank in which the strength of the lime slurry is controlled by a pH controller that is used to dose 5% lime slurry from a bulk regenerant tank.

The anion regeneration also produces gypsum, which is removed from solution by settling, and finally is discharged as slurry waste as in the cation section. The continuous precipitation of gypsum in solution in both anion and cation sections allows the regeneration solution to be reused for subsequent regenerations without a build-up of the stripped ion in solution, which further minimizes reagent consumption.

The anion resin is washed using clarified overflow from the settler to remove precipitated solids in the anion section, and is then rinsed using final product water in a similar fashion to the cation resin. The freshly regenerated resin is then returned to the anion loading section at the product end.

Waste product: In cases where high water recoveries are required, the gypsum sludge from the cation and the anion settlers can be combined to generate additional gypsum. The resulting sludge can then be filtered and disposed of as a solid and inert gypsum cake. The small volume of filtrate can be returned to the feed water tank for recycling.

2.3.3. Estimated compositions of process streams

The estimated chemical compositions of the product water quality are shown in Table 11.

Water of potable quality can be produced with the GYP–CIX process. A TDS of less than 240 and 700 mg/l can be produced, depending on how the process is operated.

2.3.4. Capital and operating costs

The capital and operating costs of the GYP–CIX process to treat the mine water is shown in Table 12.

Table 11
Feed and product water quality

Parameter	Feed 1: probable	Feed 2: probable	Feed 3: best	Feed 4: worst	GYP–CIX: product 1	GYP–CIX: product 2
TDS, mg/l	2714	3 451	1 999	4 472	<240	<700
TSS, mg/l	258	200	90	350	Unchanged	Unchanged
pH	6.0	6.1	4.5	5.8	8 - 9	8 - 9
Calcium, mg/l	340	432	200	500	<40	<50
Magnesium, mg/l	152	182	80	200	<20	<50
Sodium, mg/l	271	330	150	330	<25	<100
Potassium, mg/l	21	0.016			<5	<10
Iron, mg/l	158	247	150	285	<0.1	<0.1
Manganese, mg/l	6.7	6.9	15	10	<0.1	<0.1
Nickel, mg/l	0.7	0.565	0.71	0.565	<0.1	<0.1
Zinc, mg/l	0.015	0.023	0.015	0.023	<0.01	<0.01
Strontium, mg/l	1.02	0.9	1.0	0.899	<0.01	<0.01
Cadmium, mg/l	0.04	0.005	0.014	0.0005	<0.01	<0.01
Barium, mg/l	0.001	0.001	0.001	0.001	<0.001	<0.001
Aluminum, mg/l	0.113	0.01	0.113	0.03	<0.01	<0.01
Vanadium, mg/l	0.013	0.0015	0.013	0.0015	<0.001	<0.001
Chromium, mg/l	0.002	0.0015	0.002	0.0015	<0.001	<0.001
Boron, mg/l	0.46	0.385	0.458	0.385	<0.1	<0.1
Sulphate, mg/l	2037	2523	1182	2827	<50	<250
Chloride, mg/l	188	287	188	287	<70	<150
Nitrate, mg/l	0.03	0.02	0.03	0.02	<0.03	<0.03
Phosphate, mg/l	0.009	0.016	0.009	0.016	<0.02	<0.02
Fluoride, mg/l	0.4	0.6	0.4	0.6	Unchanged	Unchanged
Alkalinity, mg/l	51	20			<50	<100

It is interesting to note that the operating cost is significantly less when the feed water is only desalinated to a TDS of approximately 700 mg/l. It is also claimed that water recoveries in excess of 90% should be possible with this process, if water could be recovered from the brine.

2.3.5. Brine disposal

Brine disposal capital costs for the GYP–CIX process are summarized in Table 13.

Ocean disposal and brine disposal into lined evaporation ponds appear to be the most expensive brine disposal options. However, brine treatment with the enhanced evaporation spray process and disposal into unlined ponds appear to be far less expensive.

Brine disposal will be very expensive when only 54% water recovery can be obtained (36.8 Ml/d brine, worst case feed water). However, brine disposal costs become less expensive with increasing water recovery (less brine production). It is interesting to note that brine disposal costs will be relatively low when the water recovery is increased to 90% (8.0 Ml/d brine, worst case). This, however, needs to be proved on an experimental basis.

Operational costs for conventional evaporation ponds are estimated at approximately 0.56 c/kl. Operational costs for the enhanced evaporation

Table 12
Summary of capital and operational costs

Parameter	Feed 1: probable	Feed 2: probable	Feed 3: best	Feed 4: worst
Feed TDS, mg/l	2714	3451	1999	4472
GYP–CIX:				
Product 1: TDS 240 mg/l				
• Wat. rec., %	79	58.1	88.2	54
• Opex, US$/m^3	0.52	0.55	0.31	0.60
• Capex, US$:				
for 45 Ml/d	16.7 M	16.7 M	16.7 M	16.7 M
for 80 Ml/d	26.7 M	26.7 M	26.7 M	26.7
for 65 Ml/d	22.2 M	22.2 M	22.2	22.2 M
Product 2: TDS 700 mg/l				
• Wat. rec., %	82.4	68.3	91.3	58.2
• Opex, US$/m^3	0.44	0.54	0.24	0.55
• Capex, US$:				
for 45 Ml/d	16.7 M	16.7 M	16.7 M	16.7 M
for 80 Ml/d	26.7 M	26.7 M	26.7 M	26.7 M

Table 13
Estimate of GYP–CIX brine capital disposal costs

	Disposal costs, US$, M					
Disposal type	5.4 Ml/d brine[c]	16.8 Ml/d brine[d]	33.6 Ml/d brine[e]	36.8 Ml/d brine[f]	8 Ml/d brine[g]	29.9 Ml/d brine[h]
Evaporation ponds, unlined[a]	6.2 (6.7)	19.1 (20.4)	38 (34.2)	41.8 (44.9)	9.1 (9.8)	34.0 (36.7)
Evaporation ponds, lined[a]	25.6	79.8	159.6	175.1	38.0	142
Enhanced evaporation spray[b]	8.2	25.1	50.4	55.1	12.0	44.9
Ocean disposal	—	—	—	190.7	—	—

Numbers in parenthesis are based on US$1222.2/kl brine to be evaporated per day.
[a]Based on recently determined data [5].
[b]Based on US$1500/per kl brine to be evaporated per day.
[c]Based on 88% recovery (45 Ml/d feed) (best case).
[d]Based on 79% recovery (80 Ml/d feed) (probable case).
[e]Based on 58% recovery (80 Ml/d feed) (probable case).
[f]Based on 54% recovery (80 Ml/d feed) (worst case).
[g]Based on 90% recovery (80 Ml/d feed) (worst case).
[h]Based on 54% recovery (65 Ml/d feed) (worst cast).

spray process are estimated at approximately 18.9 c/kl (3.5% maintenance cost; energy consumption 2 kWh/m^3; 2.2 c/kWh). Operational cost for ocean disposal (pumping) is estimated at 15.1 c/kl (36.8 Ml/d brine).

Table 14
Economics of the desalination of the worst-case mine water with the RO, EDR and GYP–CIX processes (80 Ml/d feed) (in US$, M)

		RO	EDR	GYP–CIX
Annual operating costs:				
RO	88.0 c/kl	21.9		
EDR	47.6 c/kl		9.0	
GYP–CIX	60.4 c/kl			9.6
Brine disposal	18.9 c/kl	0.82	1.9	2.5
Annual savings:				
Water @ US$0.22		5.5	4.2	3.5
Operating costs minus savings		17.2	6.7	8.6

2.4. *Economics of the desalination of the worst-case mine water with RO, EDR and GYP–CIX processes*

The economics of the different treatment processes are summarized in Table 14.

3. Conclusions

3.1. *Reverse osmosis*

- It is predicted that water of potable quality should be produced from Grootvlei mine water (best-, probable- and worst-case waters) with SRO. It was demonstrated that it should be possible to reduce the TDS of the mine water in the TDS range of 2000–2700–3400–4500 mg/l to potable standards with SRO at a water recovery of 85%. Therefore, brine volume will comprise 15% of the treated water.
- The capital costs (pretreatment and desalination) of RO desalination plants were estimated at US$31.8 M (US$13.9 M pretreatment and US$17.9 M desalination), and US$35 M (US$15.2 M pretreatment and US$19.8 M desalination) for the probable- and worst-case feed waters, respectively (80 Ml/d plant). Total operating costs were estimated at 70.8 and 88.1 c/kl for the probable- and worst-case feed waters, respectively. Brine disposal costs varied significantly from unlined to lined ponds. Brine disposal costs for 12 Ml/d brine (80 Ml/d feed; 85% water recovery) were estimated at US$13.6 M (unlined ponds), US$57.1 M (lined ponds), and US$18 M for enhanced evaporation ponds. Total desalination costs for the 80 Ml/d plant (85% recovery, worst case) is estimated at US$53.0 M (US$39.0 M + 18.0 M).

3.2. *Electrodialysis reversal*

- It is predicted that it should be possible to produce potable water from the worst- and best-case feed water with the EDR process. It was demonstrated that TDS in the feed could be reduced from 4178 (worst case) to 246 mg/l in the EDR product at 65% water recovery. The feed can also be partially desalinated to 1519 mg/l TDS (65% recovery) if required.
- The capital cost (pretreatment plus desalination) to desalinate the worst-case feed water (65% recovery; 80 Ml/d) to potable quality is estimated at US$53.3 M (50% of the cost is due to pretreatment). The operational costs for the worst (80 Ml/d plant) case feed water are estimated at 47.6 c/kl. Operational costs for partial desalination of the feed water are estimated at 36.7 c/kl. Brine disposal costs for the worst-case feed water (28 Ml/d brine, 80 Ml/d feed, 65% water recovery) in unlined ponds, lined ponds and with the enhanced evaporation spray system are estimated at 31.8; 133.1 and US$42.0 M, respectively. The total desalination cost for the worst-case feed water (80 Ml/d plant) is estimated at US$95.3 M (US$53.3 M + US$42.0 M). Total desalination costs for partial desalination of 80 Ml/d feed (65% recovery, worst case) are

estimated at US$86.4 M (US$44.4 M+ US$42.0 M).

3.3. GYP–CIX process

- It is predicted that it should be possible to produce potable water from the worst-, probable- and best-case feed water with the GYP–CIX process. It was demonstrated that TDS in the feed water concentration range from 2000 to 4500 mg/l could be reduced to less than 240 mg/l. Partial desalination to approximately 700 mg/l TDS in the product water is also possible. Water recovery for the best- (1999 mg/l), probable- (2714 and 3451 mg/l) and worst-case (4472 mg/l) feed concentrations are estimated at 88.2, 79, 58.1, and 54%, respectively. Water recovery, however, might be increased to approximately 90% in the case of the worst-case feed water if water could be recovered from the brine.
- Capital costs for an 80 Ml/d plant are estimated at US$26.7 M (no pretreatment). Operational costs for feed waters of 2714, 3451, and 4472 mg/l (probable to worst cases) are estimated at 51.8, 55.3, and 60.4 c/kl, respectively. Brine disposal costs for the worst-case feed water (80 Ml/d plant, 54% recovery) are estimated at 41.8, 175.1, and US$55.1 M for unlined, lined and enhanced evaporation spray ponds, respectively. However, disposal costs decrease with increasing water recovery for the probable-case feed waters (80 Ml/d feed). Brine disposal costs were estimated at 38, 159.6, and US$50.4 M for lined, unlined and enhanced evaporation spray ponds when the water recovery was increased to 58%. These costs were 19.1, 79.8, and US$25.1 M when the water recovery was increased to 79%.
- However, it should be possible to decrease these costs significantly if water recovery could be increased to 90%. Disposal costs are then predicted to be 9.1, 38.0, and US$12.0 M, respectively. The total desalination costs for the 80 Ml/d plant are estimated at US$51.8 M (US$26.7+US$25.1 M) to US$81.8 M (US$26.7 M + US$55.1 M) in the feed water TDS range from 2700 to 4500 mg/l.

3.4. Overall economics

The operating costs minus the savings in water sales were estimated at 17.2, 6.7, and US$8.6 M/y for the RO, EDR and GYP–CIX processes, respectively (worst case). Therefore, the operational costs of the EDR and GYP–CIX processes are the lowest if the sale of water is taken into consideration. This may favour the EDR and GYP–CIX processes for desalination of the mine water.

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