Desalination of calcium sulphate scaling mine water: Design and operation of the SPARRO process

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Abstract

The South African mining industry discharges relatively small quantities of mine service water to the environment, but these effluents contribute substantially to the salt load of the receiving surface waters. The poor quality of mine service water also has significant cost implications on the mining operations. Of the two main types of mine service water encountered in the gold mining industry, the so-called calcium sulphate scaling type is found in the majority of cases. Preliminary testwork on this type of water using membrane desalination processes revealed that only the seeded reverse osmosis type of process showed promise. To overcome certain process problems and high operating costs with this system, a novel membrane desalination technique incorporating seeded technology, called the SPARRO (slurry precipitation and recycle reverse osmosis) process, was developed. The novel features of the new process included; a lower linear slurry velocity in the membrane tubes, a lower seed slurry concentration, a dual pumping arrangement to a tapered membrane stack, a smaller reactor and a modified seed crystal and brine blow-down system. Evaluation of the SPARRO process and its novel features, over a five-year period, confirmed its technical viability for desalinating calcium sulphate scaling mine water. The electrical power consumption of the process was approximately half that of previous designs, significantly improving its efficiency. Membrane performance was evaluated and was generally unsatisfactory with both fouling and hydrolysis dominating at times, although operating conditions for the membranes were not always ideal. The precise cause(s) for the membrane degradation was not established, but a mechanism for fouling (based upon the presence of turbidity in the mine water) and a hypothesis for a possible cause of hydrolysis (alluding to the presence of radionuclides in the mine water) were proposed. Product water from the SPARRO process has an estimated gross unit cost (including capital costs) of 383 c/m^3 (1994).

Introduction

The mining industry in South Africa uses vast quantities of water during its daily mining operations. This mine service water is recycled for reuse and can be classified as potentially calcium sulphate scaling in the majority of gold mines. The poor quality of mine service water has significant cost implications on mining and the excess water adds a considerable salt load to the environment when discharged.

Mackay et al. (1991) estimate that poor mine service water costs the gold mining industry in the region of R300 m. annually, based on its corrosive, scaling and fouling potential. The quality of mine service water varies from mine to mine but in general the total dissolved solids (TDS) concentration is above 1 800 mg/ℓ, as confirmed in a mine water survey of ten mines during 1988/1989 (Pulles, 1991). The major contributor to the high TDS of mine service water is sulphate resulting from the bacterial and chemical oxidation of pyrites which are associated with the gold-bearing reefs.

An example, which highlights the impact that large salt loads from industries can have on the limited water resources, is the contribution by three active gold mines (Grootvlei, ERPM and Durban Roodepoort Deep) over a period of 16 months during 1988/89 to the Klip River, which is the main contributor to salt load entering the Vaal Barrage (Rand Water Board, 1988; 1989). The volumetric contribution to the river from these mines amounted to 9% of the flow, while the salinity contributed 48% to the TDS load and 60% to the sulphate load (Funke, 1990).

In general, the pollution potential from the mining industry is significant. The Department of Water Affairs and Forestry through the Department of Mineral and Energy Affairs is bringing pressure to bear on the mining industry to manage the effects of the mining activities on the environment; to mitigate the negative impact while maximising the positive features. In order to enforce this responsibility the Minerals Act, 1991, requires the owner of every mine to submit and obtain approval for an Environment Management Programme Report (EMPR) before mining operations commence (Department of Mineral and Energy Affairs, 1992).

Depending upon the age of the mine and its geographical location, it is possible to have a mine service water which has high concentrations of calcium, sulphate, sodium and chloride. Table 1 shows analyses of water from selected mines in different mining areas and illustrates the range of water quality that is possible.

These analyses reveal that from a treatment point of view mine service water in the gold mining industry may be grouped into different types, namely: those that are dominated by the presence of calcium and sulphate (Mines B and C) and those that are dominated by the presence of sodium and chloride. Although Mine B has high concentrations of sodium and chloride as well as calcium and sulphate, the latter ions are more important, because of the relatively high degree of $CaSO_4$ saturation of the water (41%). Waters such as those from Mine A are termed non-scaling mine waters (relative to $CaSO_4$) while the other two are termed $CaSO_4$ scaling mine waters (Mines B and C). Most of these waters are undersaturated with respect to $CaSO_4$ and should therefore strictly be termed potentially $CaSO_4$ scaling mine waters, but the term $CaSO_4$ scaling mine waters is that which is used in this paper. The high degree of $CaSO_4$ saturation has severe implications in desalination technology due to the problems caused by scale

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TABLE 1
QUALITY OF MINE SERVICE WATER FROM DIFFERENT MINES

Determinant	Units	Mine A Free State (New)	Mine B Free State (Old)	Mine C West Rand (Old)
TDS	mg/ℓ	3 600	3 800	2 200
pH		6.9	6.5	6.0
Sodium	mg Na/ℓ	1 250	1 000	170
Calcium	mg Ca/ℓ	130	330	400
Sulphate	mg SO_4/ℓ	150	1 050	1 400
Chloride	mg Cl/ℓ	2 000	1 400	60
Magnesium	mg Mg/ℓ	15	30	50
* $CaSO_4$ degree of saturation	%	5	41	52
Concentration factor to reach 100% saturation		20	2.44	1.92

*100% = saturation

formation on membrane and other surfaces.

In view of the problems mentioned, the mining industry has devoted considerable research and development efforts to treating and desalinating mining effluents. This paper describes the work which was carried out to develop a novel seeded membrane desalination process to treat calcium sulphate scaling mine water; with the emphasis on design parameters and operating experience. Costs are also presented.

Earlier research and development on desalination of mine water

The conventional membrane desalination processes of electrodialysis reversal (EDR) and tubular reverse osmosis (TRO) have been shown to be technically viable for desalting non-scaling mine services waters, with high concentrations of dissolved sodium and chloride ions (Juby, 1992; Juby and Pulles, 1990; 1991).

However, an investigation into the quality of mine service waters revealed that most of them would become scaling when concentrated in a desalination process operating at a water recovery of 80% (Juby, 1992); that is 80% of the feed water to the plant recovered as product water. An 80% product water recovery is considered to be the economical lower limit for desalinating mine water, and implies a concentration factor of 5 between the feed stream and the brine.

Consequently, to avoid the possibility of $CaSO_4$ scaling occurring within the process equipment, the $CaSO_4$ saturation level of the feed stream should be less than 20%. Water samples taken from 26 of the large gold mines incorporating 54 shaft systems indicate that 77% have initial $CaSO_4$ saturation levels greater than 20% (Pulles et al., 1992). Table 1 presents examples of two of these waters - Mines B and C. As a result of the dominance of scaling type mine waters, processes suitable for treating them would be applied far more widely in the gold mining industry than the conventional membrane systems such as EDR and TRO, which would be suitable for treating non-scaling mine water.

The seeded reverse osmosis concept is a possible approach to treating calcium sulphate scaling waters. The concept was developed in the late 1970s by Resources Conservation Company (RCC) in Seattle, USA (O'Neail et al., 1981). Essentially a slurry of seed crystals is incorporated into the feed water of a tubular reverse osmosis system.

The seed serves as preferential growth sites for calcium sulphate, other calcium salts and silicates which begin to precipitate as their solubility product is exceeded during the concentration process within the membrane tubes (Di Benedetto, 1984). The preferential growth of scale on the seed crystals prevents scale formation on the membrane surface. Figure 1 illustrates the concept of seeded reverse osmosis.

Initial investigations with SRO

A small (0.05 ℓ/s) SRO pilot plant, designed and built by RCC, was installed at a test site at the East Rand Proprietary Mines Ltd. (ERPM), Hercules shaft, where it operated for some 5 000 h.

Harries (1984; 1985) reports that the pilot plant achieved product water recoveries of between 92 and 96%. The only

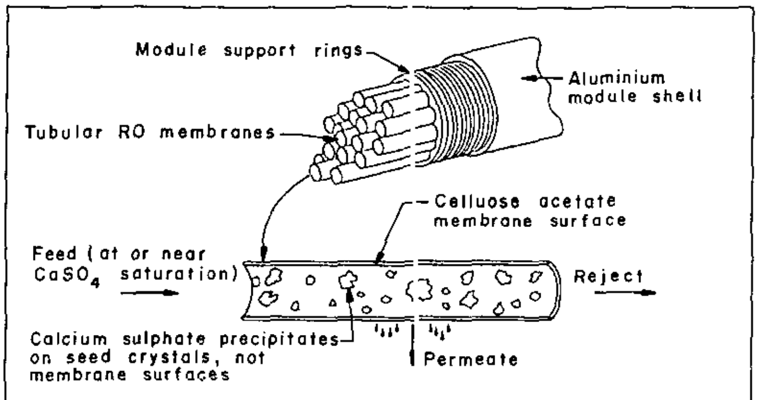


Figure 1
Conceptual illustration of seeded reverse osmosis

pretreatment of the mine water undertaken ahead of the pilot plant was coarse (50 micron) particle filtration to protect the high pressure pump, pH control to between pH 4 and 6 to reduce hydrolysis of the cellulose acetate membranes and chlorine addition to prevent biological growth within the system (Harries, 1985).

The quality of the permeate produced by the five-module plant was good for the first 3 000 h of operation during which the salt rejection was consistently above 96% (Harries, 1985). However, for the last 2 000 h the membrane salt rejection began to decrease gradually and had dropped to about 93% by the end of the study. Permeate flux stabilised during the operating period at around 650 $\ell/m^2 \cdot d$ (corrected to 25°C and 2 750 kPa).

Harries (1985) concluded that the cellulose acetate modules had performed satisfactorily in the seeded slurry mode with no evidence of fouling or scaling of the membranes. Furthermore it was concluded (Harries, 1984) that membrane hydrolysis had been the cause of the decrease in salt rejection observed and that this had been due to variations in feed water temperature and pH outside the limits of 30°C and the range 4 to 6 respectively, as specified by the membrane manufacturer.

The issues raised regarding membrane performance (the indications of hydrolysis) during the 0.05 ℓ/s study could not be investigated fully in a further study on a 0.5 ℓ/s pilot plant because only 600 h of operation were achieved. Therefore after the two pilot plant studies it was certainly not possible to make a conclusive statement in regard to expected membrane performance in any future plants.

Therefore, in 1986 COMRO embarked on a new project to develop a modified seeded reverse osmosis process. The focus of the project was to provide a treatment solution to the problems associated with both the cost of recirculating and using poor quality scaling mine service water in mines, and the discharge of such effluents to the environment. The initial work was carried out in collaboration with the Water Research Commission, Iscor and Bintech (now Membratek).

The SPARRO process

The development and evaluation of the new seeded membrane process took place over a period of more than five years. The main objective was to overcome the severe shortcomings of the existing seeded systems evaluated up to that point, which were:

- a high energy consumption of approximately 9 kWh/m^3 of product
- a high $CaSO_4$ slurry recirculation rate of about 1.4 times the product flow rate
- a poor $CaSO_4$ seed and brine mass balance control system.

The process as developed was termed the slurry precipitation and recycle reverse osmosis process, or SPARRO.

SPARRO process description

The process flow diagram (PFD) of the SPARRO pilot plant is shown in Fig. 2.

The pretreated and conditioned feed water flows into a small raw feed water tank with a hydraulic residence time of about one hour (volume = 3.5 m^3). The level in the tank is maintained by means of a float control valve. Raw feed water is drawn from the base of the tank to one of two raw water feed pumps (one duty, one standby). These 12-stage centrifugal pumps pressurise the clear raw water to a gauge pressure of 4 000 kPa.

A magnetic flow meter and pneumatic control valve on the bypass line from the raw water feed pumps ensures that the required flow rate of the raw feed water (Stream 2) is correct and also ensures that the pumps will not pump against closed valves.

The pressurised raw feed water passes through a non-return valve after which it mixes with the gypsum slurry recycle stream (Stream 13), in the ratio of 1.0 : 0.8. Stream 13 is also maintained at a gauge pressure of 4 000 kPa, by one of two slurry recycle pumps (one duty, one standby). These pumps are positive displacement triplex plunger pumps and they draw slurry recycle from the base of the reactor vessel.

Stream 13 contains about 41 g/ℓ of gypsum ($CaSO_4.2H_2O$) crystals in suspension. After mixing with Stream 2 the combined stream (Stream 3) contains about 18 g/ℓ of solids. Stream 3 is the feed stream to the module bank, and the well-mixed slurry enters the modules at a gauge pressure of about 4 000 kPa and a total dissolved solids (TDS) concentration of around 7 500 mg/ℓ.

The reject stream leaving the module bank (Stream 5) is at a gauge pressure of about 2 500 kPa, and has a TDS of about 15 000 mg/ℓ and a suspended solids concentration of about 36.5 g/ℓ, as a result of the concentrating effects within the membrane module bank. This stream passes through a pressure-reducing orifice plate after which the pressure is reduced to a gauge pressure of 510 kPa. The stream then splits and a controlled volume (Stream 6) flows into the hydrocyclone system, the remainder flowing directly into the reactor (Stream 12).

The overflow stream from the hydrocyclones (Stream 8) containing only a small quantity of solids is returned to the reactor, with a controlled and metered portion of the stream being periodically diverted to waste. The stream diverted to waste is known as the brine disposal stream (Stream 9). In a similar manner a small controlled and metered portion of the hydrocyclone underflow is also periodically diverted to waste. This is known as seed disposal stream (Stream 11). The bulk of the hydrocyclone underflow returns to the reactor vessel to maintain the required suspended solids concentration in the system.

In the reactor vessel a mechanical stirrer ensures complete mixing during the time required for desupersaturation of the supersaturated solutions which return to the reactor from the membrane bank. The hydraulic residence time in the reactor is about one hour.

Product water or permeate produced by the membranes has a TDS of about 900 mg/ℓ (Stream 4). This water is collected from all the membrane modules and gravitates to a storage tank from where it can be used to flush the membrane module bank before a plant shutdown.

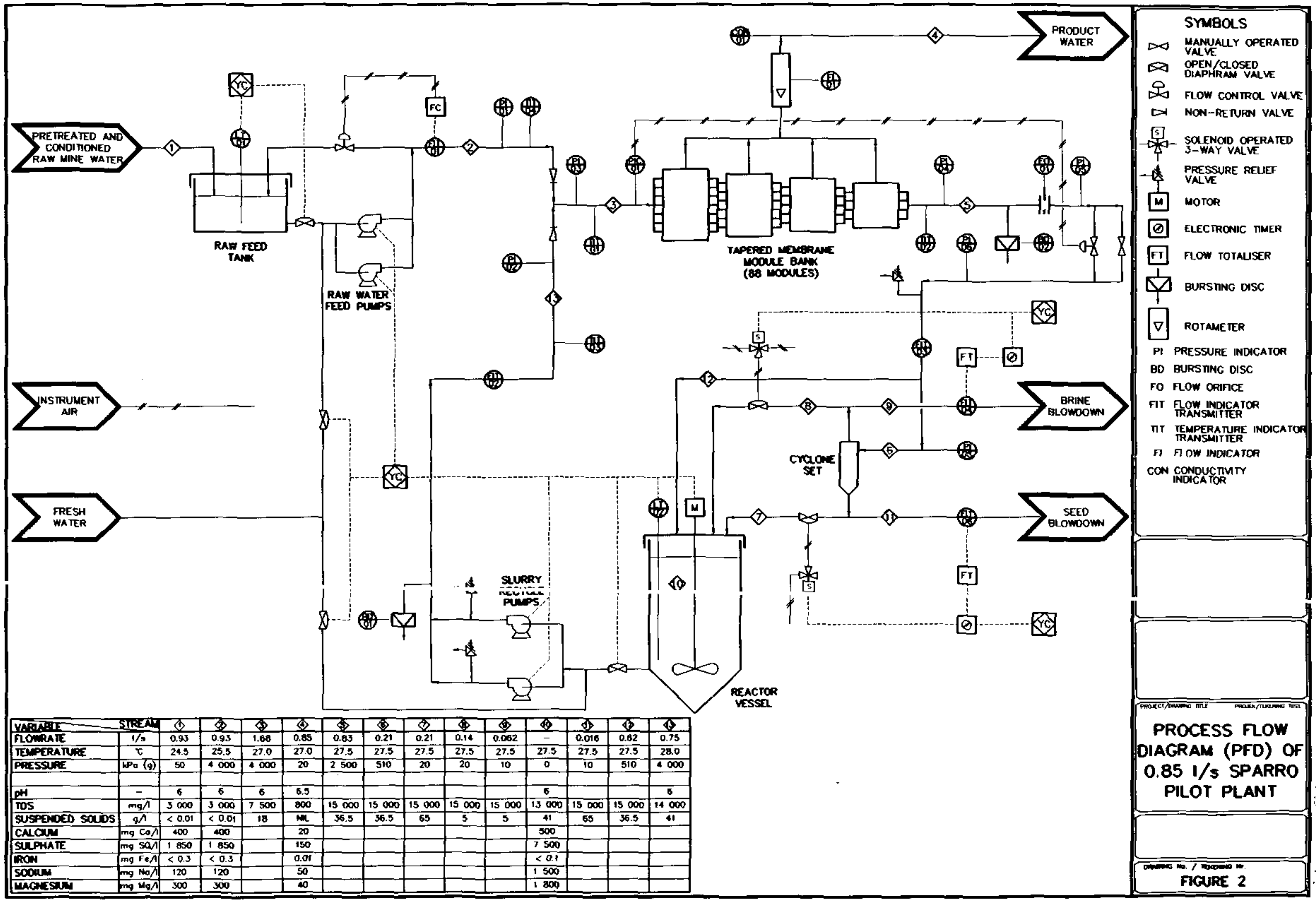
Design parameters

Membrane stack and module configuration

Tube velocity

One of the problems with the 0.5 ℓ/s pilot plant was its high power consumption. This was due to two interrelated factors; the high linear tube velocities of around 3 m/s and the requirement for interstage pumping.

Minimum velocity calculations were carried out for an assumed gypsum particle size of 200 μm at a concentration of 50 g/ℓ. Both of these figures were conservative. The calculations were repeated for several forms of the minimum or critical velocity equations checked by physically observing the slurry at various velocities.



VARIABLE	STREAM	1	2	3	4	5	6	7	8	9	10	11	12	13
FLOWRATE	l/s	0.93	0.93	1.68	0.85	0.83	0.21	0.21	0.14	0.062	–	0.016	0.62	0.75
TEMPERATURE	°C	24.5	25.5	27.0	27.0	27.5	27.5	27.5	27.5	27.5	27.5	27.5	27.5	28.0
PRESSURE	kPa (g)	50	4 000	4 000	20	2 500	510	20	20	10	0	10	510	4 000
pH	–	6	6	6	6.5						6			6
TDS	mg/l	3 000	3 000	7 500	800	15 000	15 000	15 000	15 000	15 000	13 000	15 000	15 000	14 000
SUSPENDED SOLIDS	g/l	< 0.01	< 0.01	18	NIL	36.5	36.5	65	5	5	41	65	36.5	41
CALCIUM	mg Ca/l	400	400		20						500			
SULPHATE	mg SO_4/l	1 850	1 850		150						7 500			
IRON	mg Fe/l	< 0.3	< 0.3		0.01						< 0.1			
SODIUM	mg Na/l	120	120		50						1 500			
MAGNESIUM	mg Mg/l	300	300		40						1 800			

Figure 2
Process flow diagram of SPARRO pilot plant

Tapered configuration

Both pilot plants that had been tested previously had used non-tapered module configurations in the stack. The implication with such systems is that the tube velocity diminishes as the slurry flows through the membrane stack, as a result of the permeation of the product water through the membrane walls.

The physical reduction in slurry volume with distance travelled through the membrane is one negative factor in slurry systems. The other negative factor is that a reduction in volume means that the concentration of the slurry in suspension would be increasing. A third, more complex factor, could be the ongoing precipitation of calcium sulphate from the solution onto existing crystals of calcium sulphate, thus increasing the concentration of the solid phase even further.

All three of the points mentioned above can contribute to solids sedimentation occurring in the membrane module tubes. In the SPARRO design it was decided to arrange the modules in a tapered configuration so that at the end of each section of the taper the tube velocity would not be less than the required minimum velocity of 1.25 m/s.

After various simulations to determine tube velocities and anticipated head loss, the following membrane module stack configuration was decided upon: four banks of modules, consisting of 30 modules in the first bank (arranged as three rows of ten modules in parallel), 24 modules in the second bank (arranged as three rows of eight modules in parallel), 24 modules in the third bank (arranged as four rows of six modules in parallel), and 10 modules in the last bank (arranged as two rows of five modules in parallel). This arrangement is shown schematically in Fig. 3.

The module stack therefore consisted of 88 modules which gave the plant a nominal capacity rating of about 0.85 ℓ/s.

Pressure drop and concentration factor

With the reduced slurry velocity in the membranes the pressure drop across the SPARRO membrane stack would be far lower than that experienced in the other pilot-plant studies. It was calculated that the pressure drop across each membrane module would reduce from about 300 kPa in the old pilot plants to only about 70 kPa in the SPARRO pilot plant.

Thus for the same overall pressure drop in a process operating at a feed gauge pressure of 4 000 kPa, the SPARRO plant arrangement would be able to operate with more modules in series. Therefore the overall production of product water relative to the feed flow rate, termed extraction, would increase, which would lead to a more economical design. Tables 2 and 3 summarise design values for the different process variables. Full descriptions of the design values are given by Juby (1994).

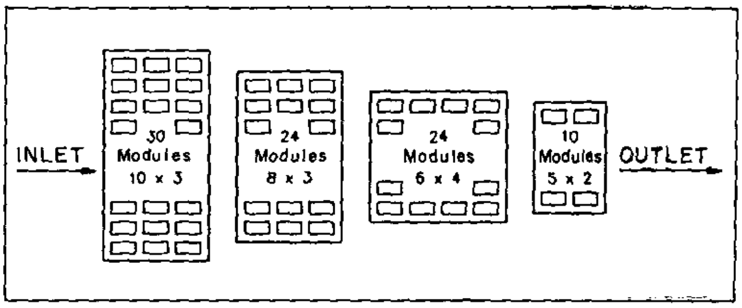


Figure 3
Schematic representation of membrane module stack in SPARRO pilot plant

Feed pumping system

One of the problems with seeded RO systems that had become quite obvious during previous investigations, was that the pumping systems used for recycling the slurry were inadequate.

There were two problems with the slurry recycle systems. The first was that the volume of slurry recycled was too high. This resulted in excessive power usage and also meant that large pumps would need to be incorporated at full scale to pump the slurry. The second problem was one of excessive pump wear.

In the SPARRO process, by adopting a novel pumping procedure, an attempt was made to overcome some of the previous pumping problems.

Instead of feeding the raw mine water into the reactor vessel, as in previous seeded RO pilot plants, a separate pumping system, incorporating a standard multistage centrifugal pump used for clean water applications, was used to pressurise the pretreated raw mine water.

By this simple system, the volume of slurry that required pumping in a given time was roughly halved. This would have a positive effect on the slurry pump life, and implies that smaller, readily available slurry pumps could be used on full-scale applications without the need for installing multiple pump sets to achieve the required flow rate. Furthermore, the slurry pump chosen for the slurry recycle application was the positive displacement triplex plunger pump, which is used widely within the mining industry for pumping slurries, and is a pump with which the mining industry is familiar.

The so-called dual pumping system of the SPARRO process was seen as a novel approach to reducing pump maintenance costs in seeded RO systems.

TABLE 2
DESIGN VALUES FOR DIFFERENT PROCESS VARIABLES

Process variable	Condition		
	Minimum	Design	Maximum
Corrected membrane flux (ℓ/m^2·d)	400	550	~600
Inlet pressure (kPa (g))	3 600	4 000	4 500
Product water recovery (%)	80	90	~99
Product extraction (conversion) (%)	40	50	unknown

Development of a process model for the SPARRO process

With all of the new features incorporated in the process simultaneously, it was not possible to predict the process performance based on any previous seeded RO investigations. Consequently, in order to check some of the new process design parameters and their influence on the expected performance of the SPARRO plant, a computer model was developed to simulate, as simply as possible, the operation of the new process. Details are given by Juby (1994).

Pilot plant operation

After the commissioning of the plant was complete and it was able to operate continuously over a 24 hour period, it was operated periodically until March 1993. During this time the plant was operated in various modes and with varying degrees of pretreatment to the raw mine water. It had been intended to operate the plant continuously for as long as possible. However, operational circumstances and difficulties at times meant that operational changes had to be made, and in some cases, delays were encountered.

The approximately five-year operating period of the SPARRO plant can be divided into four phases. The operating hours achieved for which plant performance data was obtained, were:

- Phase I - March 1989 to December 1989 (2 615 operating hours)
- Phase II - January 1990 to August 1990 (3 316 operating hours)
- Phase III - December 1991 to February 1992 (1 361 operating hours)
- Phase IV - October 1992 to February 1993 (1 100 operating hours)

Briefly, during Phase I the mechanical integrity of the process was evaluated. The plant was installed with a set of old membranes from previous pilot-plant work and therefore product water quality was not of major concern. At the start of Phase II the entire membrane inventory was replaced with new membranes and this phase was viewed as the first real testing phase for the membranes in terms of product water quality performance. Although Phase III was carried out after the plant had been shut down for over a year, mostly new membranes were installed in the plant (all but nine) at the start of Phase III and this was an additional phase in which the membrane and plant performance were evaluated. In Phase IV the plant was operated with basically the same membranes that had been used during Phase III, except for ten membrane modules which had been replaced. The configuration of the plant was changed slightly for this phase of operation in order to test a postulate proposed by Pulles et al. (1992) relating to the membrane fouling observed during Phase II (discussed later).

TABLE 3
DESIGN INFORMATION FOR SPARRO PILOT PLANT REACTOR VESSEL

Description	Value
Reactor	
Maximum capacity (m^3)	4.5
Design capacity (m^3)	3.5
Dimensions	
Dia. (m)	1.5
Height (m)	2.5
Mixer	
Power (kW)	0.75
Speed (r/min)	290
Hydraulic residence time (h) (at design)	1

TABLE 4
AVERAGE AND RANGE OF CONSTITUENTS IN COMBINED RAW MINE WATER AND SLURRY RECYCLE FEED TO SPARRO PLANT DURING PHASE II

Determinant	Units	N	Average	Std deviation	Range min	Range max
pH	-	18	4.76	0.55	4.06	5.94
Conductivity	mS/m	20	734	180	469	1 102
TDS	mg/ℓ	20	9 410	2 938	5 370	16 272
Suspended solids	g/ℓ	20	19.9	7.5	10.2	36.7
Calcium	mg Ca/ℓ	18	468	52	417	611
Magnesium	mg Mg/ℓ	18	635	243	272	1 011
Sodium	mg Na/ℓ	18	745	298	355	1 477
Chloride	mg Cl/ℓ	19	196	52	101	266
Sulphate	mg SO_4/ℓ	19	5 560	1 492	3 381	7 853
Iron	mg Fe/ℓ	18	0.61	0.3	0.30	1.27
Zinc	mg Zn/ℓ	18	6.5	4.1	1.8	14.9
Manganese	mg Mn/ℓ	17	35.8	15.5	15.0	67.0
Barium	mg Ba/ℓ	18	2.3	1.8	0.7	7.2
Nickel	mg Ni/ℓ	19	24.9	13.7	2.65	9.0
Aluminium	mg Al/ℓ	18	17.6	20.4	0.8	90.0
Total plate count	per 100 cm^3	1	12	-	-	-

TABLE 5
AVERAGE AND RANGE OF CONSTITUENTS IN PRODUCT WATER STREAM FROM SPARRO PLANT DURING PHASE II

Determinant	Units	N	Average	Std deviation	Range min	Range max
pH	-	18	5.22	0.67	4.22	7.02
Conductivity	mS/m	20	85	27	47	154
TDS	mg/ℓ	20	614	247	219	1 210
Suspended solids	g/ℓ	20	0.8	1.4	0	5
Calcium	mg Ca/ℓ	18	27	25	9	111
Magnesium	mg Mg/ℓ	18	33	16	10	68
Sodium	mg Na/ℓ	18	96	20	57	121
Chloride	mg Cl/ℓ	19	46	12	19	65
Sulphate	mg SO_4/ℓ	19	279	127	131	628
Iron	mg Fe/ℓ	18	0.17	0.06	0.1	0.3
Zinc	mg Zn/ℓ	18	0.5	0.3	0.2	1.0
Manganese	mg Mn/ℓ	17	1.7	1.0	0.8	4.2
Aluminium	mg Al/ℓ	18	1.8	2.4	0.0	7.0

Operating results

A massive amount of operating data was generated during the different phases of the project, and it is not the intention of this paper to present or discuss these operating results in detail. Such detail is presented elsewhere (Juby, 1994).

However, to illustrate the type of results obtained, some typical ones obtained during the Phase II operating period are presented below. A more detailed discussion of the operating experience specifically with regard to the novel process features of the plant is presented later in this section.

Table 4 presents the average and range of the constituents of the feed stream (a mixture of pretreated raw mine water and slurry recycle) fed to the membrane modules during the Phase II operating period, and is typical of the feed quality to the membranes during all four "operating" phases, except Phase I during which the metal ion concentrations were lower. The results are from the analytical laboratory analyses of samples taken from the plant approximately each fortnight, throughout the period of operation.

The plant product water quality is presented in Table 5. As shown, the average TDS of the product water for the whole Phase II period was 614 mg/ℓ. The maximum product water TDS shown in the table occurred during the first 500 h of operation when there were problems in the fourth bank of modules. Ignoring that value and a similar one which was also obtained during that period would lower the average product water TDS to 502 mg/ℓ. Such a product water would certainly be suitable for use as a potable water after disinfection and a slight upward adjustment of pH in order to stabilise the water.

The product water recovery during the Phase II period of operation was high, with an average of 93.3%, while the average product extraction was 37.2%.

Membrane performance

Despite the seemingly good performance of the membranes in terms of product water quality during Phase II, membrane performance was not stable, that is the salt rejection and membrane flux values could not be maintained simultaneously. Figure 4 shows the results obtained during Phase II which show a stable salt rejection but a declining membrane flux, indicative of overall membrane fouling.

As a check on individual membrane performance, module surveys were carried out. During Phase II, three full module surveys were carried out after 100 h, 2 200 h and 2 670 h on all the membrane modules in the plant. A fourth survey in which only the module flux was measured was carried out after 2 250 h of operation. The results of the surveys are summarised graphically for each row of membrane modules, in Fig. 5.

The results show that the salt rejection per module row tended to exhibit a slightly downward trend with increasing number of operating hours. Furthermore, the salt rejections of the module rows at the front of the plant were higher than those at the tail-end of the plant. Although some decrease in salt rejection would be expected due to increased salt (solute) flux resulting in the tail-end modules, the results do tend to indicate that some membrane hydrolysis was occurring. Membrane flux generally declined with operating hours, confirming the flux results shown in Fig. 4. Module surveys were found to be a valuable diagnostic tool because often the performance trends of individual modules did not match the overall performance trend of the module stack.

Briefly, the major differences in the membrane performance during Phases III and IV compared to Phase II were: During Phase III the membrane performance was the most stable of all phases of operation. Salt rejection was maintained at about 85% (because of the presence of some poor membrane modules), while the flux declined initially and then stabilised with a slight overall downward trend. During Phase IV, because of the presence of several poor membranes which had deteriorated in performance during the preservation after Phase III, the average salt rejection was less than 60%, while the flux varied between 800 and 1 000 ℓ/m^2·d. Despite the poor quality of the membranes during Phase IV, valuable performance trends were observed (as was the case in all four phases of operation), which assisted with the investigations and explanations into the possible causes of limited membrane life.

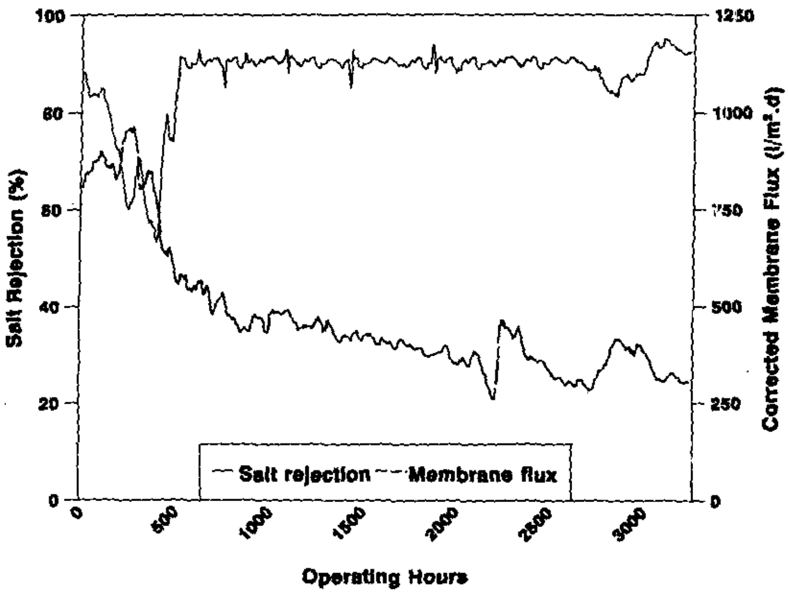


Figure 4
Variation of salt rejection and membrane flux for SPARRO plant during Phase II operating period

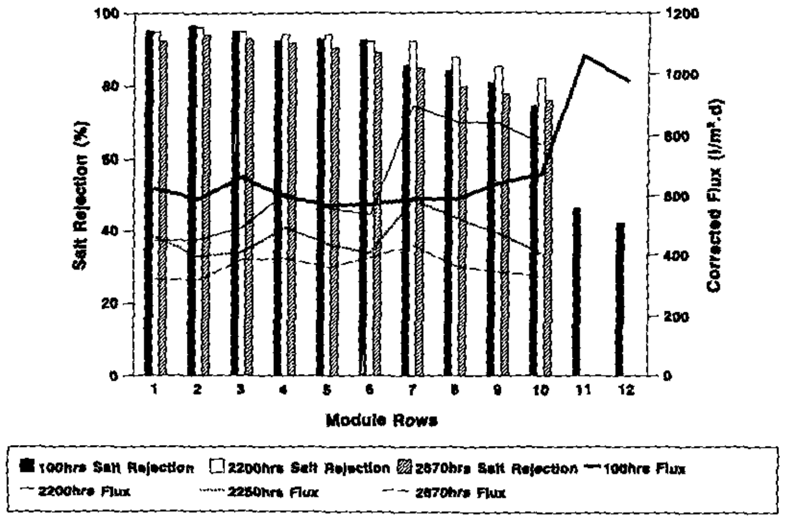


Figure 5
Variation of salt rejection and membrane flux for each row of modules in the SPARRO plant in Phase II

Summary of findings

During all phases of the work the product water quality was variable due to the changing performance of the membrane modules. However, the quality of the product water produced by the SPARRO plant was always of a quality which could be reused within the mine water circuits and in most cases it was of a quality which could be used as a potable water source, after disinfection and pH adjustment for stabilisation.

The pilot-plant studies have demonstrated that the SPARRO process is able to operate at extremely high product recovery levels, greater than 95%, while still producing an adequate quality of product water.

The SPARRO pilot plant was unable to sustain the design product extraction value of 50%, for reasons which are not clear, but are considered to be due to excessive concentration of certain constituents of the feed water which resulted in catastrophic membrane failure. The practical implication of reduced extraction values is that a larger slurry recirculation rate is required to sustain the process; a drop of 10% in extraction rate would cause an increase in the recirculation rate of over 50%. This would have both capital and operating cost implications.

However, the average extraction rate achieved during the Phase II operating period of 37% (upon which the power consumption figures were based), indicates that rates near to 40% are achievable and depending upon the raw mine water quality it is expected that extraction levels of around 40 to 45% would be achievable in practice. Sustaining the extraction level at high values is totally dependant on the performance of the membranes in terms of their membrane flux.

Readings taken during the Phase II operating period of the SPARRO plant showed that the unit power consumption of the plant was 4.82 kWh/m³ of product water produced. This compares with about 10 kWh/m³ for the 0.5 ℓ/s seeded RO plant tested in 1986.

Membrane life

During the process evaluation of the SPARRO plant all normal precautions, such as controlling the pH of the feed water to the membranes and preserving the membranes in accordance with the membrane manufacturers recommendations, were carried out. Despite this, hydrolysis of the membranes occurred to some extent during all four phases of the work, and dominated in two of the phases.

TABLE 6
BUDGET CAPITAL COST ESTIMATES FOR A SPARRO PLANT INSTALLATION TO PRODUCE 46.3 ℓ/s OF STABILISED AND DISINFECTED PRODUCT WATER (1994)

Description	Civil R(000)	Mechanical and electrical R(000)
Pretreatment (50 ℓ/s)		
Coagulation/flocculation	20	10
Clarification and chemical dosing	170	96
Cooling	24	40
Rapid gravity sand filtration	78	220
Storage in 100 m^3 tank	32	50
SPARRO Plant (46.3 ℓ/s)		
Membrane stack (4 800 + 200 spare)	-	6 075
Module racks	900	-
Building	1 300	-
Raw feed pumps (2 duty/1 standby)	10	360
Slurry recycle pumps (4 duty/2 standby)	10	570
Stainless steel pipework	-	1 000
uPVC pipework	-	200
Desupersaturation reactors (2)	120	50
Washwater tank and blowdown system	40	45
Instrumentation	-	650
Electricals and power supply	50	600
Post-treatment (46.3 ℓ/s)		
Product storage/chlorination/ stabilisation tank - 170 m^3	53	-
Lime saturator and chemical dosing equipment and instrumentation	10	30
Subtotal 1	2 817	9 996
Allowance for P&G on Civils	704	-
Allowance for Engineering	229	650
Allowance for Contingencies and Engineers site staff	628	1 499
Subtotal 2	4 378	12 145
Total estimated capital cost (excluding VAT)	**16 522**	

The results of membrane performance during the four operational phases of this study have shown that the life expectancy of the locally produced TRO cellulose acetate membranes is limited to approximately one year when operating under the conditions encountered. However, by improved pretreatment to ensure a raw mine water turbidity of less than 1 NTU and applying chlorine to produce a residual of approximately 0.5 mg/ℓ it is expected that the life of the modules can be increased to two years. The presence of a small chlorine residual is expected to provide conditions which favour fouling rather than hydrolysis, and by keeping the turbidity to less than 1 NTU the fouling properties of the raw mine water should be reduced to a minimum.

A definitive statement in respect of the cause(s) of the observed membrane hydrolysis, cannot be made. However, theoretical calculations have shown that the presence of radionuclides in the process feed water would in all likelihood deliver an α-radiation dosage to the membranes, particularly in the absence of free chlorine. While this is not fully quantifiable with the data available, it is considered that any dosage of radiation to the membrane, and in particular α-radiation, could in some way influence its chemical structure. The influence of free chlorine in the water on the chemistry of uranium ions in solution, is also seen as a contributing factor, as well as the organophilic nature of the radium ion. Overall it is speculated that radioactivity effects on the membranes as a result of treating mine service water containing radionuclides could be contributing to the observed membrane hydrolysis, but this should be investigated further in a separate study.

The membrane fouling that was observed is thought to be caused in some way by the presence of turbidity in the pretreated raw mine water. A turbidity of greater than 1.0 NTU appears to be unacceptably high. This apparent anomaly, in a process in which there is a requirement for the presence of seed crystals, can be explained by the fact that colloidal particles which contribute to turbidity are normally in the range of 0.1 to 0.3 µm and could agglomerate to 10 µm or larger in an RO system (Osta and Bakheet, 1987) during destabilisation; whereas the average size of the seed crystals was between 50 and 60µm. These factors and others could result in an accumulation of the destabilising colloidal particles near the membrane wall. This is explained in more detail in a hypothesis proposed to explain the mechanism of membrane fouling caused by the presence of colloidal material and exacerbated by the presence of free chlorine (Juby 1994).

Conceptual design for full-scale plant and cost estimates

A conceptual design was made for a full-scale plant (46.3 ℓ/s) incorporating the novel features developed and evaluated on the pilot plant. The design considerations are discussed in detail by Juby (1994).

The plant would consist of around 4 800 modules arranged in three module banks in a tapered configuration. Such a plant could be constructed as a demonstration plant which could be used as one module of a plant operating at a large mine. The projected quality of product water from the plant would be similar to the average values in Table 5. Post-treatment would include lime stabilisation as well as disinfection.

TABLE 7
ESTIMATED OPERATING COSTS FOR A SPARRO PLANT INSTALLATION TO PRODUCE 46.3 ℓ/s OF STABILISED AND DISINFECTED PRODUCT WATER (1994)

Description	Annual cost R(000)
Pretreatment	
Alum @ 10 mg/ℓ maximum	33
Polymer @ 0.5 mg/ℓ	10
Acid/alkali for pH correction	18
Chlorine @ 2 mg/ℓ maximum	10
Electrical power @ 10 c/kWh	11
General maintenance based on capital cost	22
SPARRO plant	
Membrane replacement - 2-year life	1 404
Electrical power @ 4.82 kWh/m^3	701
Membrane washing	20
Plunger/packing/valve replacement	24
General maintenance based on capital cost	186
Post-treatment	
Lime addition @ 25 mg/ℓ	11
Chlorine @ 2 mg/ℓ maximum	10
Electrical power @ 10 c/kWh	8
General maintenance based on capital cost	2
Staff	
One plant manager - time	144
Four plant operators - ½ time	
One plant cleaner - full time	
Total estimated operating cost (excluding VAT)	**2 614**

Estimated capital and operating cost

The costs presented in this section have a base date of 1994. The cost of civil works is based upon 1994 tender prices for similar installations and the cost of mechanical and electrical plant is based upon estimates provided by various equipment suppliers.

Capital cost

Based upon a conceptual full-scale SPARRO installation (consisting of pretreatment, the SPARRO plant itself and post-treatment), estimates of the capital cost were made.

The estimated budget costs are summarised in Table 6, and as shown total approximately R16.5 m (excluding VAT). It can be seen that the pretreatment plant accounts for roughly 4.4 % of the total cost, while the post-treatment plant accounts for only about 0.5% of the total cost. The biggest single capital cost would be for the initial membrane inventory, in this case taken to be 4 800 modules for the plant plus 200 spare modules, which has a capital cost of over R6 m, almost 40% of the total plant cost.

Operating costs

The operating costs were calculated from the experience gained during the pilot-plant study and the estimated costs are presented in Table 7.

As was the case with the capital costs, the cost of the membranes would be the major contributor to the overall operating costs, in this case accounting for over 53% of the total annual estimated cost of about R2.6 m. The second biggest cost would be for the electrical power of the SPARRO plant; about R700 000 annually or about 27% of the total cost.

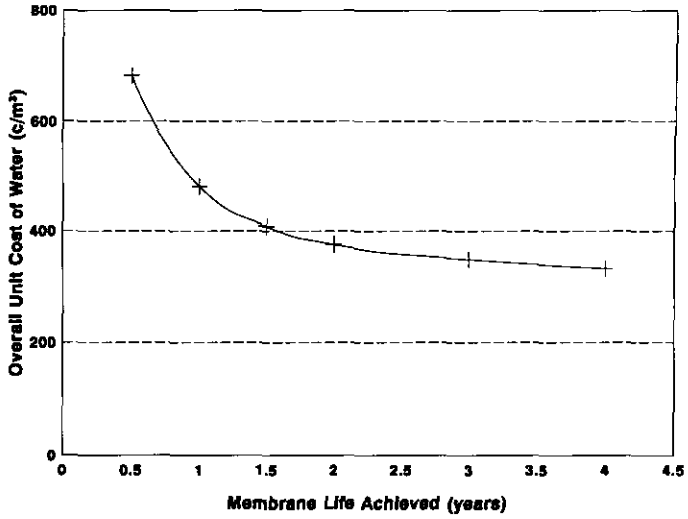
Overall costs

To provide an order of magnitude estimate of the annual costs associated with repayment of a capital sum of R16,522 m., it was assumed for the purpose of this work that the civil portion would be repaid over a period of 25 years at an annual gross interest rate of 15%, while the mechanical and electrical portion of the loan

TABLE 8
SUMMARY OF TOTAL ANNUAL COSTS AND UNIT COSTS FOR CAPITAL AND OPERATING REQUIREMENTS (1994)

Description	Annual cost R(000)	Unit cost (c/m^3)
Interest and redemption on rapital		
Civil (25 years @ 15%)	677	48
Mechanical and electrical (15 years @ 15%)	2 076	148
Estimated operating costs	2 614	187
Total	**5 367**	**383**

Figure 6
Sensitivity curve for the variation in overall unit cost of product water with membrane life achieved



would be repaid over 15 years at the same interest rate.

The annual costs associated with the interest and redemption on capital as described above, are shown in Table 8, together with the total annual operating cost from Table 7. In addition to the annual costs, the costs are also expressed as a unit cost, calculated from the total volume of product water produced during an average of 350 d of operation at the design output each year. As the figures in the table indicate, the total cost of product water, calculated by this method would be approximately 380 c/m^3, and that the contribution by the operating and capital cost portions would be roughly equal - the capital cost portion slightly the greater of the two.

Influence of membrane life on overall cost

To illustrate the sensitivity of the membrane life achieved on the unit cost of product water, calculations were carried out for a range of membrane life periods between 6 months and 3 years. The effect of the membrane life on the overall unit cost is shown in Fig. 6, which illustrates the exponential relationship between membrane life and the overall unit cost of the product water.

Discounting for the cost of purchasing water from a local water authority and for the selling price which could possibly be secured for the high quality gypsum by-product from the plant, the extra cost that the mine would have to pay for the potable water, would be approximately 190 c/m^3 (1994).

Conclusions

The four operating phases of the SPARRO process, under different operating conditions, proved the mechanical integrity of the process configuration and showed that from an engineering point of view the process is technically viable.

All of the novel features of the process were evaluated and found to be practical alternatives to overcoming the process difficulties with the previous systems. The unit power consumption of the process was evaluated and found to be approximately half of the value of the previous plants.

The plant was able to operate at very high product recovery values (above 95%) while still producing a product water which could be re-used in the mines. In most cases the product water produced was of potable quality (accepting that final disinfection and pH correction would be required). Product extraction levels varied but were generally above 37% and values of between 40 and 45% are expected to be achieved in practice.

A cost estimate was made, based on a conceptual design of a full-scale SPARRO plant (46.3 ℓ/s) to treat a scaling mine water and produce a disinfected and stabilised potable water. The capital cost of the plant was estimated to be about R16.5 m (1994) (excluding VAT). The estimated gross operating cost, which includes interest and redemption on the capital sum amounts to 383 c/m^3 (1994), while the nett operating cost after discounting for the value of the product water and high quality gypsum is 190 c/m^3.

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