increased acceptance of membrane separation technology by all sectors of the industry, even by the more conservative sectors such as the mining sector.

Wastewater treatment and waste minimization has become a rapidly growing area for membrane separation technology. Although membrane separation technology has become an established separation technique that is widely used in drinking water production; it is still considered an emerging technology in the mining industry. This technology offers many potential benefits to the mining industry, for instance, high percentages of water recovery from a wastewater will significantly reduce the volume of the stream requiring further treatment. This could translate to smaller neutralization plants with lower capital costs, particularly if small membrane plants are developed to remove contaminants where they are the most concentrated in the process.

The versatility and modularity of the membrane separation technology facilitates its introduction into practically any process with a suitable feed stream. With increasingly stringent environmental regulations, the use of membrane separation technology for achieving acceptable environmental performance may be unavoidable in the future. Although membrane separation does not always provide the ultimate solution to water and wastewater treatment, it can be economically added to conventional liquid waste treatment technologies (Weber and Bowman, 1986).

3. MEMBRANE BASED WATER AND WASTEWATER TREATMENT TECHNOLOGIES

In the most general sense, a membrane is a thin barrier that permits selective mass transport. The rate of mass transfer across a membrane depends on different driving forces that influence permeation. Membranes, depending on their composition and conditions of fabrication, can have different types of structures, but all membranes of practical industrial application are invariably of asymmetric or composite structures (Porter, 1990).

Separation in membrane processes take place as a result of differences in the transport rates of different chemical species through the membrane matrix, which for the applications of interest to the scope of this review, is usually polymeric or ceramic.

3.1. Membrane Processes

There are many requirements for separations in the mining industry but they can be classified into two major areas; 1) where materials are present in different phases, and 2) where different chemical species are present in a single phase. Membrane separation can be applied to both scenarios through their ability to; change the composition of solutions, selectively permeate certain species while rejecting others, regulate permeation of certain species and conduct electric current.

Membrane based separation processes have become viable alternatives to many of the physical methods of separation in a given process, such as selective adsorption, absorption, solvent extraction, distillation, crystallization, etc. Membranes are primarily used as a means of separation, recovery and purification, although they also have other applications that are beyond the scope of this review.

Membrane filtration is a separation process that uses a semipermeable membrane to divide a liquid or gas feed stream into two portions: a permeate, which contains the material passing through the membranes, and a retentate, which consists of the species left behind (Mallevialle *et al.*, 1996). Transport of the permeating species through the membrane matrix is achieved by the

application of a driving force across the membrane which provides a basis for the classification of membrane separation processes. This classification is based on the type of driving force which drives mass transport across the membrane, which can be mechanical (pressure), concentration (chemical potential), temperature, or electrical potential (Porter, 1990). Membrane filtration can be further classified in terms of the size range of the permeating species, the mechanisms of rejection, the driving forces employed, the chemical structure and composition of membranes, and the geometry of construction. The most broadly applied types of membrane filtration are pressure driven processes and include reverse osmosis (RO), nanofiltration (NF), ultrafiltration (UF) and microfiltration (MF).

3.1.1. Pressure driven membrane processes

Table 3.1 gives an overview of the four pressure driven membrane processes mentioned above. It should be noted that there are more commercially available polymer choices for each of the membrane processes than those listed in Table 3.1.

Table 3.1. Comparison of the four pressure driven membrane processes (Wagner, 2001).

Membrane	Reverse Osmosis	Nanofiltration	Ultrafiltration	Microfiltration
	Asymmetric	Asymmetric	Asymmetric	Asymmetric Symmetric
Thin film Thickness	1 micron 150 micron	1 micron 150 micron	1 micron 150-250 micron	10-150 micron
Rejection of:	High and low molecular weight compounds, NaCl, glucose, amino acids	High molecular weight compounds, mono-, di- and oligosaccharides, polyvalent ions	Macromolecules, proteins, polysaccharides, vira	Particles, clay, bacteria
Membrane materials	Cellulose acetate (CA) thin film	CA, thin film	Ceramic, PS, PVDF, CA, thin film	Ceramic, PS, PVDF, CA
Pore size	< 0.002 micron	< 0.002 micron	0.02-0.2 micron	0.02-4 micron
Module configuration	Tubular Spiralwound Plate-and-frame	Tubular Spiralwound Plate-and-frame	Tubular Hollow fiber Spiralwound Plate-and-frame	Tubular Hollow fiber
Operating pressure	15-150 bar	5-35 bar	1-10 bar	< 2 bar

Reverse Osmosis (RO) or hyperfiltration is the most widely used pressure driven membrane separation process. RO membranes are the tightest membranes in liquid/liquid separation. RO is aimed at the separation of ionic solutes, metals, and macromolecules from aqueous streams such as industrial wastewaters, mine water and mill effluents. Water is, in principle, the only material that would pass through the membrane; essentially all dissolved and suspended materials, organic and inorganic, are rejected by RO membranes. The operating pressure of RO depends on the osmotic pressure of the solution and typically is in the range of 15-150 bar. Separation of species is a function of the shape and size of permeating species, their ionic charge, the membrane material properties and composition and its interaction with the permeating species. The more open types of RO membranes are sometimes confused with nanofiltration (NF) membranes.

Nanofiltration (NF) systems typically operate at lower pressures than those used for RO. NF systems have higher fluxes and their permeate quality is lower than that achieved with RO, but they have a selectivity that is not possible with RO. Since NF systems operate at lower pressures, they have lower energy consumption than conventional RO systems. True NF rejects multivalent ions and dissolved materials such as sulphate, phosphate, Mg and Ca, according to the size and shape of the molecule. The molecular weight cutoff of NF membranes is around 200 Daltons. Typical rejections with a NF membrane, at 5 bar and 2000 ppm of TDS, are 60% for NaCl, 80% for calcium bicarbonate and 98% for magnesium sulphate.

Some of the specific applications of NF are removal of Total Organic Carbon (TOC), hardness, salinity, radium, Total Dissolved Solids (TDS) and multivalent ions from surface water, groundwater and wastewater. In some plating operations, NF is used for the separation of EDTA-metal complexes and can deliver almost complete sulphate rejection from water (Scott and Hughes, 1996).

Ultrafiltration (UF) is another pressure driven membrane separation process. UF membranes are used to remove particles in the size range of 0.001-0.02 μm. Solvents and salts of low molecular weight pass through the UF membranes while larger molecules are rejected or retained. The primary application of the UF process is the separation of macromolecules, but they can be used

Due to the complexity of transport phenomena and interactions between membrane and contaminants in membrane filtration, the design practices of membrane filtration remain largely empirical. Consequently, site-specific bench and pilot testing are often recommended and necessary to assess treatment feasibility and to provide process parameters for plant scale-up. Common considerations and experimental protocols to conduct such tests were summarized by Mallevialle *et al.* (1996).

A membrane plant is sensitive and very specific to its feed composition and conditions. Any change in the feed composition beyond the limits set in the original plant design could have significant negative impact on its operational efficiency and performance. Lack of attention to feed characteristics, adequate pretreatment requirements and membrane module design, membrane material selection, cleaning methods and to proper selection of membrane system component materials has been the cause of numerous failed attempts in membrane separation applications.

3.1.4. Mass transport and fouling control

Membrane fouling is probably the most significant process problem that is encountered in mining applications of membrane separation and the major cause of membrane failure, and severely impacts productivity, effluent quality and membrane life. It has been observed that less soluble salts, dissolved organic compounds, colloids, fine particles, and biological growth can cause membrane fouling (Braghetta *et al.*, 1997a,b; Cho *et al.*, 1999; Fu *et al.*, 1994; Jacangelo *et al.*, 1995; Wiesner *et al.*, 1989; Zhu and Elimelech, 1995).

The major problem with the application of membrane separation to AD treatment occurs as a result of calcium sulphate and ferric hydroxide. Membrane fouling with iron can be managed by the introduction of additives and by maintaining the pH below 3. Calcium sulphate fouling has been a more challenging and limiting factor, although with high flux membranes and proper pretreatment and feed conditioning, fouling could be managed. Additionally, streams that are treated with cationic or anionic flocculants pose different and challenging fouling problems. As

the performance problems arise and escalate, operating costs increase and process economics become compromised (Zibrida et al., 2000).

The type of foulant most commonly encountered in membrane separation is scaling due to inorganic precipitation and fouling. Some of the common forms of mineral scale are calcium carbonate, magnesium carbonate, calcium sulphate, calcium phosphate, barium sulphate, strontium sulphate, iron hydroxide and silicon dioxide (silica). Other types of membrane fouling include: colloidal fouling, biological fouling, and organic fouling. The latter two are not as significant in AD treatment applications.

Figure 3.6 shows the SEM photo of a deposited amorphous solid residue on a high flux RO membrane coupon that was tested at CANMET-MMSL using untreated AD, under feed pressures of 100-500 psig, without an antiscalant. The total sulphate concentration in the feed was 3500 ppm while the levels of calcium and iron were 290 ppm and 380 ppm respectively. The fouling layer shown in Figure 3.6 did not impact the permeate flux under operating pressures up to 400 psig. At 500 psig, a decline in the permeate flux and permeate quality was observed which was probably due to compaction of the cake layer. Figure 3.7 shows the SEM photo of a crystalline fouling layer, primarily $MgCO_3.2H_2O$ crystals, which were formed during another CANMET-MMSL test. In this test, a high rejection RO membrane was tested with a mine effluent containing 13 g/L of TDS, 10 g/L of sulphate and approximately 600 ppm of magnesium at pH 9.7 and an operating pressure of 450-600 psig.

RO and NF membranes are typically able to remove 90-99% of TDS and other contaminants in an effluent. If such high rejections are combined with high water recoveries, the likelihood of membrane fouling and serious drops in water flux increases. The feed components that contribute to fouling are calcium carbonate, calcium sulphate, silica, iron, and other species that exceed the saturation index (Zibrida et al., 2000; Koseoglu and Guzmann, 1993).

precipitation of the salts in which case the requirement for a clarifier in the treatment process could be eliminated. This process would result in the production of a clean permeate stream and solid precipitated salt.

Membrane processes such as UF for solid separation have been used as a pretreatment step to RO and NF membranes in order to reduce the TSS of the feed stream for spiral wound and hollow-fiber applications (Porter, 1990). UF and MF membranes have been used for solid-liquid separation in hybrid processes where an adsorbent, such as activated alumina, is initially used to adsorb a contaminant and separated by the membrane (Mortazavi *et al.*, 1999).

3.1.6.2. Inorganic contaminants

The removal of inorganic contaminants by NF and RO remains the largest application in water treatment. A survey showed that there were more than 4000 land-based RO plants worldwide in 1989 with a combined desalting capacity of approximately 3.8×10^6 m^3/d (AWWA Membrane Technology Research Committee, 1992). This only includes plants with capacities larger than 95 m^3/d. The feasibility of the application of RO and NF and even UF for the removal of hardness, nitrate, ammonia, heavy metals and oxyanions has been demonstrated in the published literature (Rautenbach and Groschl 1990; Waypa *et al.*, 1997). The USEPA considers RO as a best available technology to meet anticipated regulations for small surface-water plants without existing facilities and groundwater treatment plants.

3.1.7. Hybrid membrane processes

UF and MF are very effective methods of solid–liquid separation and operate at low pressures. UF and MF have been combined with other conventional treatment processes to remove dissolved species. Examples include the use of membrane filtration and powdered activated carbon, adsorption and MF. The use of polymers and resins with UF or MF and membrane bioreactors has shown great promise in water and wastewater treatment (Mortazavi *et al.*, 1998, 1999; Legualt and Tremblay, 1994).

8. CASE HISTORIES

Several summaries of large scale and pilot processes have been presented above. The selected case studies in this section provide examples of membrane separation performance, associated capital and operating costs, where available, and a comparison of membrane separation with conventional treatment options.

These case studies cover different membrane applications in various scenarios and provide comparative examples of membrane and conventional wastewater and effluent treatment technologies. All the examples show that the application of membrane separation technology to mitigation of AD and water management in mining and metal processing operations provides opportunities for water recovery and recycle. As well, the examples show that membrane separation could provide the possibility of improving process economics and performance while exceeding environmental water discharge criteria.

8.1. ASARCO Inc. Globe Plant in Denver Colorado

This case study presents a summary of a feasibility study conducted at Asarco's Globe Plant to reduce the operating costs, sludge volume and discharge water quality of the existing wastewater treatment facility. A number of process configurations were examined including a membrane separation polishing system (Green *et al.*, 1993).

Asarco Inc. is a large producer of non-ferrous metals such as copper, zinc, lead, silver and gold. The Asarco Globe plant has been a metal refining facility since 1886, producing a wide range of non-ferrous metals. In 1986, the company installed and operated a chemical precipitation system to treat wastewaters containing arsenic, selenium, lead, zinc, cadmium, nickel, iron, manganese, copper, chromium and silver.

The Asarco Globe Plant system utilized ferric sulphate to first remove arsenic and selenium followed by lime addition for the removal of manganese, cadmium and zinc. The pH is adjusted

to from 4 to 9.8. The system operating temperature was maintained at 10-20°C. No complexing agents were added to the wastewater.

The wastewater feed entered the process train into a surge/blend tank where sodium carbonate was added (1), followed by ferric sulphate (2), the sludge was then filtered in press filters and dewatered and sent to sludge processing. Lime and sodium sulphide at pH 9.8 were added (3), followed by filtration and sludge dewatering. The final effluent water pH was adjusted to pH 7.5 before final discharge. The total operating cost of the wastewater treatment, including the depreciated initial capital cost was $58.34 USD (in 1993) per 1000 US gal of treated wastewater. Figure 8.1 shows the block diagram of the precipitation process. Table 8.1 shows the concentration of the contaminants in the feed and the discharge water from the precipitation system.
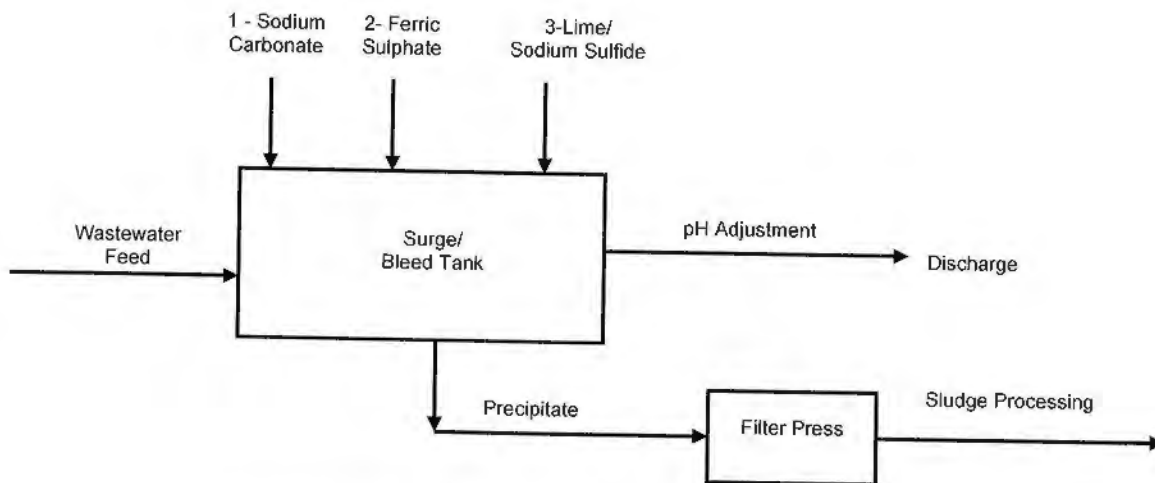

Figure 8.1. Block diagram of Asarco's precipitation process

Table 8.1. Asarco's Globe Plant precipitation system performance.

Component	Wastewater Feed (mg/L)	Treated Water (mg/L)
pH	4.0	~7
TDS	3000-10000	<3000
As (mg/L)	10.1	0.024
Se (mg/L)	0.056	<0.010
Cd (mg/L)	14.5	0.10
Zn (mg/L)	35.5	0.35
Pb (mg/L)	3.07	<0.050
Ni (mg/L)	0.060	0.025
Fe (mg/L)	0.986	0.100
Mn (mg/L)	3.33	0.120
Cu (mg/L)	0.07	0.020

In 1991, the company investigated the use of various types of encapsulated biomass for the extraction of the contaminants present in the Asarco wastewater; however, the biomass did not have the capacity for the levels of contaminants present. In a full-scale pilot plant study, the wastewater, after pretreatment, was first subjected to a membrane separation stage for reduction of all contaminants. This was followed by a biomass media extraction stage for heavy metal polishing, and finally a mineral media extraction stage for arsenic and selenium polishing. The final concentrate, which had a significantly reduced volume, was then put through the existing precipitation system that primarily treated the concentrate from the membrane stage and the stripping solution from the biomass polishing stage. Figure 8.2 shows the block diagram of the membrane separation system. Table 8.2 shows the compositions of the feed and the discharge water produced from the membrane separation system.
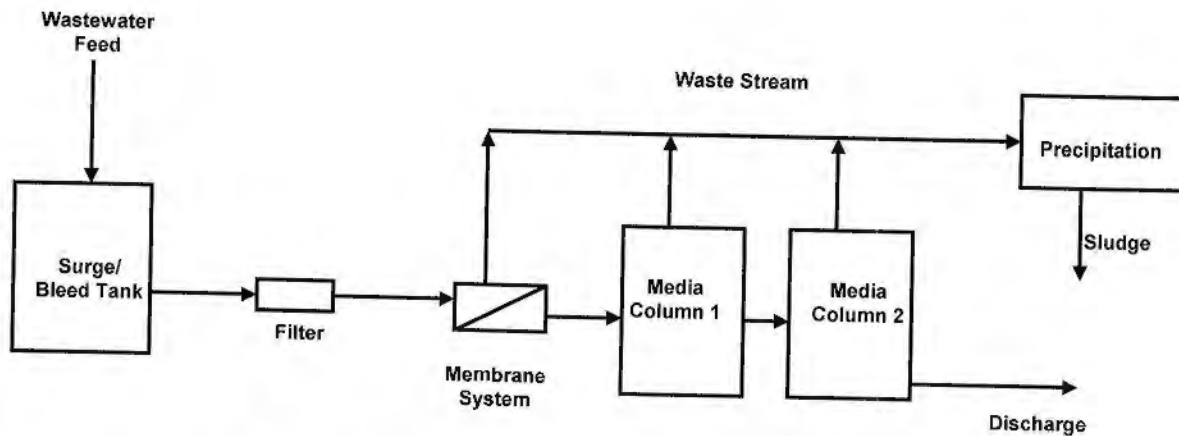
Figure 8.2. Block diagram of Asarco's membrane separation process

Table 8.2. Asarco's Globe Plant membrane separation system performance.

Component	Wastewater Feed (mg/L)	Treated Water (mg/L)
pH	4.0	~7
TDS	3000-10000	<1000
As (mg/L)	10.1	0.006
Se (mg/L)	0.056	<0.010
Cd (mg/L)	14.5	0.02
Zn (mg/L)	35.5	0.010
Pb (mg/L)	3.07	0.050
Ni (mg/L)	0.060	0.050
Fe (mg/L)	0.986	0.10
Mn (mg/L)	3.33	0.050
Cu (mg/L)	0.07	0.012

Water recovery from the treatment process was 80% and the treated water was discharged with contaminant levels below the discharge standards. The concentrated water from the membrane separation stage and the metal stripping from the biomass polishing stage were treated in the existing precipitation system. An overall 85% sludge reduction was realized.

Some of the difficulties encountered were; the variability of the feed water metal, the variability of contaminant concentrations and the presence of algae and bacteria in the feed water. These issues were handled by providing adequate pretreatment through the addition of an antiscalant and by adjusting the frequency of bio-medial regeneration cycles.

The operating cost of treating 1000 US gal of wastewater with the membrane system, including the depreciated capital costs, was reduced to $15.67 USD (1995) from $58.32 USD (1993).

Table 8.3 presents the costs and benefits of the precipitation and membrane system.

Table 8.3. Cost comparison between the Asarco's precipitation and membrane systems per 1000 US gal treated wastewater. Costs are based on 1995 dollars unless otherwise specified.

Costs Items	Precipitation System	Membrane Separation System
Water Quality	Meets Discharge Criteria	Meets and Exceeds Discharge Criteria
Capital Cost (USD)	$1,000,000 (1986)	$300,000 (1993)
Reagent Cost (USD) (per 1000 US gal)	$9.88	$0.93
Direct Operating Cost (USD) (per 1000 US gal)	$10	$3.33
Sludge production (per 1000 US gal)	160 lbs	24 lbs
Total Treatment Cost (USD) (per 1000 US gal)	$58.34 (1993)	$15.67 (1995)

As the above table shows the membrane system reduced the amount of the generated sludge by 85% and reduced the operating cost by 73%, while producing better discharge water quality than the precipitation system.

8.2. Mexicana de Cananea mine in Cananea, Mexico

Open pit mines can accumulate enormous amounts of water, even in dry areas, from water run off, rain and snow melt and intrusion of ground water. Water accumulation can severely impact the hydrogeology of the area, can create a risk of flooding and may result in an interruption of mining operations. This situation was observed at the Berkley pit in Butte, Montana.

The Mexicana de Cananea mine, located in northern Mexico, was facing closure due to insufficient water and a threat of flooding of the operational parts of the mine because of an accumulation of water in the Cananea pit. The Cananea pit started accumulating water in the

1980's. At the beginning of mine operation, the company leached minerals directly in the pit and then utilized it as a reservoir for the pregnant leachate from dump leaching operations around the pit. As well, in the years prior to the study, the pit was used to hold the excess water from the hydrometallurgical system which was transferred to the pit in order to avoid spills. Eventually, the pit solution volume reached 17 Million cubic meters (4.49×10^9 US gal) with an average copper concentration of 0.70 g/L. At this point, pit water levels and evidence of channeling within the pit had jeopardized the operation of the mine (Harrison Western, 1997).

The mine decided to install a full-scale membrane plant to control the water levels. The plant was successfully used to; remove water from the pit, recover water from the tailings thickener, increase the copper concentration in the acid leach water feed from the Cananea pit to the copper extraction plant, remove excess water from the leach circuits, and produce clean water for mine process water application. The membrane system was able to; dewater the leach solution, increase its copper concentration, recover clean water for reuse, reduce acid costs, and recover copper from excess raffinate and electrowinning electrolyte.

In 1995, initial pilot tests were completed with a small 10 - 15 US gpm membrane system to assist with the design of a system for a 4000 US gpm membrane plant capable of generating a 2000 US gpm concentrate stream with 1.6 g/L copper and 2000 US gpm of clean permeate suitable for reuse. The objectives of the large scale membrane plant were to; improve the feed copper concentration in the feed to SX/EW plant, increase the cathode copper production by more than 14%, create savings of $212K in process water cost and $27K in sulphuric acid costs, and reduce the pit water level.

Table 8.4 shows the concentrations of different components in the Cananea pit solution. The pit solution was processed in a full scale 4000 US gpm membrane plant. The pit solution was at pH 1.1, with a total TDS of 113,000 mg/L.

Table 8.4. Cananea pit solution composition.

Component	Concentration (mg/L)
Al	4400
Ca	220
Cu	800
Fe	21600
Mg	220
Zn	110
SO₄	85500

The full scale membrane plant was commissioned in February 1997. Figure 8.3 shows the general diagram of the membrane plant and its performance with average stream flows. The copper recovery from the pit solution was reported to exceed 99% and the water recovery was about 43%. The concentrate solution reached an average copper concentration of 1.2 g/L. The permeate from the membrane plant was combined with the tailings solution and sent to the grinding and ore flotation circuit. The overall water balance of the pit indicated that its level would continue to drop by about 3.5 meters per year, which is equal to approximately 1 billion gallons per year.
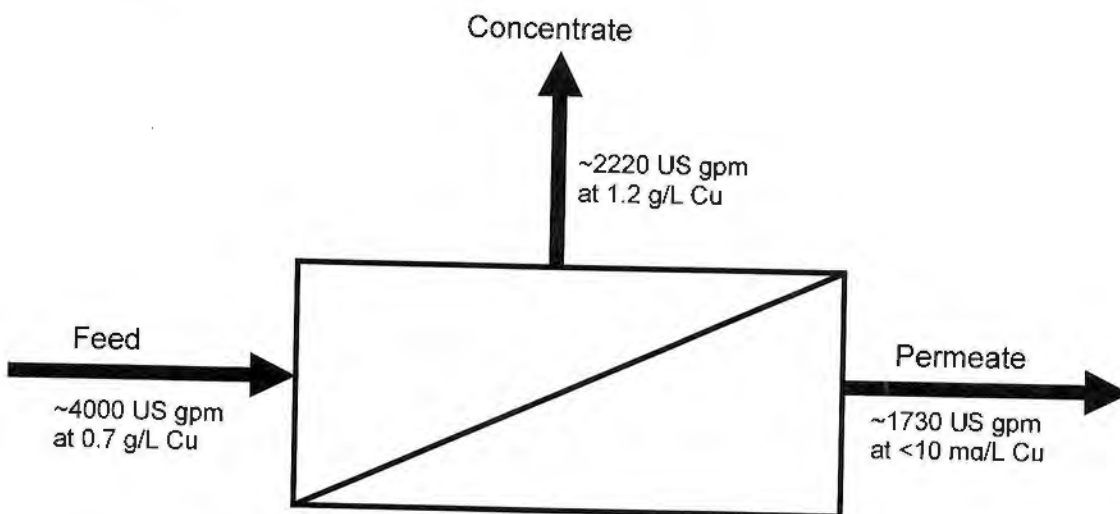

Figure 8.3. General diagram and performance of the Cananea membrane plant showing the average stream flows and concentrations

Process Economics

The typical capital cost for the membrane plant according to Harrison Western (1997) was $1.5-$2.5 USD/US gal/day. The operating costs included; power consumption, prefiltration and pretreatment operations, chemicals, membrane cleaning, and membrane replacement. Operating credits were the value of the copper increase in the feed to the SX/EW plant and the recovered permeate. The typical operating costs were $1.00-$2.00 USD/1000 US gal water recovered. The payback period for the plant was 1-3 years. The cost for a lime precipitation system, accounting for the loss of copper, would be approximately $5.00 USD/1000US gal removed.

8.3. Desalination and Reuse of Acidic Drainage and Ash Water

This case study reviews the membrane plant put in place at the Sasol Technology Limited operation in Secunda, Republic of South Africa. At this plant, Tubular and spiral RO and EDR were used to treat AD and ash water and production of boiler feed water (Nieuwenhuis *et al.*, 2000). Sasol Technology Ltd. was able to design and operate a successful membrane operation which reduced water intake volumes.

The processes that were installed consisted of a tubular RO (TRO) system followed by a second RO system with spiral wound modules (SRO) for the recovery and treatment of ash water. For the treatment of AD, a combination of EDR and SRO was used to successfully convert AD to boiler feed water. The operating costs of the two processes were similar at R3.50/m^3 of the final boiler feed water.

System Design

TRO-SRO system Design

Figure 8.4 shows a block flow diagram of the process used for the recovery of ash water and boiler feed water production.

Figure 8.4. Block flow diagram of TRO-SRO plant for recovery of ash water

In this process the pretreatment of the feed included pH adjustment to pH 3.0 to 6.5 with sulphuric acid. Suspended solids (SS) were then reduced to 40 mg/L by sand filtration and an antiscalant was added. Chlorination was used and the ash water was heated to 27°C to minimize biofouling.

The membrane plant (TRO) consisted of 11 units, each of which was made up of 80 parallel branches with 10 modules in series, for a total of 800 modules with a total capacity of 230 m³/h. Flow reversal and sponge balls were used in thirty minute intervals for system cleaning. The plant was operated at a constant water recovery rate. Each unit in the plant produced 23 m³/h of permeate. The concentrate stream was treated using three falling film evaporators.

The permeate from the TRO plant was further treated with the SRO units. The SRO units had a production capacity of 128 m³/h and were operated at 90% recovery. The membranes used were high rejection polyamide RO membranes in a 10:5:3 configuration.

Performance criteria for the pretreatment process included targets for pH, SS removal, feed temperature and microbial counts. The operation did not always meet the performance criteria and problems were reported with sand filtration and hypochlorite dosing. During the operation, large variations in the SS concentrations in the feed stream were encountered with an average SS concentration of 146±33 mg/L achieved. The sand filters were able to remove an average of 50% of the SS content of the feed stream.

TRO performance criteria included targets for permeate quality, salt rejection, standard membrane flux (flux at 4000 kPa (580 psig) and at 25°C) and CIP frequency. The feed to the TRO had a great deal of variability and the concentrations of feed constituents such as organic compounds, chloride, sodium, barium and calcium were at a higher concentration than the plant's original design values. The TRO was, however, effective in treating the feed and generating a permeate of a relatively constant quality. Table 8.5 shows the feed and permeate compositions and variability. The TRO system's standard flux was 524 ± 65.5 L/m^2.day. The reported variation in flux was due to the feed variation. No membrane fouling was observed and the average salt rejection, calculated based on conductivity measurements, was 94.5%. It was seen that the operation of an effective CIP regime was effective in preventing fouling in the TRO plant. The high TOC levels in the feed were an issue with respect to biofouling; however, the risk was managed by the introduction of an inorganic biocide.

Table 8.5. TRO feed and permeate compositions.

Component	Feed Concentration (mg/L)	Permeate Concentration (mg/L)
TDS	3998 ± 786	96 ± 38
Ca	422 ± 94	4.6 ± 3.2
Ba	0.2 ± 0.09	>0.2
Na	917 ± 79	48 ± 7
Cl	828 ± 238	44 ± 4
SO$_4^{2-}$	3254 ± 842	7.5 ± 5.1
F	18 ± 4.9	2 ± 0.3
TOC	52 ± 14	<10

The permeate from the TRO units was sent to the SRO units for further upgrading of the permeate to boiler feed water. Table 8.6 illustrates the performance criteria for the SRO plant.

Table 8.6. Performance criteria and the average performance of the SRO units of the TRO plant.

Component	Target	Average
Water Recovery (%)	90	88 ± 8.8
Conductivity (µS/cm)	<30	26 ± 6.1
CIP/Train/month	1	2.4
Flux (L/m².h)	25	23.5 ± 1.7
Feed Pressure (kPa)	1350	1390 ± 159

The major issue that was reported with the standard CIP was that it was not completely effective in mitigating biofouling which resulted in the lower reported flux in the SRO plant.

EDR-SRO system Design

The EDR-SRO plant was designed for the treatment of AD and the generation of boiler feed water. Figure 8.5 presents the process block diagram of the EDR-SRO plant.
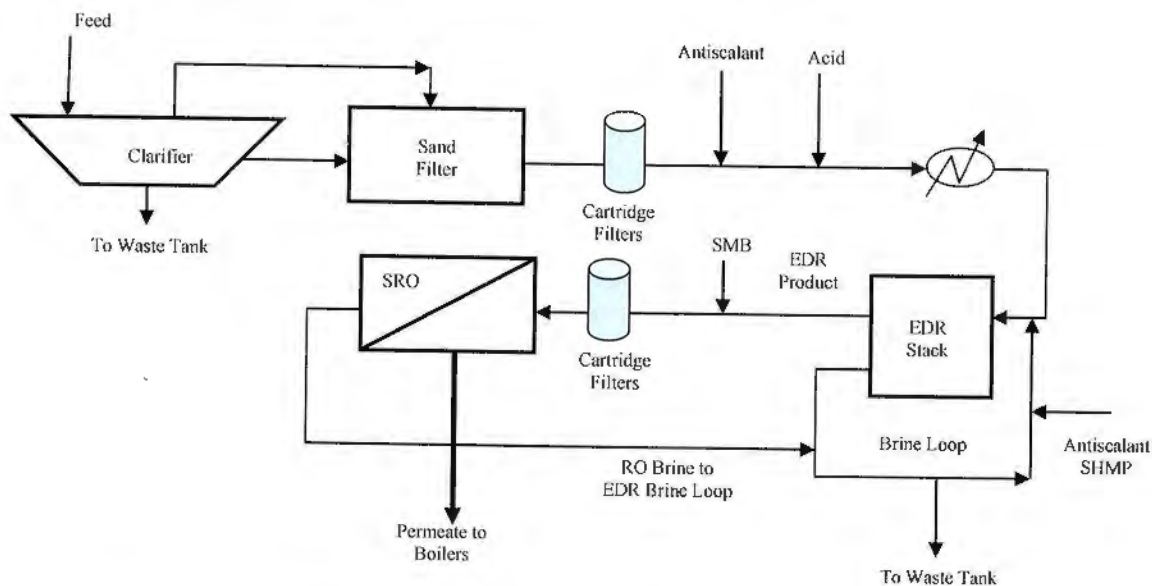

Figure 8.5. Block flow diagram of EDR-SRO plant for recovery of water from AD

Performance criteria for the EDR pretreatment processes were based on Silt Density Index (SDI) and on the iron and manganese concentrations leaving the cartridge filters units. The pretreatment train was able to produce an EDR feed with an SDI of less than 5. However the iron and manganese concentrations were higher than the design values, which was related to a decrease in mine water pH from 7.9 to 5.8, this issue was resolved by the addition of soda ash. Table 8.7 shows the targets and the average performance of the pretreatment train.

Table 8.7. Performance data for the EDR plant pretreatment process train.

Factor	Target	Maximum Value (mg/L)	Average (mg/L)
Fe input		0.62	0.37
Fe output	0.2	0.18	0.1
Mn input		0.81	0.42
Mn output	0.1	0.53	0.03
SDI	<5	>5	3.9

The EDR performance criteria were based on the frequency of EDR stack cleanings, salt rejection and water recovery. Table 8.8 shows the performance data for the EDR stacks. Based on the data presented, the EDR stacks performed well but the salt rejection was lower than the design target value which suffered because of higher water recovery.

Table 8.8. Performance data for the EDR stacks.

Component	Feed Concentration (mg/L)	EDR Permeate Concentration (mg/L)
TDS	3998 ± 786	1435 ± 438
Ca	422 ± 94	36 ± 15
Na	917 ± 79	358 ± 151
Cl	828 ± 238	121 ± 42
SO_4^{2-}	3254 ± 842	701 ± 487
TOC	2.12 ± 1.1	1.98 ± 0.4

The permeate from the EDR stacks was then sent to the SRO units. The performance criteria for the SRO units were permeate quality, water recovery, CIP frequency and permeate flux. Table

8.9 shows the performance data for the SRO section of the EDR plant. Given the calcium and sulphate content of the SRO feed, a CIP was implemented that used 1% EDTA, 1% tri-sodium-polyphosphate, and sodium-dodecyl-sulphate at pH 10.5 and 35°C.

Table 8.9. Performance data for the SRO units of the EDR plant.

Component	Target	Average
Water Recovery (%)	85	79 ± 1.6
Conductivity (μS/cm)	80	33 ± 9
CIP/Train/month	1	2.5
Flux ($L/m^2.h$)	25	20.1 ± 3.2
Feed Pressure (kPa)	1350	1350

The overall EDR plant water recovery was 76%. The SRO concentrate was recycled to the EDR stacks as brine makeup.

8.4. Canonsburg, Pennsylvania Uranium Mill Tailings Remedial Action Program (UMTRA) Site

Tiepel and Shorr (1985) applied their system design to a wide number of treatment scenarios including the groundwater clean-up of operation at the Canonsburg, PA, Uranium Mill Tailings Remedial Action Program (UMTRA) site.

Site Description

The Canonsburg site was operated as a vanadium and radium processing facility from the 1900's through to the 1920's. In 1933, the facility was utilized to extract uranium, vanadium and radium from various residues, ores and concentrates. It primarily processed uranium for the U.S. Atomic Energy Commission (AEC) until 1953 and solid and liquid wastes were deposited and discharged at different locations on the site. Some remedial action was performed on site by burying solid waste under a layer of steel mill slag.

Remedial Action

The remedial action consisted of the identification of all contaminated areas and excavation of the contaminated soils into a central engineered encapsulation cell. Groundwater from the contaminated soil and tailings excavation area and any surface runoff from the exposed