

Modified Nano-Filtration Membrane Treatment of Hyper-saline Goldfields Water- an Overview and Benefits to Gold Plant Operation and Economics

B Tapley¹, N Stoitis², L Lien³

ABSTRACT

The treatment of hypersaline waters originating from numerous gold projects within the Western Australian Goldfields region via membrane technology has been assessed. A combination of Ultra-filtration (UF) and Nano-filtration (NF) using Ecotechnol's modified elements effectively removes the polyvalent ions responsible for the high lime and cyanide consumption as well as the prevalence of scale formation characteristic. Through removing up to 98% magnesium, 94% calcium and 99.7% sulfates from hypersaline water with total dissolved solids from 40,000mg/L to 250,000mg/L, Ecotechnol and subsequent independent tests have demonstrated up to an 87% reduction in lime consumption and through the ability to increase the operating leach pH, reduce cyanide consumption by up to 28%.

1. MAusIMM, Business Development Manager, Ecotechnol Pty Ltd, 38 Hobbs Tce, Dalkeith WA 6009, btapley@ecotechnol.com
2. MAusIMM, Director, Ecotechnol Pty Ltd, 38 Hobbs Tce, Dalkeith WA 6009, nstoitis@ecotechnol.com
3. Director, MDS membrane development specialist 921A S. Andreasen Dr, Escondido, CA 92029, l.lien@cox.net

INTRODUCTION

Membrane separation is commonly thought to only include reverse osmosis (RO) in Australia with seawater desalination projects or containerised plants providing potable water to remote mine sites (Levy, 2006). Recently though, a new style of filtration, namely Ecotechnol's modified nanofiltration (NF) has emerged as a viable treatment alternative thus gaining more acceptance in the mining industry to remove and/or concentrate valuable ions or waste products either for achieving environmental discharge standards or to improve the efficiency of the beneficiation process.

Unlike reverse osmosis, nanofiltration operates at much lower pressure and higher flux rates due to a more open weave resulting in lower operating costs and a smaller footprint (Dunne 2010). NF though is still much less prevalent in Australia than RO despite payback of capital of less than one year (Bernard, 1999). Essentially though, a nanofiltration installation is just modified reverse osmosis with the membrane being the only main difference.

Of particular importance to the Australian gold industry is the ability to effectively and economically pre-treat the hypersaline water prevalent in the Goldfields region of Western Australia through a combination of ultrafiltration and modified nanofiltration. As will be detailed in this paper, nanofiltration effectively separates and concentrates into a small stream the ions responsible for the high lime and cyanide consumption and also the occurrence of scale at gold projects operating with hypersaline water. This is possible due to nanofiltration's ability to separate and concentrate multivalent ions such as magnesium from monovalent ions such as sodium and chloride.

This paper outlines the findings of a test work program undertaken by Ecotechnol Pty Ltd on borefields water from a number of sites across Western Australia. Two projects will be discussed with lime demand and bottle roll tests undertaken to ascertain the benefit to lime and cyanide consumption.

This paper also provides an overview of membrane development, the operating principles behind nanofiltration and current global reverse osmosis and nanofiltration applications in order to highlight other potential opportunities within the Australian industry.

Nanofiltration plants has been successfully operated ranging from a relatively small modular and portable 20m³/hr modified nanofiltration plant at a refinery in the USA, to a 3,500m³/hr reverse osmosis plant in Peru which has operated since 2003.

MEMBRANE OVERVIEW

Membrane technology has been successfully applied to a number of industries over the past forty years. With recent developments in polymer chemistry, spiral wound element design and construction, improvements in pre-treatment, namely ultrafiltration, and an improved fouling mitigation techniques, nanofiltration has been used in a number of mining and oil/gas installations worldwide. Early applications of nanofiltration were in the pharmaceutical industry, chromium recovery in the leather industry and the treatment of textile effluents for reuse (Nel, 2103).

As illustrated in Figure 1 nanofiltration falls in the region between reverse osmosis and ultrafiltration. Wale (1989) indicates that ultrafiltration utilises sieving action at a molecular level to facilitate separation whilst reverse osmosis relies on the physico-chemical properties of the membrane transmitting water but actively rejecting salts. Operating pressures increase with a decrease in the weave aperture ranging from 0.2-1 bar for microfiltration to 80 bar for reverse osmosis.

Nanofiltration thus provides the opportunity to undertake several novel separation techniques which were once only possible through traditional techniques such as chemical precipitation, flocculation and coagulation, biological techniques or ion exchange (Levy, 2006; Dunne, 2010).

Table 1 outlines the opportunity for a purely physical separation technique through the use of typical nanofiltration by providing the rejection characteristics of a number of species (Van Der Merwe, 1998).

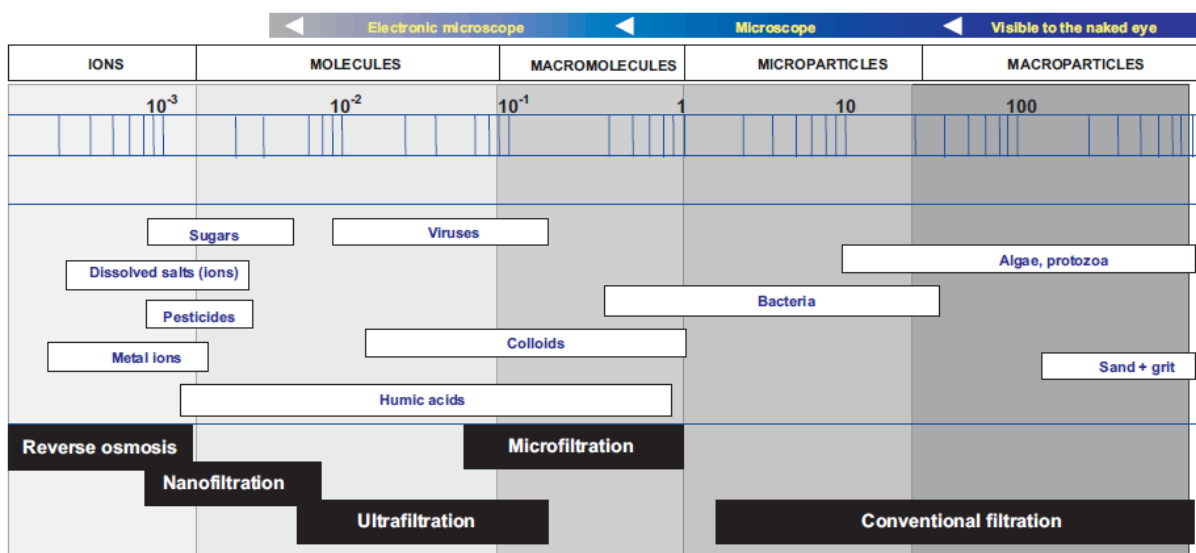


Figure 1: Filtration overview

Table 1: Comparative Rejection Values (Van der Merwe, 1998)

Species	Reverse Osmosis (%)	Nanofiltration (%)	Ultrafiltration (%)
NaCl	99	0-50	0
Na ₂ SO ₄	99	99	0
CaCl ₂	99	0-60	0
MgSO ₄	>99	>99	0
H ₂ SO ₄	98	0-10	0
HCl	90	0	0
Fructose	>99	>99	0
Sucrose	>99	>99	0
Humic Acid	>99	>99	30
Virus	>99.99	>99.99	>99.99
Protein	>99.99	>99.99	>99.99
Bacteria	>99.99	>99.99	>99.99

During operation, the feed solution is pumped under pressure along the membrane and via cross-flow filtration produces two streams; the permeate being the fluid which has passed through the membrane and the concentrate being the fluid which contains most of the dissolved and all of the suspended solids. Figure 2 illustrates this process.

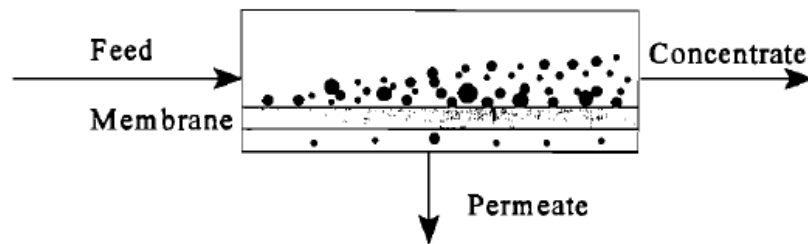


Figure 2: Cross flow Filtration Schematic

Lien (2008) and Van der Maerwe (1998) outlines the following key mechanisms of nanofiltration separation:

- Rejection by size

The membrane element has a molecular weight cut off (MWCO) between 150 and 300 Dalton meaning large uncharged, dissolved materials and some positively charged ions such as heavy metals can be rejected.

- Rejection due to the ionic charge

Multivalent anions such as sulfate (SO_4^{2-}) and carbonate (CO_3^{2-}) are strongly rejected by the negatively charged membrane element due to electric repulsion (95-99%). Conversely, the rejection of monovalent ions such as chloride (Cl^-) is considerably lower being between 5 and 45%. To ensure electroneutrality in the concentrate stream, cations associated with multivalent anions such as magnesium (Mg^{2+}) with sulfate are also strongly rejected. However cations associated with monovalent anions readily pass through the membrane such as sodium (Na^+) and chloride (Cl^-).

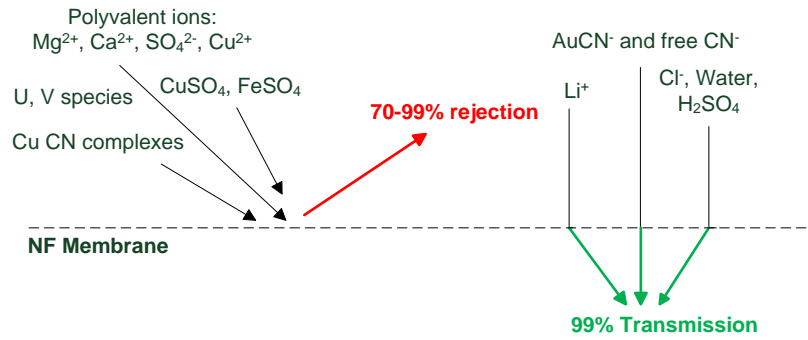


Figure 3: Rejection of Ions using Ecotechnol's Modified Nanofiltration Membranes

As illustrated in Table 1 and Figure 3, sulfuric acid (H_2SO_4) readily passes through the NF membrane due to the dissociation of the sulfate ion into the monovalent bisulfate ion as presented in Equation 1.



This phenomena leads to the possibility of separating sulfuric acid from metals associated with the sulfate ion.

One of the main parameters for the performance of nanofiltration is operating pressure. Pressure is dictated mainly by the required operating flux (Feini, 2008), the total dissolved solids (TDS) of the solution with a higher TDS requiring a higher pressure to maintain the same flux rate and also the temperature of the solution with a higher temperature leading to a higher flux rate. Pressure though is limited by membrane compression at high pressures leading to a decrease in membrane porosity and pore size.

To illustrate the impact of pressure on rejection, Feini (2008) conducted an experiment assessing the rejection of magnesium, sodium, calcium, chloride and sulfate. Reverse osmosis and nanofiltration membranes were tested with rejection levels presented in Figure 4. Reverse osmosis gave the highest rejection for all anions however nanofiltration resulted in the lowest rejection for chloride but highest for sulfate. Feini (2008) notes that nanofiltration provided good selectivity between monovalent and divalent ions with a low rejection of sodium chloride but high rejection of sulfate, magnesium and calcium. Due to these characteristics, Feini (2008) goes on to state that nanofiltration is more distinguished than reverse osmosis membranes in high hardness water treatment.

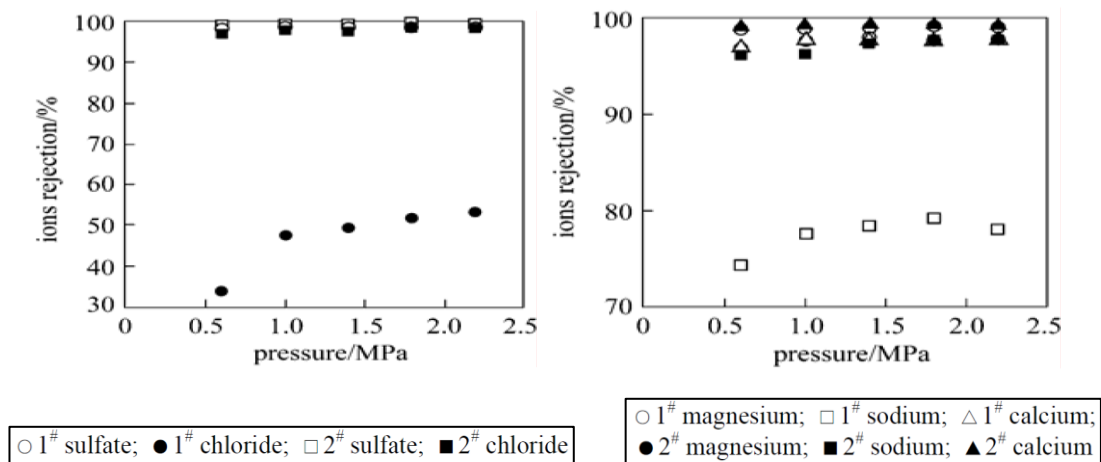


Figure 4: Rejection of Anions and Cations- NF (1[#]) Vs RO (2[#]) against Operating Pressure

Nel (2013) conducted a similar experiment on a synthetic nickel solution to demonstrate the impact of pressure. Nickel having a positive valence of two had consistently high rejection levels owing to the charge separation mechanism mentioned above and also by Ahn (1999). Acid (H^+) and sodium (Na^+) had the lowest levels of rejection however increased with pressure due steric and charge effects. Based on these findings, Nel (2013) states that it is possible to recover acid from a spent nickel electrolyte solution using nanofiltration principles.

With improvements to the design and manufacture of the polymeric Ecotechnol modified nanofiltration membranes, this separation technology can be applied at temperatures up to $140^{\circ}C$, pH ranges from 0 – 14, viscosities up to 300 centipoise and high soluble and suspended solids (Lien, 2009).

In order to minimise footprint and maximise flux rate, the spiral wound element has a very high surface area ($20\text{ cm} \times 100\text{ cm} = 30\text{ m}^2$) and able to operate at pressures up to 200 Bar. Additionally it has been found that the choice of spacer weave thickness is critical to provide sufficient turbulence on the membrane surface to minimise fouling (Bernard, 1999). The construction of a nanofiltration membrane is presented in Figure 5 and Figure 6. The expected membrane element life is between 4 and 5 years.

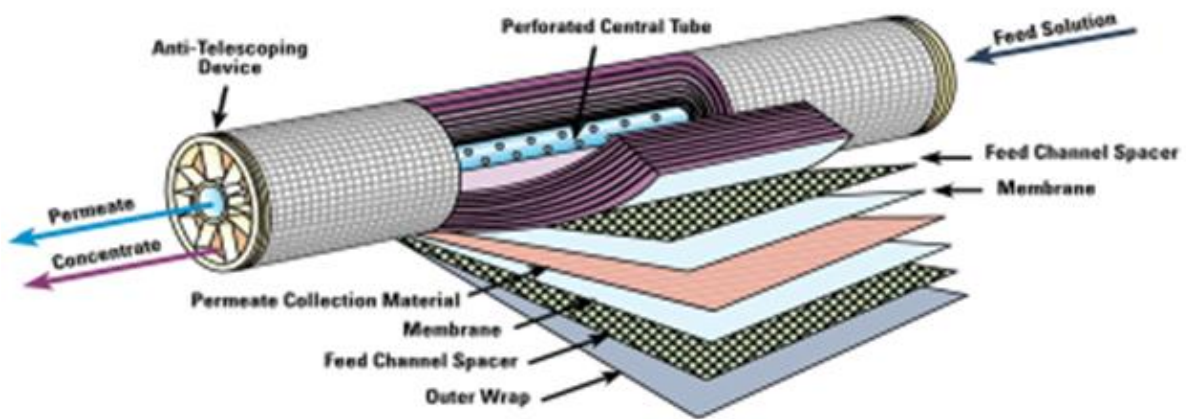


Figure 5: Spiral Wound Element (Bernard, 1999)



Figure 6: Customised Ecotechnol Spiral Wound Element with Unique Feed Spacers and Stainless Steel Centre Tube

Many previous installations of reverse osmosis plants have suffered due to a lack of adequate pre-treatment. The use of cartridge or sand filters in high TDS environments have led to the fouling of the downstream membranes, higher power and maintenance cost and ultimately their early replacement. The advent of ultrafiltration and the appropriate dosage and type of antiscalant has led to an

improvement in the operation and cost of reverse osmosis plants through the removal of large particles from the water such as algae and clays. For this reason, ultrafiltration is usually placed prior to nanofiltration together with an appropriate antiscalent to protect the membrane elements from fouling (Levy, 2006).

GOLD PROCESSING IN HYPERSALINE WATERS

Hypersaline water of the Western Australia Goldfields

The processing of gold ores within the Goldfields Kalgoorlie region of Western Australia is commonly performed in waters classified as hypersaline with magnesium and TDS levels up to 10,000 mg/L and 280,000 mg/L respectively (Landgate 2015, Kern, 1995). According to Landgate (2015), hypersaline water is termed as any water more saline than seawater which is approximately 35,000 TDS. Although the use of seawater is not uncommon in mineral processing; in some cases leading to improved metallurgical performance (Dunne, 2010; Aral, 2010), the elevated magnesium (Mg^{2+}), calcium (Ca^{2+}) and sulfate (SO_4^{2-}) levels in hypersaline water lead to particularly high reagent consumptions and reduced metallurgical performance in the cyanide leaching of gold (Dunne, 2010; Stoitis, 2011; La Brooy, 1994; Aral, 2010; Chryssoulis and McMullen, 2005).

Ecotechnol has compiled an extensive list of Western Australian gold operation water analyses (Table 2) indicating the water quality varies quite significantly in regards to TDS, magnesium, calcium and sulfate concentrations.

The Southern Cross, Kalgoorlie and Norseman localities have the most hypersaline water whereas operations in the Murchinson and Wiluna have water classified as marginal to brackish (Landgate 2015). Similar Government studies on the variation of TDS and salts across Australia have also indicated the prevalence of hypersaline water in the Goldfields region (Kern, 1995). Figure 7 clearly illustrates the dominance of hypersaline water in the goldfields compared to the rest of the state.

Table 2: Borefields Water Analysis of Western Australian Gold Operations

Site- Locality	pH	TDS mg/L	Cl mg/L	Ca mg/L	Mg mg/L	Na mg/L	Sulphate mg/L
Southern Cross	6.76	249,500	144,630	820	10,020	79,900	13,244
Norseman	7.3	280,000	130,000	780	9,100	82,000	18,000
Kalgoorlie- East	6.54	260,000	150,000	665	8,150	76,000	17,500
Kalgoorlie- East	6.4	124,000	143,400	584	7,940	78,100	-
Kalgoorlie- North East		88,000	69,000	101	7,210	36,300	-
Laverton- South	7.07	100,000	82,000	970	5,300	47,000	11,000
Laverton- South	7.4	230,000	130,000	840	5,000	78,000	15,000
Southern Cross	6.76	109,450	62,950	307	4,327	33,200	6,513
Kalgoorlie- South	5.9	104,600	58,700		4,300	34,800	6,100
Laverton- South		98,000	70,000	1,300	4,100	41,000	8,200
Wiluna- South East	7.6	42,528	65,518	739	4,073	35,610	12,563
Southern Cross	6.65	86,800	49,500	314	4,000	24,900	5,700
Menzies- West	7.72			3,000	4,000		800
Coolgardie	7.7	90,200	47,000	1,500	3,800	28,000	-
Southern Cross	6.9	83,200	48,000	254	3,800	24,600	5,400
Southern Cross	7.1	78,900	45,500	244	3,650	23,400	5,500
Southern Cross	6.6	73,100	41,000	239	3,550	20,900	4,950
Southern Cross	7.1	77,700	44,500	244	3,450	23,400	5,550
Kambalda	3.8	-	49,000	500	3,272		2,060
Kalgoorlie- North East	6.6	91,000	49,000	320	3,100	28,000	7,300
Southern Cross	4.5	70,000	37,000	300	3,000	19,000	4,900
Meekatharra- North West		39,300	23,000	510	2,600	14,100	
Coolgardie	-	66,300	36,750	635	2,570	20,300	1,900
Kalgoorlie- South East	5.85	57,950	11,800	475	2,490	25,000	4,365
Southern Cross	6.97	63,726	35,900	681	2,264	20,100	4,210
Coolgardie- West		47,000	30,000	210	2,100	17,000	
Southern Cross	7.2	37,000	19,000	610	1,900	7,100	2,400
Kalgoorlie- East	7.7	45,100	24,645		1,795	13,450	
Southern Cross	3.9	72,000	43,000	2,200	1,200	20,000	3,000
Southern Cross	5.6	51,000	28,000	1,300	1,200	15,000	1,500
Murchinson	8.3	24,000	-	150	850	7,000	2,900
Murchinson	8.5	6,200	2,900	98	240	1,900	730
Wiluna	7.7	4,525	1,600	143	215	945	985
Murchinson	7.8	4,500	1,800	285	138	1,300	
Meekatharra- North West		2,380	970	16	120	680	-
Wiluna- East	7.8	1,200	300	80	62	190	252
Wiluna- South West	7.69	740	170	102	40	59	120

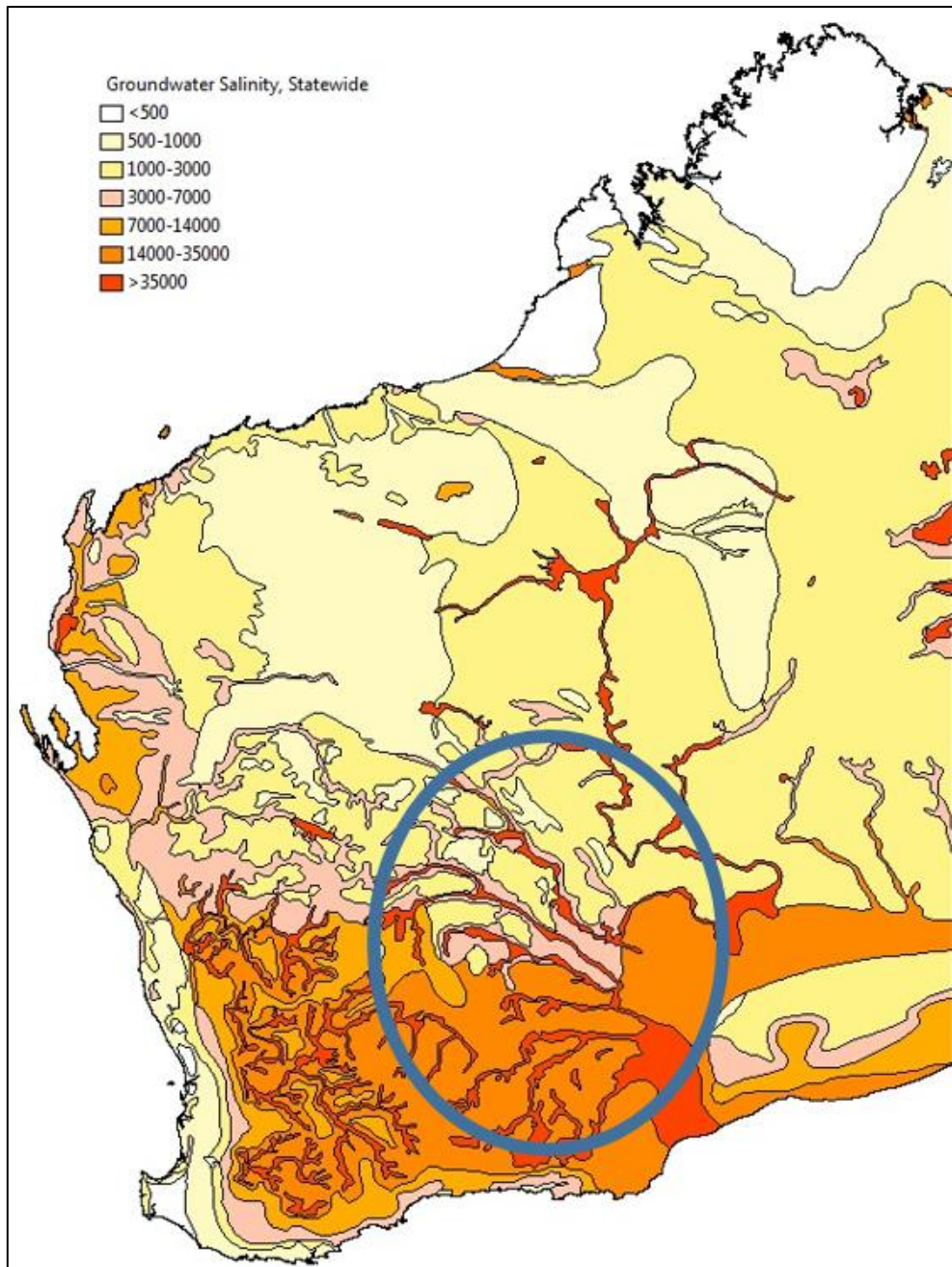


Figure 7: Groundwater Salinity Map- Goldfields region circled (Department of Water, 2015)

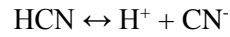
The hypersaline water of the goldfields occurs mainly in palaeochannels, bed rock and adjacent alluvial flats and playa lakes and also in greenstone belts; geological formations all associated with gold mineralisation in the Goldfields. Lesser brine waters (3,000-30,000mg/L TDS) can be found in Archaean bedrock and at the head waters of palaeochannels. As described by Kern (1995) the salts in the groundwater derive from marine aerosols rather than from the weathering of the bedrock. These salts may have been accumulating for hundreds of thousands of years.

For optimal operations, it is imperative for gold operations to source the lowest salinity borefields water within an economic distance of the processing plant. Operations which do not have this ability are forced to operate with these hypersaline waters.

Effect of Hypersaline Water on the Cyanide Leaching of Gold

Impact on Operating Leach pH

Cyanide leaching of gold is usually carried out at approximately pH 10.0-10.5 to ensure the highest possible economical concentration of free cyanide ions (CN^-). Marsden (2006) reported the pK_a of cyanide is 9.31 at 25°C in fresh water meaning that at this pH, 50% of the cyanide is present as CN^- and 50% as HCN (Figure 8). As noted by Aral (2010) and Rumball (1997), aqueous HCN is significantly less effective in leaching gold.



(Equation 2)

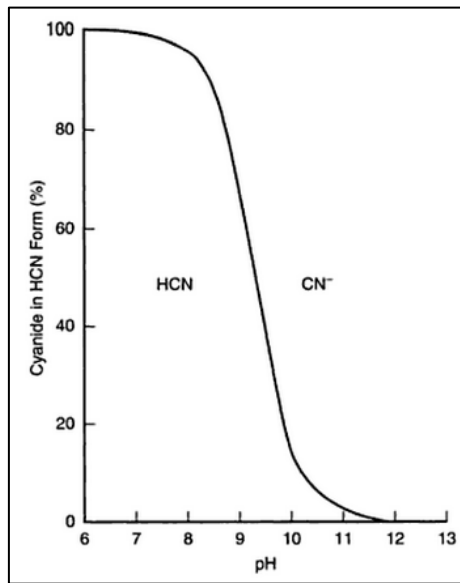
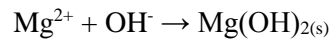


Figure 8: Speciation of cyanide and hydrogen cyanide in aqueous solution as a function of pH (Marsden, 2006)

The presence of hypersaline water containing elevated concentrations of magnesium in the resulting leach slurry buffers the pH with the addition of lime to provide hydroxide ions (OH^-) as per Equation 3 (Stoitis, 2014; Aral, 2010). This leads to lower operating pH, typically 9.0-9.2 (La Brooy, 1994; Perry, 1999) and sometimes as low as pH 8.5-8.7 mainly depending on the quality of the available site water.



(Equation 3)

Figure 9 shows that the pH will remain buffered until all the magnesium has precipitated out of solution. The lower than desired operating pH leads to a greater proportion of the cyanide to be present as HCN which contributes to cyanide consumption.

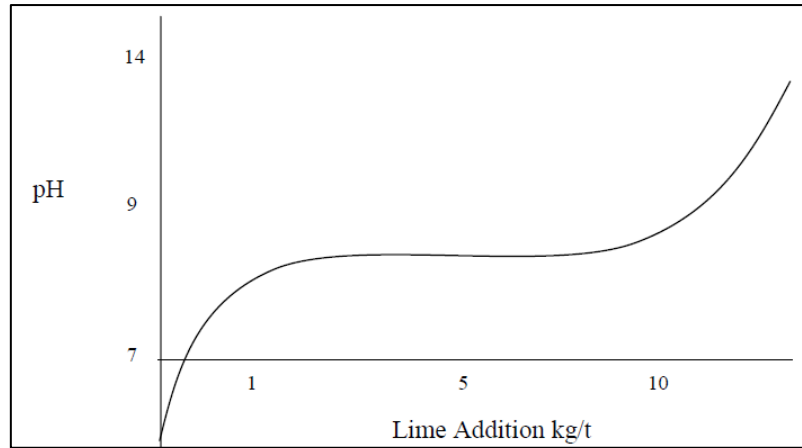


Figure 9: Example of a Lime Buffering Graph (Stoitis, 2014)

Negus (2013) highlighted the impact of the quality of water on the Sunrise Dam gold operation located 45km south east of Laverton. The quality of raw water varies from approximately 30k TDS to 200k TDS leading to an approximate 300% increase in lime consumption to maintain a desired leach pH and 20% increase in cyanide consumption. This is illustrated in Figure 10.

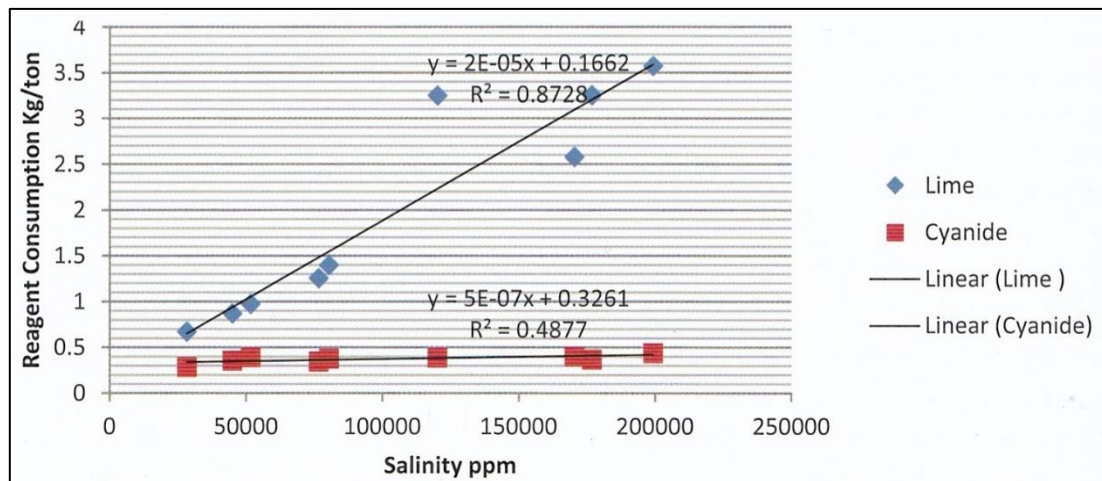


Figure 10: Impact of Salinity on Lime and Cyanide Consumption (Negus, 2013)

Salinity also has a varied impact on the pKa of cyanide as noted by La Brooy (1994), Verhoeven (1990) and as illustrated in Figure 11. This indicates that the cyanide pKa is not constant and is dependent upon the salinity of the water, amongst others. For waters containing 30-60k TDS, the pKa is lower meaning that a greater proportion of the cyanide is present as CN^- . However at elevated TDS, the pKa increases above that of fresh water resulting in higher cyanide consumptions.

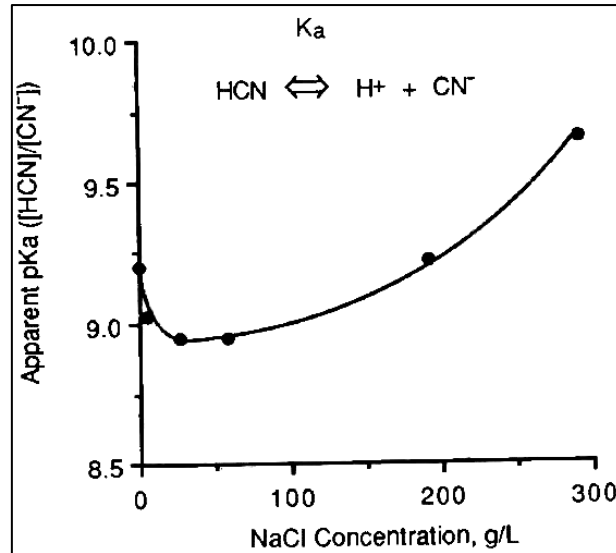


Figure 11: Impact of Salinity of Apparent pKa of HCN/CN⁻ (La Brooy, 1994)

Impact on Dissolved Oxygen Levels

An increase in process water salinity results in a decrease in the solubility of oxygen leading to lower dissolved oxygen levels in the pulp (Dunne, 2010; La Brooy, 1994 and La Brooy, 1991). As expressed by the Elsner equation (Equation 4), both cyanide and oxygen are essential to the leaching of gold. The impact of TDS is presented in Figure 12 whereby the oxygen solubility and gold leaching rate decrease with an increase in TDS. To overcome the negative impact of a hypersaline water, gold operations inject oxygen into the leaching tanks in the form of liquid oxygen (LOX) or oxygen gas derived from an onsite oxygen plant (eg PSA plant).

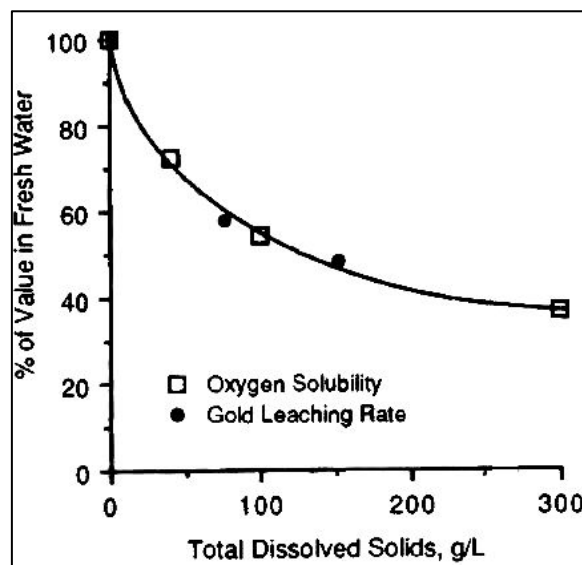


Figure 12: Effect of Salinity on Oxygen Solubility and Rate of Gold Leaching at 27°C (La Brooy, 1991)

Impact on Scale Formation

Hypersaline water is characterised by high concentrations of calcium, carbonates and sulfates. The presence of these ions lead to the formation of gypsum (CaSO_4) and calcium carbonate (CaCO_3) in the pH range of gold cyanide leaching (Johnson, 2003; Kern, 1995). These scales are prolific in operations which utilise hypersaline borefields water leading to blocked screens, reduced efficiency of centrifugal concentrators and blocked water pipelines (Figure 13).



Figure 13: Scale Formation in a Pipeline

To reduce the formation of these scales, gold operations dose antiscalant in the process water, water used in heat exchangers and in the gravity circuit of the plant.

Impact on Carbon

Carbon Fouling and Regeneration

Gypsum and calcium carbonate absorb onto activated carbon reducing the ability of the carbon to adsorb gold, ie. reduce carbon activity.

The presence of calcium and magnesium salts during the thermal regeneration of carbon post elution catalyses carbon decomposition with weight losses up to 20% being observed (La Brooy, 1994). Additionally, due to the over activation of the carbon during regeneration, the resulting carbon has a much lower resistance to attrition thus resulting in greater carbon losses once reintroduced back to the adsorption circuit.

Therefore to minimise the impact of the presence of magnesium, calcium, sulfates and carbonates, loaded carbon is washed in hydrochloric acid prior to elution.

CURRENT MEMBRANE APPLICATIONS

Reverse Osmosis technology in Industrial Applications

Further to generating potable water from seawater and brackish water, several operations have utilised reverse osmosis and nanofiltration in other applications:

- ***Queensland Nickel (QNI) Refinery*** in Townsville installed a microfiltration and reverse osmosis plant in 2002 to remove concentrated salts from the process water increasing the amount of water that was able to be re-used;
- ***ERA Ranger Uranium Mine*** in Northern Territory installed a reverse osmosis plant in 2006 to treat contaminated water prior to discharge into the environment (Levy, 2006).
- ***Yanacocha Gold Mine*** in Peru installed a modified reverse osmosis plant treating 3,500 m³/hr of Barren Leach Solution in 2003. Table 3 outlines the performance of the installation allowing

safe discharge after chlorine treatment into the environment. The Client noted that chlorine consumption reduced by 75% and the overall operating cost compared to the previous precipitation plant reduced by 70%.

Table 3: Yanacocha Reverse Osmosis Plant Assays

Ion	Feed mg/L	Permeate mg/L	Concentrate mg/L	Discharge Limit mg/L	Limit Achieved
pH	10.1	8	9.7	6.0-9.0	Y
CN WAD	46.7	<0.05	117.5	0.2	Y
Arsenic	0.4	<0.01	1.5	.5-1	Y
Mercury	0.0025	<0.0005	0.0076	0.002	Y
Nitrite	5.19	0.09	17.11	-	n/a
Nitrate	27.5	0.64	89.8	-	n/a
Copper	3.1	0.1	11.6	0.3	Y
Zinc	17.2	0.3	65.1	1	Y

Nanofiltration Globally

Concentration of Diluted Copper Leach Solution and Acid Purification

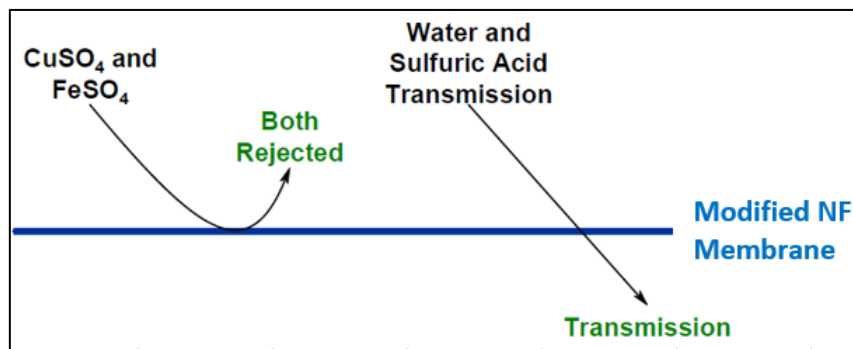


Figure 14: Copper and Iron Rejection from Acidic Medium with Specialised Acid NF Membranes

- Copper SX/EW Plant- Mexico

A nanofiltration system was installed in 1996 at a copper Solvent Extraction/Electrowinning (SX/EW) operation in Mexico (Bernard, 1999; Lien, 2009). The primary aim was to treat approximately 15 GL of Acid Mine Drainage (AMD) impacted abandoned pit water at a copper grade of between 0.6-0.8 g Cu/L. At the time of installation, this solution equalled the SX/EW raffinate solution grade thus was not deemed as suitable feed. In order to obtain safe access to an adjacent pit the contaminated water needed to be relocated.

Both lime precipitation and a nanofiltration system were evaluated. Unlike lime precipitation where no saleable products were generated, the installed 900m³/hr nanofiltration plant generated a concentrated copper stream which was sent directly to EW. A secondary benefit was the production of a clean permeate stream free of heavy metals with copper and iron grades of less than 10 ppm and 50 ppm respectively which was reused in the process. At the time of publishing, approximately half the pit water had been treated over an 18 month period recovering more than USD 10M of copper ensuring a payback of less than 12 months. The alternative method of using lime was estimated to have cost USD 7M just in operating cost. A photo of the membrane plant is presented in Figure 15.



Figure 15: Nanofiltration Plant in Mexico

- *Copper Dump Leach- USA*

A second example was a copper dump leach in USA whereby a nanofiltration pilot plant was installed to increase the efficiency and extraction capacity of the SX/EW plant (Eriksson, 1996). The pilot plant successfully doubled the concentration of the pregnant leach solution (PLS) in the concentrate stream at a recovery of 50%. This could also apply prior to a Sulphidisation, Acidification, Recycling, Thickening (SART) plant whereby the incoming feed is further concentrated into a small stream thus reducing the footprint of the SART plant and operating costs, namely through reduced acid and caustic consumption (Picaro, 2011).

- *Freeport-McMoran Copper*

A third case study is a system consisting of both a reverse osmosis and nanofiltration installed and still operating at Freeport-McMoran copper operation since 1993. In this installation, a copper rinse stream containing low concentrations of copper (0.1 g/L) and sulphuric acid (1%) are upgraded to approximately 30 g/L copper sulfate for sale and 100 g/L sulphuric acid for re-use. This flowsheet is presented in Figure 16.

The recovery of rinse water and copper paid for the initial capital within 9 months of commissioning.

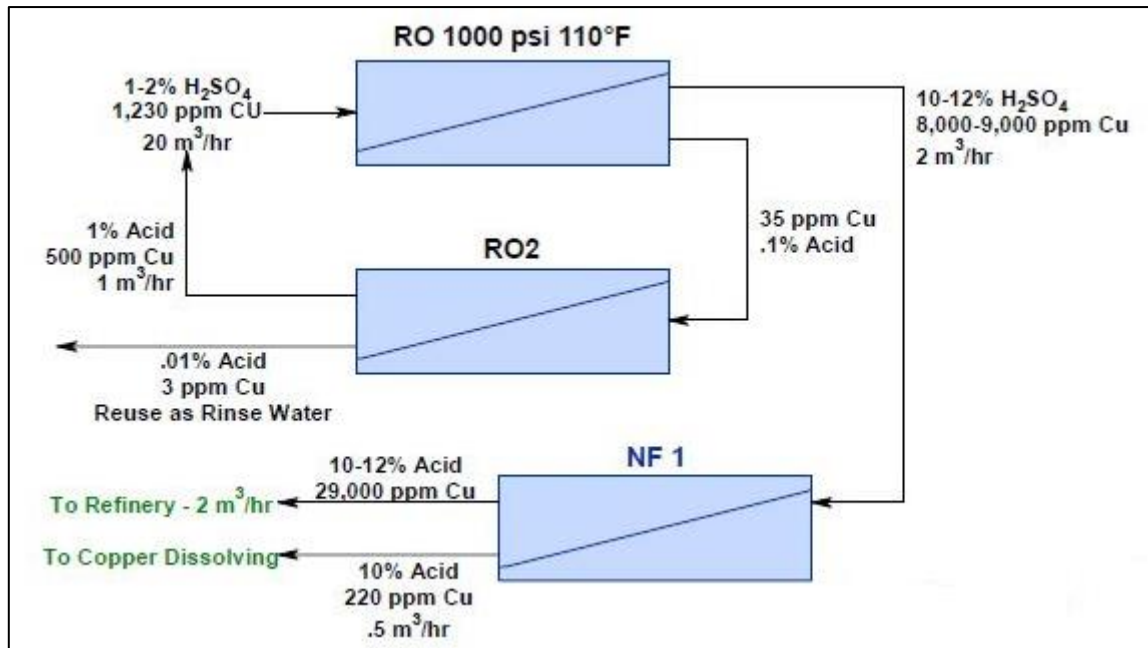


Figure 16: Freeport-McMoran Flowsheet

Copper-Gold Separation

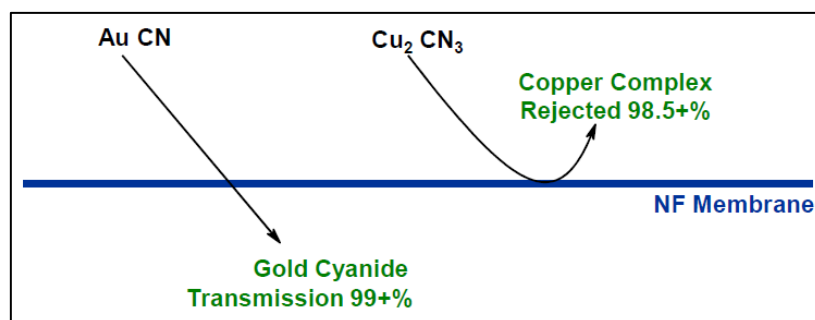


Figure 17: Copper-Gold Separation with Modified Nanofiltration Membranes

Test work undertaken at Ecotechnol has demonstrated the ability of nanofiltration to concentrate copper and complexed cyanide species whilst transmitting AuCN^- and free cyanide to permeate as illustrated in Figure 17. Copper is concentrated into approximately 10% of the initial feed volume at a recovery of 99% resulting in a permeate containing ~99% of the gold and less than 40 ppm copper.

When a nanofiltration plant is installed on a PLS stream high in copper from a gold heap leach, the nanofiltration plant will concentrate the copper allowing for the potential installation of a SART plant resulting in the generation of saleable copper and the recycling of cyanide back to the circuit (Jay, 2001). At the very least, the clean permeate virtually devoid of copper will report to the carbon adsorption circuit ensuring improved adsorption rates, lower cyanide requirements and improved bullion quality.

This flowsheet is illustrated in Figure 18 additionally indicating the possible inclusion of a RO plant on the Barren Leach Solution (BLS) on a water positive site to recover cyanide before disposing of the clean permeate.

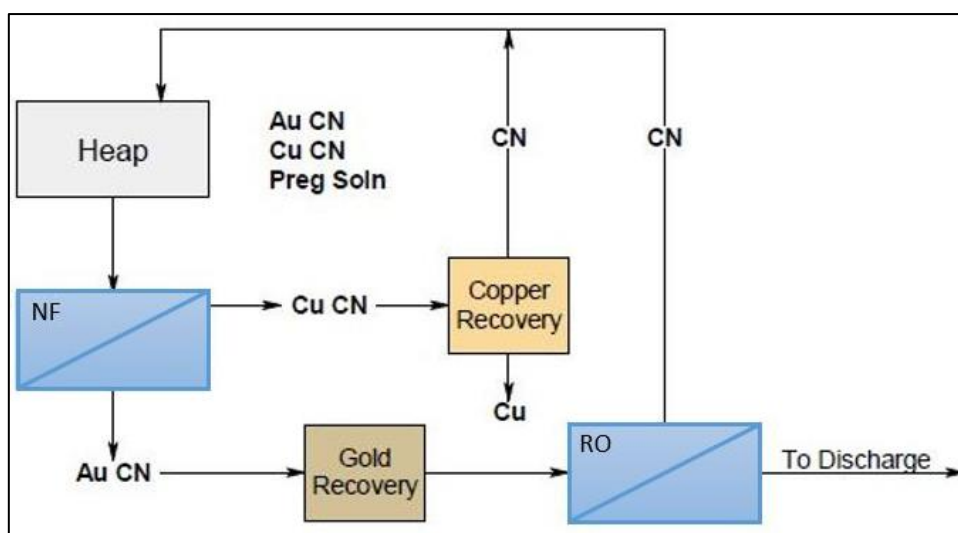


Figure 18: Copper-Gold Fractionation Process

With this separation possible, it is feasible to promote the installation of a nanofiltration plant on the tailings return water of a gold operation high in soluble copper. The plant will remove the recycled copper and other complexed cyanide species reducing the return of WAD cyanide species and cyanide consuming copper in the solution.

Uranium Applications

Nanofiltration membranes are effective in concentrating uranium ions whether generated from alkaline or acidic leach processes. Test work undertaken by Ecotechnol has demonstrated that Uranyl Sulfate (UO_2SO_4) and Uranyl Carbonate (UO_2CO_3) are strongly rejected by a modified nanofiltration membrane as demonstrated by the results presented in Table 4 and pictured in Figure 19. An example of how a membrane system could be implemented in a uranium heap leach is presented in Figure 20 indicating that 1-4 g/L Uranyl Sulfate is concentrated into approximately 10% of the original flow resulting in concentrations of 10-40 g/L Uranyl Sulfate. The purified acid containing less than 0.001 g/L Uranyl Sulfate is returned to the leach for reuse.

Table 4: Alkaline Leach Assays

Analyte	Feed mg/l	Permeate mg/l	Concentrate mg/l	% Rejection
Calcium, Ca	11.6	1.8	32.7	92%
Magnesium, Mg	113	9.6	580	97%
Vanadium, V*	302	64.2	1,539	93%
Sulfate, SO_4^{2-}	4,600	170	17,000	98%
Carbonate, CO_3^{2-}	898	311	2,413	81%
Bicarbonate, HCO_3^-	19,098	21,041	16,236	-19%

* Vanadium was used as an acceptable substitute for Uranium in this test work

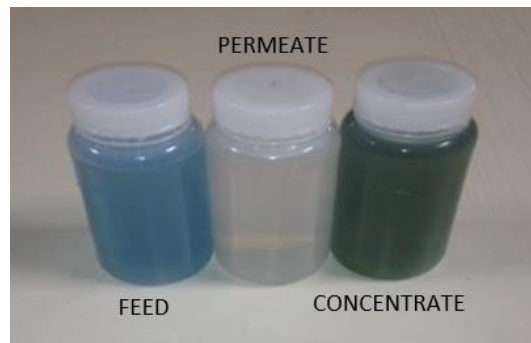


Figure 19: Alkaline Leach- Feed, Permeate and Concentrate

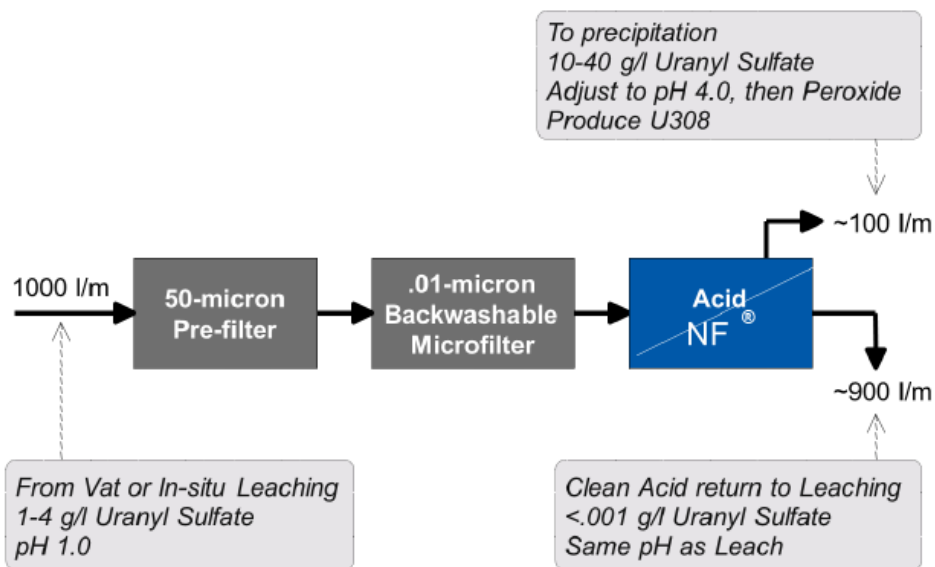


Figure 20: Uranium Acid Vat or Heap Leach Application of Nanofiltration

Environmental Applications

The ability of nanofiltration to concentrate heavy metals such as cadmium (Feini, 2008) and separate multivalent ions such as copper and carbonates from monovalent ions has enabled its implementation across a range of environmental projects in order to achieve discharge limits.

- Acid Mine Drainage- Copper Operation

Ecotechnol undertook a test work program in 2011 with the aim to produce permeate capable of being discharged into the environment at a client's operation (Ecotechnol, 2011). Table 5 presents the results of the program indicating reductions of greater than 95% ensuring the environmental discharge limit on all but one (copper) was met. Ecotechnol was confident in achieving less than 2.5 mg/L copper in the final permeate in an operating environment.

Table 5: Acid Mine Drainage Test work Results

Parameter	Units	Limit Type	Discharge Limit	Feed Solution	1st Stage Permeate	2nd Stage Permeate	% Rejection	Limit Achieved
% recovery to permeate					75%	85%		
Copper	mg/L	Maximum	2.5	190	17	2.7	98.6%	N
Arsenic	mg/L	Maximum	0.25	0.03	<0.03	0.001	~95%	Y
Cobalt	mg/L	Maximum	5	18	1.5	0.22	98.8%	Y
Lead	mg/L	Maximum	0.05	<0.02	<0.02	<0.001	~95%	Y
Zinc	mg/L	Maximum	10	3.8	0.35	0.07	98.2%	Y
Sulfate	mg/L	Maximum	3500	4700	390	51	98.9%	Y
pH	pH units	Range	6.5-8.5	2.9	3.7	4.8		n/a*
TDS	mg/L	Maximum	7000	7100	540	100	98.6%	Y

- Note- pH adjustment was not in the scope of the test work program

- *New Zealand Gold Mine Effluent Treatment Prior to Environmental Discharge*

Antimony, Arsenic and Selenium discharge concentrations of less than 5 ppb were commonly exceeded through the use of a conventional water treatment system consisting of chemical precipitation followed by clarification. This observation was supported by Kempton (2003) noting a lime precipitation system alone typically cannot meet regulatory standards. A nanofiltration membrane plant was installed on the clarifier overflow prior to environmental discharge which reduced the antimony, arsenic and selenium concentrations to less than 2 ppb, 0.01 ppb and 0.2 ppb respectively at a volumetric recovery to permeate of 85%.

- *Refinery Waste Water Reclamation Project*

A precipitation system was installed in 1985 to commence treating contaminated groundwater; a legacy of over one hundred years of operation. Due to high sustaining capital and operating costs, and common discharge limit exceedances in 1993 a nanofiltration system was installed prior to precipitation in order to reduce the volume to precipitation.

Following its installation, the volume to precipitation reduced by 80% significantly reducing chemical consumption and the amount of sludge produced and hence disposed of at great cost. The flowsheet is presented in Figure 21 with corresponding assays of the feed and permeate in Table 6 highlighting the reduction in targeted elements through nanofiltration alone.

With the introduction of a nanofiltration system, the operating cost of the water treatment plant reduced by 90% mainly due to a reduction in sludge disposal.

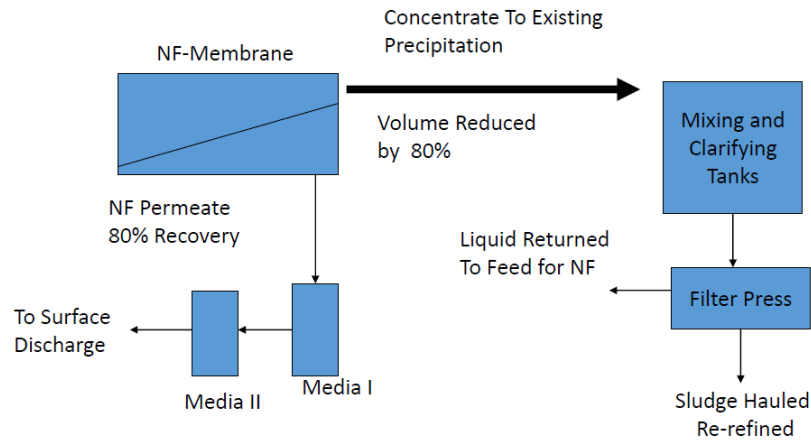


Figure 21: Installation of a Nanofiltration Plant in a Refinery Waste Water Plant

Table 6: Feed and Permeate Assays with the Installation of a Nanofiltration Plant

Element	Feed	Permeate	Rejection
As	10.1	0.081	99.2%
Cd	14.5	0.05	99.7%
Zn	33.5	0.01	~100.0%
Pb	3.07	0.05	98.4%
Cu	0.073	0.01	86.3%
Fe	0.983	0.1	89.8%
Mn	3.33	0.5	85.0%
Total Metals	67.9	0.583	99.1%

ECOTECHNOL TESTWORK

Overview

Using Ecotechnol's lab scale filtration unit capable of ultrafiltration, nanofiltration and reverse osmosis membranes, Ecotechnol developed an experimental protocol to assess the separation of multivalent ions, particularly calcium, magnesium and sulfate from hypersaline borefields water provided by numerous gold operations across Western Australia. Noting nanofiltration's performance in the case studies described, it was anticipated that at least 80% of magnesium, calcium and sulfate would be recovered into a small concentrate stream at low power requirements. The removal of magnesium as previously discussed should result in reduced lime buffering resulting in a lower lime consumption to achieve a given pH, or alternatively permit a higher operating pH benefiting cyanide consumption.

The lab scale unit is presented in Figure 22 and Figure 23. This allows up to 60 L of untreated water to be passed through a desired membrane with the following test work parameters measured:

- Operating pressure
- Permeate flow rate
- Feed temperature, and
- Final water recovery to permeate based on volumes.

The only manipulative variable for the test was the ability to increase pressure within the filter membrane to achieve an acceptable permeate flow. As witnessed in the test work, a water higher in TDS would require a greater pressure in order to achieve acceptable flow. An antiscalent based on the

feed assay results conducted prior to the test work was added in order to minimise precipitation of calcium sulfate thus increasing recovery to permeate and reduce operating pressure.

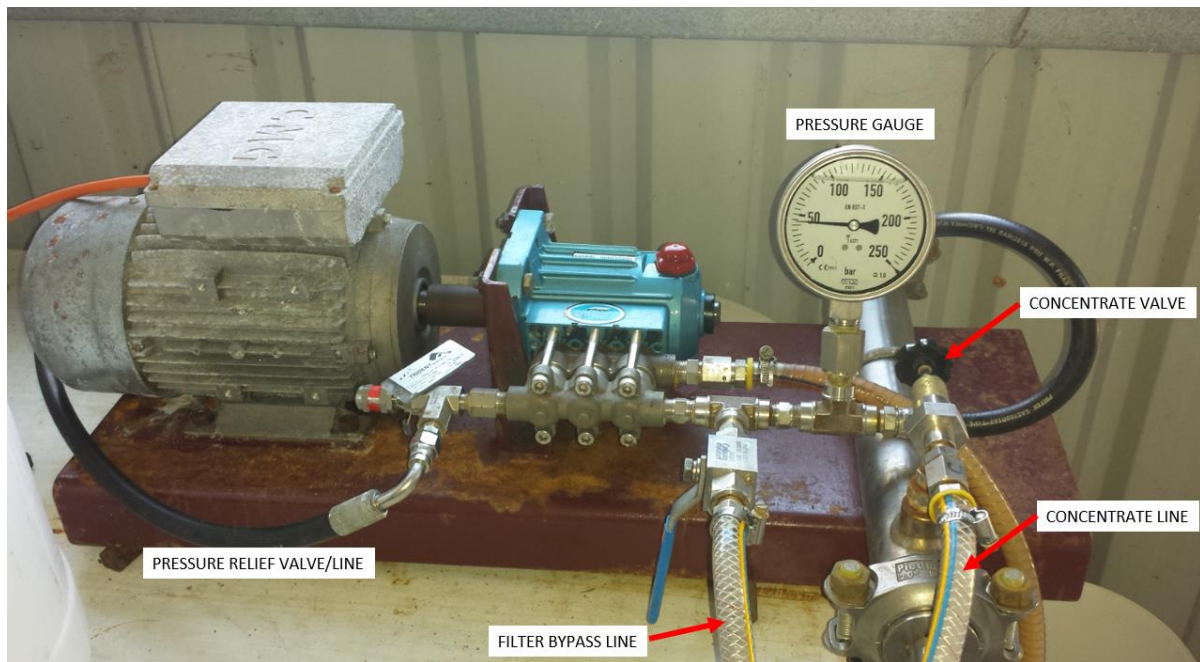


Figure 22: Ecotechnol Laboratory Scale Filtration Unit- Front View

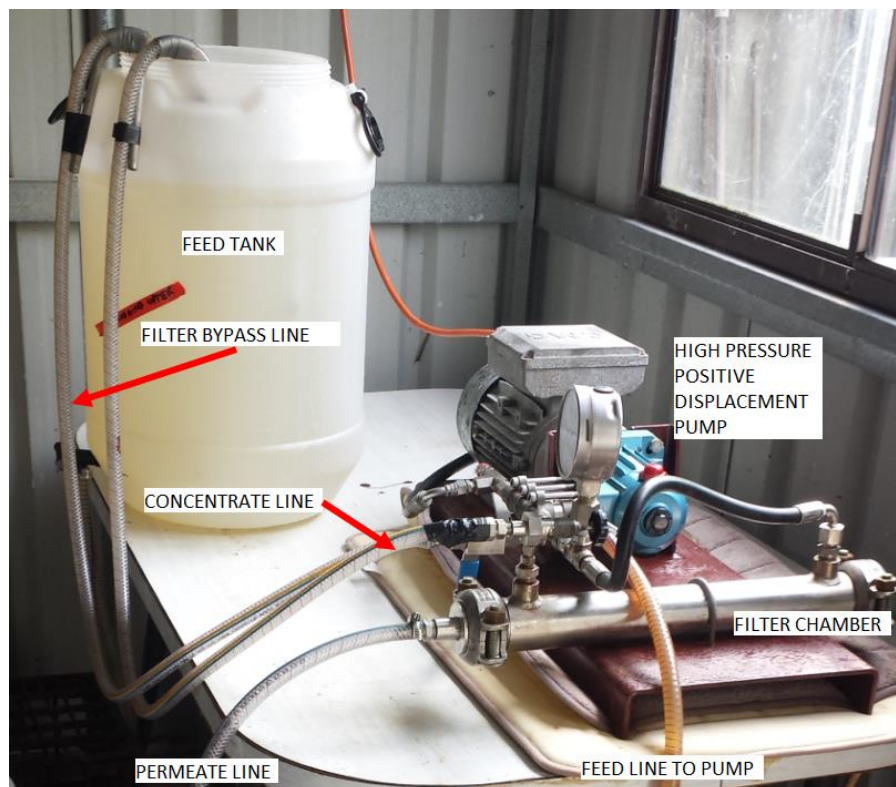


Figure 23: Ecotechnol Laboratory Scale Filtration Unit- Side View

The operation of the lab scale unit involves passing the feed solution through a high pressure positive displacement pump to the filter chamber. The flow through the filter membrane is controlled by the concentrate valve. The concentrate then recycles back to the feed tank.

All waters are firstly treated through an UF membrane (0.04 μ m) thus removing all particulates which commonly impact the efficiency of nanofiltration. An example of the effectiveness of an UF membrane is presented in Figure 24. The volumetric recovery to permeate for a UF test is approximately 98-99% of the initial feed volume with permeate flow rates up to 300 mL/min (65 mL/m²/min).



Figure 24: Borefields Water Sample and Treated UF Permeate

The remaining solution free of particulates is then passed through a NF membrane. The filter is operated similarly to the UF membrane, however as it has a tighter weave the operating pressures are higher (30-55 bar) and flow rates lower at 100 mL/min to 180 mL/min (21- 39 mL/m²/min). The concentrate is recycled back to the feed container and the test continues until either the desired water recovery is achieved or the permeate flow rate becomes unacceptably low. This is typically because the saturation point of calcium sulfate is reached, which can to a certain degree be mitigated with the use of an antiscalant. With an increasing concentration of these key ions in the feed solution period the flow rate reduces over the test period as the operating pressure remains constant.

Feed, permeate and in most tests concentrate samples are collected and submitted for assay at a NATA accredited laboratory.

Effectiveness of Nanofiltration in the Rejection of Key Ions

A total of nine tests of various scopes were conducted by Ecotechnol. For all tests, the supplied borefields water was passed through the Ecotechnol lab scale UF and NF unit with feed, permeate and concentrate assayed to determine the rejection of key ions (Table 7 and Table 8). The volumetric recovery to permeate and operating pressure for these tests are presented in Table 9.

As directed by the client, lime demand and cyanide bottle roll test work was conducted on borefields water from the Bullabulling Gold Project located west of Coolgardie and Saracen Carosue Dam located south of Laverton.

Table 7: Hypersaline Water Test work Assays

Site	Feed Concentration				Permeate				Concentrate			
	TDS	Mg	Ca	SO ₄ ²⁻	TDS	Mg	Ca	SO ₄ ²⁻	TDS	Mg	Ca	SO ₄ ²⁻
	mg/L (k)	mg/L	mg/L	mg/L	mg/L (k)	mg/L	mg/L	mg/L	mg/L (k)	mg/L	mg/L	mg/L
Bullabulling	49	2,400	300	4,300	39	87	17	130		9,100	930	15,000

Carosue Dam	100	5,300	970	11,000	100	710	320	180				
3	280	7,940	683	16,800	210	1,850	299	54	280	18,000	730	57,000
4	56	2,500	360	3,600	42	52	20	52				
5	39	67	2,400	2,300	25	<5	180	59				
6	98	4,100	1,300	8,200	90	330	230	79	99	9,200	1,100	16,000
7	250	8,600	740	11,000	220	1,400	270	450	270	16,000	530	28,000
8	230	5,000	840	15,000	210	1,200	450	1,200	260	7,600	570	25,000
9	72	1,200	2,200	3,000	57	180	320	160	91	3,200	3,800	2,700

Table 8: Hypersaline Water Test work Rejection of Key Ions

	Ion Rejection (%)		
Site	Mg	Ca	Sulfate
Bullabulling	96	94	97
Carosue Dam	87	67	98
3	81	65	99.7
4	98	94	99
5	93	93	97
6	92	82	99
7	84	64	96
8	76	46	92
9	85	85	95

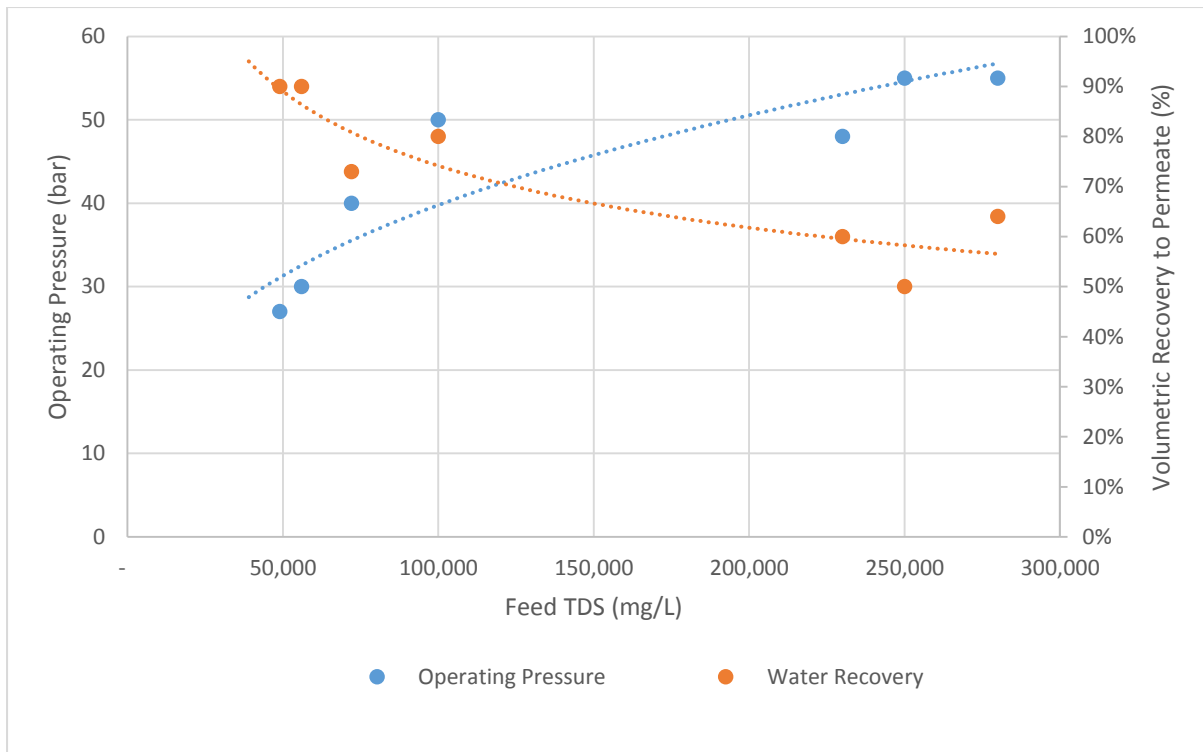
Table 9: Volumetric Recovery and Operating Pressure

Site	% Water to Permeate	Pressure (bar)
Bullabulling	90% **	27
Carosue Dam	80%	50
3	64%	55
4	90%	30
5*	50% **	38
6	50% **	48
7	50%	55
8	60%	48
9	73%	40

* Process Water tested ** No antiscalant added

These tests proved that nanofiltration is effective in removal of between 76% and 98% of magnesium and 92% and 99% of sulfate in the tested hypersaline waters. The volumetric recovery of water to permeate ranged from 50% to 90%, depending mainly on the TDS of the initial feed solution and the test work directive of the client to maximise either recovery or permeate quality.

As illustrated in Figure 25, an increase of the feed water TDS impacts adversely on the operating pressure and volumetric recovery to permeate. The volumetric recovery increases with the addition of antiscalant and using multi-stage filters for plant operation.



* Tests 5 and 6 were removed from chart due to being a process water and not having antiscalent added respectively

Figure 25: Impact of TDS on Operating Pressure and Volumetric Recovery of Water

Lime Demand and Bottle Roll Test work

Lime demand and cyanide bottle roll tests carried out on Bullabulling site water and Carosue Dam site water clearly illustrates the impact of removing key multivalent ions, namely magnesium from the borefields water (96% and 87% magnesium removal respectively). Test work commissioned by Ecotechnol and clients were undertaken at independent metallurgical laboratories.

Bullabulling Gold Project

A total of three lime demand tests were initially conducted on Bullabulling with two different nanofiltration membranes assessed (ET-NF-A and ET-NF-D). Both membranes led to a significant improvement in lime consumption as outlined in Figure 26 with the buffering effect of magnesium at approximately pH 9 essentially removed. These lime demand tests were conducted in the presence of water only.

Noting the strong buffering of the untreated bore fields water at pH 9, it is likely that Bullabulling once operational would operate at close to pH 9.0-9.5 leading to a lime consumption of approximately 1.01 g/L. To achieve the same pH when using treated water produced via the NF-D membrane, lime consumption would reduce to 0.08 g/L or 92.1%. In reality though, the operating leach pH would be increased to approximately pH 10.5 leading to a lime consumption of 0.13 g/L; still an 87% lime reduction from operating at pH 9 with untreated water.

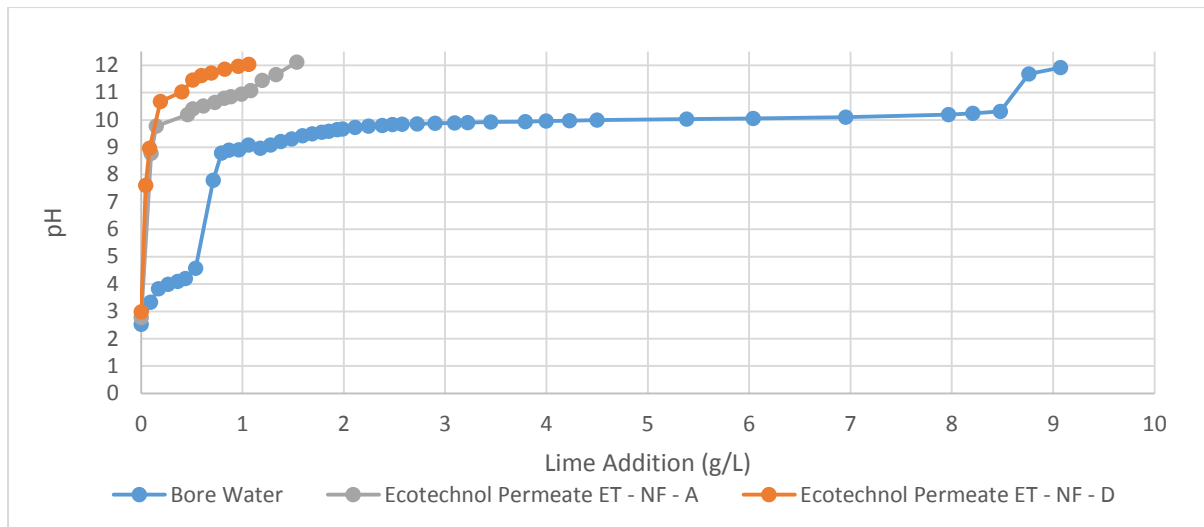


Figure 26: Bullabulling Lime Demand Test

Table 10: Bullabulling Lime Reduction

pH	Bore water	NF-D Permeate	Reduction in Lime (%)
9	1.01	0.08	92.1%
10	4.5	0.13	97.1%
11	8.3	0.40	95.2%
12	9.07	1.06	88.3%

Four subsequent cyanide leach bottle rolls were conducted on weathered and primary ore supplied by Bullabulling to ascertain the reduction in cyanide consumption possible by operating at a higher leach pH. The conditions of these bottle rolls are presented in Table 11 with lime and cyanide consumption results in Table 12.

Table 11: Cyanide Bottle Roll Test work Conditions

Condition	ET-NF-D Permeate		Bore Water	
	Initial	Maintained	Initial	Maintained
pH	10.2	10.2	9.5	9.5
NaCN (ppm)	500	250	500	250
Grind Size (um)	90	90	90	90
Pulp Density (% solids)	40%	-	40%	-
Oxygen Addition	O ₂ Sparging	-	O ₂ Sparging	-

Table 12: Lime and Cyanide Consumption Rates from Cyanide Bottle Rolls

	Oxide Ore			Fresh Ore		
	ET-NF-D Permeate	Bore Water	Reduction (%)	ET-NF-D Permeate	Bore Water	Reduction (%)
Lime Consumption (kg/t)	5.8	12.4	53%	1.6	8.3	81%
Cyanide Consumption (kg/t)	1.21	1.39	13%	0.9	1.45	38%

The removal of magnesium from the water makes it possible to operate at a higher leach pH hence the pH was increased from pH 9.5 using the borefields water to pH 10.2 using permeate. This simulates the probable leach conditions of the future operating plant. As noted in Table 12, the consumption rate of lime reduced by 53% in the oxide ore and 81% in the fresh ore remembering the leach pH increased

by 0.7 pH units. This higher pH led to a reduction in cyanide consumption of 13% in the oxide ore and 38% in the fresh ore.

Based on these encouraging results, Bullabulling commissioned further subsequent independent metallurgical tests to confirm the reduction in lime and cyanide consumption (Proactive Investors, 2014). According to this ASX release, a total of 18 bottle roll leach tests were carried out in the presence of a blended site ore ground to the targeted P_{80} . Untreated borefields water required 1.53 kg/t of lime to achieve a pH of 9.5 whilst treated water required only 0.20 kg/t to achieve the same pH. An 87% reduction. Across all tests, cyanide consumption was 28% lower on average with the use of treated water.

The benefit of pre-treating the borefields water is presented in Figure 27, notably the ability to operate at higher operating leach pH. Bullabulling notes that the implementation of a nanofiltration pre-treatment plant will lead to significant savings in operating costs.

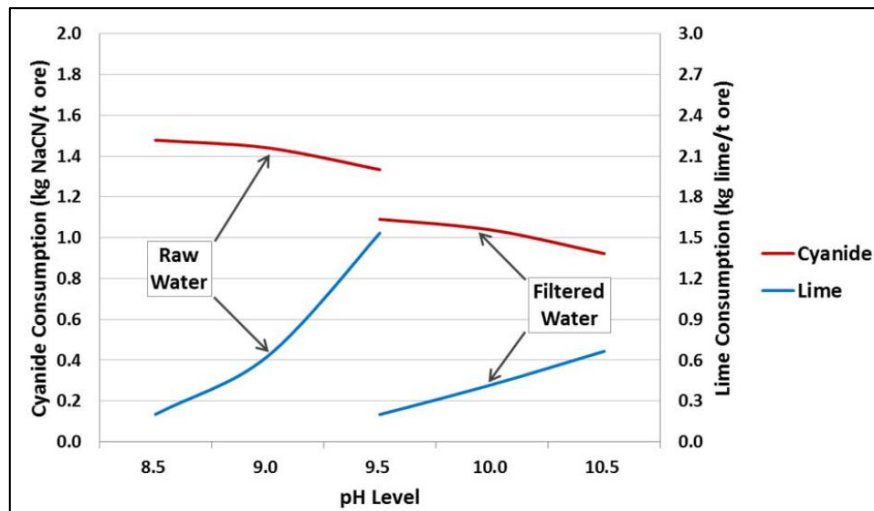


Figure 27: Bullabulling Lime and Cyanide Consumptions

Saracen Carosue Dam Gold Operation

Four lime demand tests were undertaken on Carosue Dam site water in the presence of site ore ground to the leach circuit target P_{80} . In these tests, varying compositions of treated and initial feed water were assessed. The impact of the removal of magnesium on the buffering effect is again clearly demonstrated with the untreated water buffering at pH 9.3-9.5 whilst the 100% permeate weakly buffering at approximately 9.8 before increasing sharply to pH 12 (Figure 28).

The metallurgical staff (Private communication, 2015) at Carosue Dam advised that the actual lime consumption of the process plant averaged 6.26 kg/t to achieve a target leach pH of 9.3. This aligns well with the lime demand test of 6.5 kg/t. Based on these results, lime consumption will reduce by 95% if the pH remains at pH 9.3, or in all likelihood reduce by 62% (2.5 kg/t Vs 6.5 kg/t) since it will now be possible to increase the leach pH to 10.5.

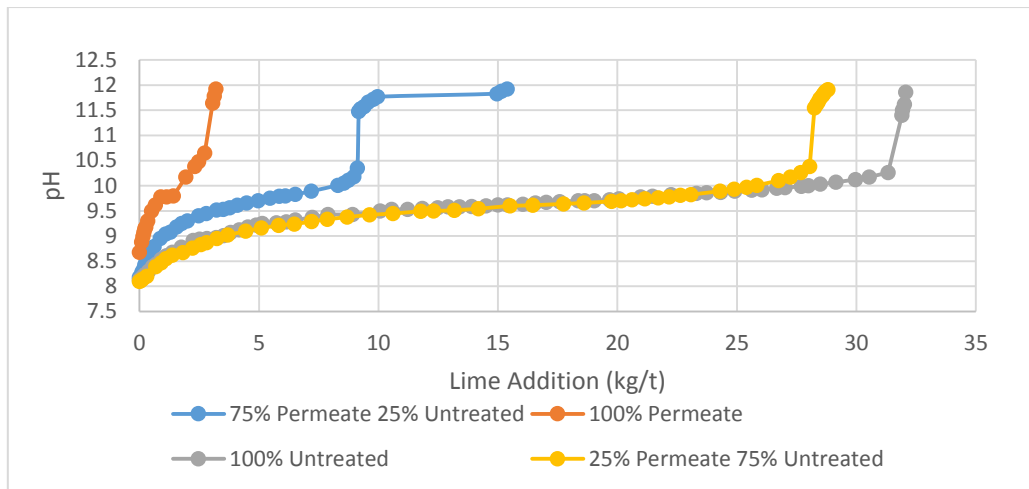


Figure 28: Saracen Lime Demand Test

Table 13: Saracen Lime Consumption at target pH

pH	Bore water	100% Permeate	Reduction in Lime (%)
9.3	6.5	0.35	95%
10	28	1.68	94%
10.5	31.5	2.5	92%

Cyanide leach bottle rolls were conducted on 100% permeate and untreated borefields water with results presented in Table 14. The reduction in lime consumption of only 32% to achieve a pH of 9.6 was not supported by the positive results in the initial lime demand test. It is thought that due to the length of time between the two bottle rolls (approximately 3 months), the water and/or the crushed ore aged impacting the results.

Table 14: Cyanide Bottle Roll Results- Lime and Cyanide Consumption

	Bore Water (pH 9.3)	Permeate (pH 9.6)	Reduction (%)
Lime Consumption (kg/t)	4.7	3.2	32%
Cyanide Consumption (kg/t)	0.9	0.75	17%

APPLICATION OF NANOFILTRATION IN GOLD OPERATIONS

The removal of the key ions responsible for high reagent consumption and scale formation at gold operations using hypersaline water is best performed through the use of a UF and NF system treating the incoming borefields water. Depending on site specifics, a separate feed tank servicing the membrane plant should be installed with appropriate bypass lines in times of maintenance.

As presented in Figure 29, a gold operation requiring 250m³/hr raw water flow to the plant would require 333m³/hr incoming borefields water flow since the test work indicated a water recovery of 75%. Based on site operation, higher water recoveries than shown in test work can be achieved by operating at higher water pressures, multiple stages and with antiscalant optimisation. The brine concentrate consisting of both UF and NF rejects would total 83m³/hr which must be appropriately discharged.

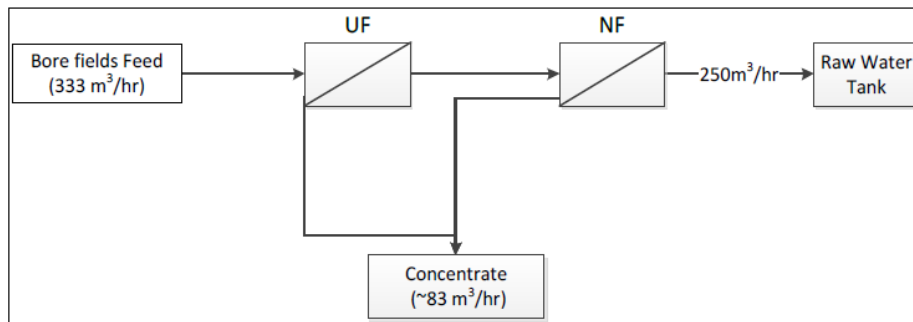


Figure 29: Block Flow Diagram of a Membrane Pre-Treatment System

Brine disposal options include the following:

- In-pit disposal- brine discharge reports to abandoned open pits
- Evaporation ponds (Madin, 2007)- brine discharge is sent directly to lined evaporation ponds
- Seeded Reactors (Madin, 2007) recovering 95-99% of wastewater as a high purity supernatant by precipitating calcium and magnesium with the remaining solids sent to an evaporation pond
- Reinjection of the brine concentrate into the saline groundwater system as previously conducted at Bellevue and Bounty gold projects (Dundon, 2000), and
- Dust Suppression- Esperanza copper/gold mine in Chile sprays brine concentrate onto haul roads etc as a dust suppressant (Aral, 2010).

Aral (2002) has investigated the potential sale of separate constituents of the brine concentrate generated from the membrane treatment of waters from the Murray Darling Basin. Ions which are prolific in hypersaline waters and are further concentrated through the Nanofiltration process are presented below.

- Value Added Magnesium
 - o Epsom Salt (Magnesium sulfate heptahydrate)
 - o Magnesium chloride and
 - o Magnesium hydroxide/Magnesium oxide.
- Gypsum (Calcium sulfate dihydrate) and
- Potassium salts (K_2SO_4).

According to Aral (2002), there is an economic case in treating the saline waters of the Murray Darling basin which are significantly softer than the hypersaline waters of the goldfields region.

FINANCIAL OVERVIEW

A case study is presented with the key design inputs presented in Table 15 and an operating and capital cost breakdown presented in Table 16 and Table 17 respectively. With a TDS of 110,000 mg/L, this water is considered to be relatively easy to treat through an ultrafiltration and nanofiltration facility. The resulting permeate is expected to contain 90% less magnesium and 95% less sulfate than the feed water leading to a 80% and 20% reduction in lime and cyanide consumption.

Table 15: Parameters of UF and NF Plant- Case Study

Parameter	Unit	Value
Permeate Production	m ³ /hr	250
TDS	mg/L (k)	100-110
Water Recovery	%	90
Power Cost	c/kWhr	20
Water Temperature	°C	20
Exchange rate	AUD:USD	0.75
NF Pressure	bar	40

Table 16: Operating Cost of UF and NF Plant Treating Hypersaline Water

Item	\$/m ³ of Permeate	%
Power	\$0.448	78.15%
Chemicals	\$0.024	4.25%
NF Membrane Replacement (4 years)	\$0.037	6.58%
UF Membrane Replacement (5 years)	\$0.021	3.72%
Filters	\$0.001	0.07%
Cleaning	\$0.002	0.32%
Labour	\$0.040	6.92%
	\$0.573	

Table 17: Case Study Capital Cost for a 250m³/hr Permeate UF and NF Plant

Item	A\$ k
Vendor Supplied UF and NF Plant (delivered to site)	4,200
Site Installation (inc piping, feed tank, civils, electrical)	1,000

The cost of power is the largest contributor to the operating cost at nearly 80%. Therefore the source of power to site is of utmost importance with a preference for grid power due to the lower unit cost. All other maintenance and labour costs are included in the operating cost leading to a total variable cost of \$0.573/m³ of permeate produced.

A high level financial analysis is presented in Table 18 calculating the benefit of reducing the lime and cyanide consumption by the test work derived and conservative 80% and 20% respectively. Based on the parameters presented, the payback on initial capital which includes the membrane system and installation costs was 16 months.

A sensitivity analysis is presented in Figure 30 outlining the impact of a reduction in reagent consumption against capital payback. This indicates in this case study, a reduction in lime had the most profound impact on the reduction in overall operating cost.

Table 18: Financial Analysis of Case Study

	Unit	Cost	Comments
Capital Cost			
- Vendor UF/NF Plant	\$M	4.2	+/- 25%
- Site Installation and Associated Costs	\$M	1	Allowance
Operating Cost	\$/m3	0.574	+/-15%
Throughput	ktpa	2,500	Client Provided
Availability	%	92%	Client Provided
Water treatment requirement	m3/hr	250	Client Provided
Lime consumption	kg/t	6.5	Client Provided
Cyanide consumption	kg/t	0.5	Client Provided
Reagent Unit Cost			
- Lime	\$/t	330	Recent cost data
- Cyanide	\$/t	3,300	Recent cost data
% reduction in lime consumption	%	80	Test work Results
% reduction in cyanide consumption	%	20	Test work Results
Cost Comparison			
- Lime (no treatment)	\$M	5.4	Calculation
- Lime (with UF/NF)	\$M	1.1	Calculation
- Cyanide (no treatment)	\$M	4.1	Calculation
- Cyanide (with UF/NF)	\$M	3.3	Calculation
Total Cost Saving in Reagents	\$M pa	5.1	Calculation
Total Operating Cost of NF	\$M pa	1.16	Calculation
Payback	months	15.8	Calculation

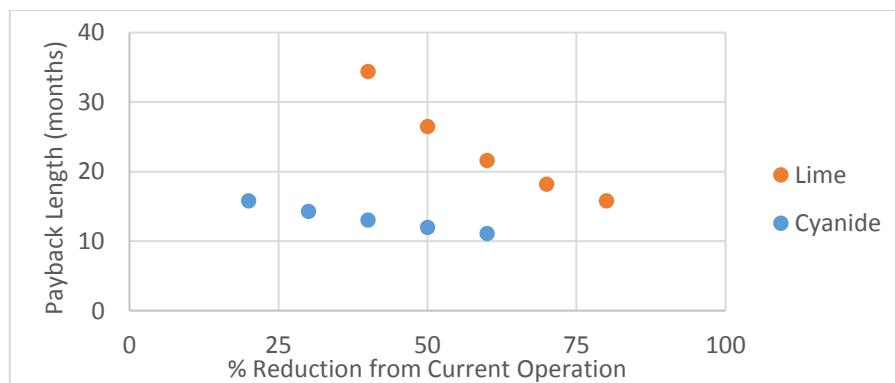


Figure 30: Sensitivity Analysis

CONCLUSION

Membrane technologies for water treatment have been successfully implemented for over forty years across a number of industries and applications. Test work undertaken by Ecotechnol and independent laboratories on borefields waters sourced from a number of Western Australian Goldfields gold operations has now demonstrated the ability to economically and effectively remove the ions responsible for the high lime and cyanide consumption rates characteristic of this region.

More than 80% of magnesium, calcium and sulfate can be concentrated into approximately 20% of the initial feed stream via an ultrafiltration then nanofiltration plant. The remaining purified water can then be utilised in the processing plant improving plant economics and ease of operation.

ACKNOWLEDGEMENTS

The authors would like to acknowledge Bullabulling Gold and Saracen Carosue Dam for their support and permission to publish the test work results.

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