TREATMENT OF LANDFILL LEACHATE FROM HAZARDOUS AND MUNICIPAL SOLID WASTE

Report to the

Water Research Commission

by

J J Schoeman, A Steyn, J L Slabbert & E A Venter

Division of Water, Environment and Forestry Technology, CSIR
P O Box 395, Pretoria 0001

WRC Report No : 1167/1/03
ISBN No : 1-86845-976-4

FEBRUARY 2003

Disclaimer

This report emanates from a project financed by the Water Research Commission (WRC) and is approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the WRC or the members of the project steering committee, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

EXECUTIVE SUMMARY

1. INTRODUCTION AND AIMS

1.1 General Background

Continuing industrial and commercial growth in many countries around the world in the past decades have been accompanied by rapid increases in Municipal Solid Waste (MSW) and Industrial Solid Waste (ISW) generation. At present, landfilling is the most popular way of solid waste disposal and landfilling will continue to be the primary means of MSW and ISW disposal in future. Besides scarcity of available landfill sites in certain regions, a large amount of leachate (originating from water which has percolated through emplaced refuse) generated from a landfill site poses a major problem of landfill disposal of MSW and ISW. Proper treatment of the leachate has therefore been a challenging problem confronting the local authorities.

Landfill leachate has been generally known as a high-strength waste water that is most difficult to deal with. This is due primarily to its large variability of organic, inorganic and heavy metal contents, strongly depending on the age and type of solid waste of a landfill site. Satisfactory treatment of leachate is thus no easy task. The most popular treatment of landfill leachate in the past was anaerobic digestion or the aerobic activated sludge method. These methods were known to be inadequate in handling such a difficult treatment task. In more recent decades, search for alternative treatment methods had focussed on various sophisticated technologies. These included advanced biological, chemical and physical treatment methods. Researchers have studied leachate treatment using the sequencing batch reactor (SBR) method. Other researchers studied the aerobic treatment of a high-strength leachate preceded by ion-exchange and lime addition to effect inorganic removal prior to biological treatment. An efficient biological activated carbon fluidized bed process was developed. Chemical oxidation using strong oxidising agents, such as Fenton's reagent, photo-assisted H_2O_2, ozone or UV-Vis light, was reported. Membrane processes have received considerable attention in the past decades, as reported by many investigators. In fact, it has been indicated that membrane technology is widely practiced in many European countries in dealing with leachate treatment. In two recent investigations, researchers employed an electrochemical method in decomposing the refractory organic and inorganic components in the leachate. Both investigations found efficient removal of the chemical oxygen demand (COD) and ammonia concentration in the leachate.

The above brief review of the literature has indicated that physical (membrane) and chemical methods appear to offer quite good alternatives to biological treatment. Reverse osmosis should be a very suitable technology for the treatment of hazardous leachates with a relatively low TDS concentration. However, the salinity concentration of certain industrial leachates may be too high to consider RO for treatment. Consequently, evaporation and electrodialysis technologies are the remaining technologies that should be considered for the treatment of high TDS concentration leachates. Consequently, chemical methods and

membrane technology (RO for low TDS leachate and ED for high TDS leachate) were selected to study the treatment of two selected leachates in South Africa.

1.2 South African Background

The minimum requirements for waste disposal to landfill sites according to the Department of Water Affairs and Forestry (DWAF) in South Africa is that all hazardous waste sites should have a leachate management system. The minimum requirements for the classification, handling and disposal of hazardous waste according to DWAF states that all leachates are hazardous, and DWAF is beginning to push waste disposal companies to manage leachate treatment effectively. Therefore, suitable technologies will be required in the near future for the successful treatment of these hazardous leachates.

Two types of hazardous leachates are produced in landfill sites in South Africa. The one is a high TDS (50 to 100 g/ℓ), high organic (10 to 80 g/ℓ COD) concentration leachate containing hazardous chemicals like phenols, sulphides, ammonia-nitrogen, chromium, etc. (Industrial Solid Waste Leachate, ISWL). This ISWL is unique to South Africa, with its relatively low rainfall compared with Europe, and past waste disposal practices. The other leachate is a low TDS (2 to 6 g/ℓ), low organic (2 to 6 g/ℓ COD) concentration leachate (Municipal Solid Waste Leachate, MSWL). The ISWL cannot be discharged into the municipal sewer system, because the high salinity levels, as well as the other hazardous chemicals (phenols, chromium, heavy metals) will adversely affect the biological treatment process. However, the MSWL can, in some cases, be directly discharged into the municipal sewer system without any problems, depending on the dilution water available. Municipal solid waste leachate cannot be discharged into the municipal sewer system where little dilution water is available, and where hazardous chemicals are present in the leachate. Positive water balances at landfill sites can also result in pollution of the water environment. Therefore, the leachates should be treated prior to disposal, to prevent adverse effects on municipal biological treatment processes, and to prevent pollution of the water environment.

Industrial solid waste leachate (50 to 100 g/ℓ TDS) cannot be treated with reverse osmosis (RO) for salinity removal, because the salinity levels are too high. Electrodialysis (ED), however, which can handle high salinity levels, has the potential to desalinate this type of leachate effectively. However, membrane fouling may be a problem. Pretreatment of ISWL with adsorbents or absorbents (ash) and other physical/chemical (coagulation-flocculation, filtration, softening etc.), and membrane methods is, therefore, considered to be important for the removal of most of the membrane foulants prior to ED desalination. Electrodialysis membranes are also available, which are claimed to be more fouling-resistant than conventional electrodialysis membranes, and these membranes were used in the study.

Preliminary work on an ISWL from Holfontein has shown that approximately 50 to 60% of the salinity could easily be removed with electrodialysis desalination. However, it was not possible to remove the remaining salinity, consisting of organic complexes, with

electrodialysis. Reverse osmosis, however, has the potential to desalinate the electrodialysis diluate further to very low concentration levels.

It should be possible to desalinate MSWL effectively with RO for direct discharge into the water environment. This will help to prevent water pollution effectively. However, very little information is available in South Africa regarding the performance and economics of RO for the desalination of MSWL. The newly developed spiral wrap RO system from Grahamtek Systems which is claimed to have the ability to treat high fouling effluents effectively without chemical pretreatment, was evaluated in the study. Tubular cellulose acetate RO membranes, which are manufactured in South Africa, were also evaluated in the study, together with tubular polyamide RO membranes, which are applied overseas for the treatment of MSWL. The tubular cellulose acetate RO membranes are claimed to be more resistant to membrane fouling than polyamide tubular and conventional spiral wrap RO membranes.

There are approximately eight strong leachate-producing landfill sites in South Africa and many more weak leachate-producing sites. Several industries are also producing strong effluents. Therefore, the process technology that has been developed in this study could find wide application in South Africa.

1.3 Aims

The main aim of this project is to develop ED and RO process technology for the treatment of hazardous industrial leachate with high TDS and high organic concentrations (ISWL; Holfontein waste disposal site, Springs). The secondary aim of this project is to develop RO process technology for the treatment of hazardous municipal leachate (MSWL) with low TDS and low organic concentrations (Bisasar Road waste disposal site, Durban). The different tasks include : -

(a) Concise literature overview.

(b) Characterisation of the leachates.

(c) Determination of the biodegradability of the leachates.

(d) Evaluation of pretreatment of the leachates for the removal of suspended material, organics and inorganics.

(e) Determination of the fouling potential of the leachates for ED and RO membranes.

(f) Evaluation of membrane cleaning strategies of fouled ED and RO membranes with commercially available membrane cleaning agents.

(g) Evaluation of ED and RO performance for the desalination/ concentration of the leachates.

(h) Determination of the preliminary economics of the processes.

(i) Technology transfer through ED and RO pilot tests on site, and determination of the economics of the processes.

(j) Development of the capacity of Technikon students to operate and maintain laboratory and membrane pilot plants.

2. LITERATURE OVERVIEW

2.1 Industrial Solid Waste Leachate (High TDS and COD)

(a) Various methods of leachate treatment (membrane bioreactor, activated sludge, activated carbon, chemical oxidation, reverse osmosis, etc.) may be combined in various modes with other standard chemical engineering unit processes, providing a huge range of variable options are available from which a solution can be engineered to optimise the balance between cost and quality. Some examples of combinations of these processes that have been shown to be successful in previous studies are as follows : -

(i) Chemical oxidation - biological oxidation - chemical precipitation
(ii) Biological oxidation - chemical oxidation - biological oxidation - chemical preciptiation
(iii) Chemical precipitation - chemical oxidation - biological oxidation
(iv) Chemical precipitation - biological oxidation - chemical oxidation - biological oxidation.

In researching leachate treatment requirements in some detail, it has become apparent that the solution may be found not only in leachate treatment technology, but in chemical process technology as well. Alternatives inclues : -

(i) The Phenosolvan processes for the extraction of phenols from water and other process streams,
(ii) Activated carbon or related chemical dosing technology (ash treatment, chemical treatment, coagulation-flocculation, filtration, etc.)

(b) A number of laboratory and small-scale tests were carried out in 1999 in South Africa to test the application of various treatment options on ISWL (Holfontein). The general conclusions from early test work were : -

(i) Evaporational/crystallisation appears to be the best technology for the first step in the process. It does not, however, on its own, produce water of the required quality. Further treatment of the condensate produced by the evaporator/crystalliser is required.

(ii) Almost no biological activity occurs in the raw leachate due to the toxic nature of the leachate. To obtain significant biological activity requires substantial dilution of the feed.

(iii) Inadequate reduction in COD and dissolved salts were obtained with electrodialysis treatment of the leachate. Membrane fouling was also experienced. However, ED with fouling resistant membranes hold promise for the desalination/concentration of this high TDS, high COD-type leachate.

Pilot studies were subsequently conducted to evaluate COD removal with an evaporator/crystalliser. This showed that with a leachate containing 80 000 mg/ℓ COD, the condensate would contain about 14 000 mg/ℓ COD. Dissolved salts in the leachate of 100 000 mg/ℓ were reduced to about 300 mg/ℓ. By raising the pH to 10,5, the COD in the condensate could be reduced to about 6 000 mg/ℓ COD. At a pH of 12,5, the COD was reduced to 1 300 mg/ℓ.

No COD reduction was found with anaerobic biological treatment of the condensate from the evaporator/crystalliser. Aerobic biological treatment, in contrast, reduced the COD content of the condensate from 3 000 mg/ℓ to about 400 mg/ℓ.

Activated carbon removed the COD from the condensate to about the required level of 65 mg/ℓ for discharge to the water environment.

Ammonia was not removed adequately from the condensate (1 000 mg/ℓ) in either the activated carbon columns or the aerobic biological treatment plant. A number of treatment options to reduce the ammonia are being tested. Treatment options include :

(i) Distillation of the final condensate after activated carbon treatment;
(ii) Air stripping of the condensate prior to biological treatment.

The treatment options for the treatment of the leachate are :

(i) Evaporator/crystalliser plus activated carbon and
(ii) Evaporator/crystalliser plus aerobic biological treatment plus activated carbon.

2.2 Municipal Solid Waste Leachate (Low TDS and COD)

(a) Reverse osmosis plants in operation in Europe and other parts of the world have proven the suitability of the process for the treatment of landfill leachate and similar waste waters. This does not only apply to the unsurpassed discharge qualities, but also to the high availability of the installations (in general >90%).

Reverse osmosis plants may be operated with or without biological pretreatment. In general, a multi-staged RO is required if biological pretreatment is not installed. With biological pretreatment, a 1-stage RO is in most cases sufficient.

The ideal process combination for the treatment of small and medium-sized leachate quantities is biological pretreatment and removal of solids by ultrafiltration. Sludge separation by ultrafiltration, which allows one to obtain a high content of solids in the biological reactor. This makes the installation of small volume reactors possible. The process combination for the treatment of small and medium-sized leachage quantities is biological pretreatment/reverse osmosis, fluidized bed granulator.

With the new drying technique of evaporation and fluidized bed granulation a solution has finally been found to the much discussed issue of how to properly treat RO concentrate.

(b) Results obtained during the operation of an increasing number of plants under very different conditions prove that RO is a very effective instrument for the purification of landfill leachate if all design criteria and requirements specific for landfill leachate have been taken into consideration, and if an adapted module system as well as correlated technologies are used. This includes high pressure RO, with operating pressure up to 120 bar and/or NF in combination with a controlled crystallisation process, that allows permeate recovery rates of more than 95%.

The elimination of the negative impact of landfill leachate on the environment can be achieved with membrane filtration due to the dramatic minimization of residual waste to be processed or immobilized and due to the high quality of the purified water discharged back to nature. The combination of processes designed for this purpose is one example for a sustainable environmentally friendly development.

(c) The treatment of highly polluted water by RO is a reliable and economic operation and can be considered as state of the art. The process can produce water of any required quality - if necessary in a cascaded operation.

A major problem of waste water treatment is the water recovery rate, which should be near to 100%, realized in a simple and energy-efficient process combination. As demonstrated by long-term experiments in pilot plants and on technical scale, this can be achieved by the addition of NF and 200 bar high pressure RO to the 60/120 bar RO stage. The integration of

the simple mechanical unit operation crystallizer/hydrocyclone/filtration promises an almost zero discharge process.

(d) In RO/NF, research and development concentrate on shifting the limits of processes to very high water recoveries, i.e. the development of 'almost zero discharge' processes. This is also strongly related to module development.

(e) In the course of the treatment of landfill leachates by reverse osmosis (single and multi-level) a concentrate of 15 to 25% of the amount of the raw leachate accumulates, the difference being determined by the chosen concentration factor. In principle, there are three possibilities for the disposal of concentrate :

(i) Return to the landfill;
(ii) Evaporation and drying-out;
(iii) Incineration (for example, in a waste incineration plant).

(f) A leachate treatment pilot plant has been established and commissioned at the Bisasar Road Landfill Site in Durban to access the treatability of the landfill leachate. Complete biological nitrification and denitrification of landfill leachate has been achieved inside the same SBR unit, and the viability of establishing full-scale treatment plants at landfill sites has been assessed. Levels of some 2 000 mg/ℓ of ammonia-nitrogen have shown to be completely and consistently nitrified to nitrate-N, and in turn released as harmless nitrogen gas, resulting in only negligible levels of nitrogenous components in the treated effluent. Sludge build-ups are minimal, and sludge waste has proven to be an infrequent occurrence. Future and ongoing research work involves the assessment of denitrification processes utilising waste molasses, and the feasibility of further polishing treatment aimed in particular at the removal of residual COD levels, using constructed wetlands.

3. CHARACTERISATION OF THE INDUSTRIAL SOLID WASTE LEACHATE

The ISWL from Holfontein contains very high inorganic (TDS approximately 105 600 mg/ℓ) and organic (COD approximately 64 000 mg/ℓ) concentrations. The inorganic cations consist mainly out of sodium (67% of cations), ammonium-nitrogen (12,5%) and potassium (11%) while the major anions are chloride (56,4% of anions), sulphate (31,4%) and bicarbonate (12,3%). Significant quantities of magnesium (5,6%) and calcium (3,9%) ions are also present. Other hazardous compounds include phenols, lead, chromium, arsenic, nickel, etc. It is interesting to note that the phenol concentration of the leachate is high (900 mg/ℓ). The BOD/COD ratio is low (0,26). Therefore, it will be difficult to biodegrade the leachate. The relatively high volatile fatty acid concentration (approximately 6 600 mg/ℓ) shows, that some of the organics will be readily biodegradable. Biodegradability of the leachate, however, could be inhibited by toxic compounds in the leachate.

4. BIODEGRADABILITY OF THE INDUSTRIAL SOLID WASTE LEACHATE

Biodegradable tests (respirometer) have shown that it should be possible to biodegrade the ISWL from Holfontein to some extent. COD removals, however, were low (7,3 to 8%). Significantly higher COD removals (30%) were obtained on the ED treated leachate because most of the salinity had been removed from the leachate. Dilution (5x) of the leachate improved COD removal to 34,9%. The addition of biosupplements to the leachate also improved its biodegrability somewhat. COD removal was improved from 5,4% (no biosupplement) to 7,8 to 8% with the addition of biosupplements. Bio-enhanced treatment of the leachate in an SBR unit has shown that approximately 20 to 45% of the COD could be removed.

5. ASH PRETREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR ORGANICS AND INORGANICS REMOVAL

Very good phenol removals (900 mg/ℓ to 58 mg/ℓ, 93,6%) were obtained when the ISWL from Holfontein was treated with Iscor ash (200 g/ℓ). COD removal, however, was not that good (64 000 mg/ℓ to 54 700 mg/ℓ, 14,5%). COD removals increased with increasing ash dosage and similar COD removals were obtained with Lethabo and Sasol ash samples. The ashes added significant quantities of calcium (approximately 500 mg/ℓ to 3 000 mg/ℓ Ca at an ash dosage of 200 g/ℓ) to the leachate. No significant removals of magnesium, manganese, barium, strontium and silica were obtained at relatively low ash dosages. However, significant quantities of magnesium, manganese, ammonia and iron were removed at high ash dosage in the case of Sasol ash. Sludge volume comprises approximately 27% of the treated volume in the case of Iscor ash (200 g/ℓ). Iscor ash appears to be the best candidate for treatment of the leachate because it can remove significant quantities of phenol from the leachate.

6. LIME, CAUSTIC SODA AND SODA-ASH PRETREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR ORGANICS AND INORGANICS REMOVAL

Excellent magnesium (858 to 25mg/ℓ, 97,1%), manganese (29,1 to 0,45 mg/ℓ, 98,5%), barium (0,32 to 0,03 mg/ℓ, 90,6%) and iron (91,6 to 12,6 mg/ℓ, 86,2%) removals were obtained with caustic soda treatment of the ISWL from Holfontein at high pH (pH 12). Good strontium removal was also obtained (3,83 to 1,5 mg/ℓ, 60,8%). Significant amounts of COD (64 500 to 34 500 mg/ℓ, 46,5%) and phenols (900 to 380 mg/ℓ, 57,8%) could also be removed with caustic soda treatment. However, a high concentration of sodium was added to the leachate while almost no calcium was removed. High quantities of manganese (29,1 to 0,24 mg/ℓ, 99,2%), barium (0,32 to 0,03 mg/ℓ, 90,6%) and iron (91,6 to 15,7 mg/ℓ, 82,9%)

could be removed from the leachate with lime treatment at high pH. However, high concentrations of calcium were added to the leachate. A significant amount of COD (64 000 to 38 000 mg/ℓ, 40,6%) could also be removed with lime treatment, while very little phenol was removed. High concentrations of calcium (983 to 164 mg/ℓ, 83,3%) could be removed from the leachate with soda ash treatment. Significant amounts of manganese (29,1 to 4,0 mg/ℓ, 86,3%) and barium (0,32 to 0,08 mg/ℓ, 75%) could also be removed from the leachate with soda ash and lime treatment. Caustic soda and soda ash treatment of the leachate showed that excellent removals of calcium (983 to 166 mg/ℓ, 83,1%), magnesium (858 to 247 mg/ℓ, 71,2%) and barium (0,32 to 0,1 mg/ℓ, 68,8%) could be obtained. However, sodium was added to the leachate while some phenol removal was also obtained. Excellent removals of calcium (983 to 130 mg/ℓ, 86,8%), manganese (29,1 to 2,3 mg/ℓ, 92,1%) and barium (0,32 to 0,1 mg/ℓ, 68,8%) were obtained with treatment of the leachate with Iscor ash, caustic soda and soda ash (200 g/ℓ ash, 12 g/ℓ Na_2CO_3, 6,9 g/ℓ NaOH). Strontium and iron removals, however, were not that good. A significant amount of sodium was also introduced into the leachate. Poor COD removals were also obtained. However, the removal of phenol (900 to 66 mg/ℓ, 92,7% removal) which can attack membranes was good. Therefore, the abovementioned combination of ash and chemicals were selected for the treatment of the leachate prior to desalination. Chemical treatment costs were determined at R20,40/kℓ for soda ash and R22,08/kℓ for caustic soda (Total R42,48/m^3). (Note: the chemical dosages are not necessarily the optimum dosages for the desalination of the leachate).

7. COAGULATION/FLOCCULATION PRETREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR THE REMOVAL OF SUSPENDED MATERIAL AND ORGANICS

Good suspended solids removals 63 to 82% (3 778 to 1 405 mg/ℓ, and 3 778 to 698 mg/ℓ) could be obtained with treatment of the ISWL from Holfontein with alum, ferric chloride, poly-aluminium chloride and a polyelectrolyte. Good suspended solids removal could also be obtained with filtration (3 778 to 988 mg/ℓ, 74%) of the leachate. Lowering of the pH also helped to increase the removal of suspended solids (3 778 to 515 mg/ℓ, 86,3%) in the case of ferric chloride treatment. This type of effluent will require further filtration for the successful removal of suspended solids prior to desalination. Organics removals, however, were poor.

8. NANOFILTRATION FOR THE TREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE

Good organic removals were obtained when the ISWL from Holfontein was treated with nanofiltration (approximately 59% (69 000 to 28 500 mg/ℓ COD), to 42% (69 000 to 40 000 mg/ℓ) (AFC 40, MPT31 and MPT 36 membranes). However, a significant amount of organics permeate the membranes (24 to 37%) showing that there are low molecular mass organics present in the leachate. The monovalent ions (Na^+, Cl^-) permeate the membranes

preferentially with the result that a good separation could be obtained between the monovalent and divalent ions (Ca^{2+}, SO_4^{2-}). Approximately 52,5% chloride ions, for example, was present in the permeate compared to divalent sulphate ions (18%, AFC 40 membranes). Also, significantly more monovalent sodium ions were present in the permeate (70%) than divalent calcium (14,6%) and magnesium (15%) ions. More monovalent chloride ions (56,2%) were present in the permeate than sulphate ions (10,3%) (MPT 31 membranes). The same applied to the monovalent sodium (44%) and calcium ions (31,3%). The chloride ions in the permeate in the case of the MPT 36 membranes were 58,1% while only 25,1% sulphate ions were present. The same applies to the monovalent sodium (48%), potassium (42,2%) and the divalent calcium (30,9%) and magnesium ions (33%). Therefore, the nanofiltration permeate should be more suitable for further desalination because most of the scale-forming chemicals have been eliminated from the permeate. The membranes, however, were fouled by the leachate. Preliminary tests, however, have indicated that it should be possible to clean the membranes with chemicals. Nanofiltration has potential for the treatment of leachate and further investigations should be done in this regard.

9. FOULING POTENTIAL OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR ELECTRODIALYSIS MEMBRANES

The fouling potential of the ISWL from Holfontein for several ion-exchange membranes was evaluated in a membrane fouling test cell. It was found that most of the anion membranes (Selemion AMV, Tokuyama Soda ACS, Selemion ASV) were fouled by the leachate. However, the Tokuyama Soda AFN anionic membrane, Ionics AR204SZ RA anionic membrane and the Tokuyama Soda AXE anionic membrane were far less fouled by the leachate. The AFN anionic membrane appeared to be more resistant to fouling by the leachate as indicated by the potential drop across the membrane during the fouling studies and by membrane resistance measurements. The AFN anionic membrane was therefore selected for further studies.

10. LONG-TERM FOULING STUDIES AND MEMBRANE CLEANING STRATEGIES OF THE FOULED ELECTRODIALYSIS MEMBRANES WITH COMMERCIALLY AVAILABLE MEMBRANE CLEANING AGENTS

Long-term membrane fouling studies were conducted in a membrane fouling test cell (approximately 900 hours) to further evaluate the fouling potential of the leachate (Holfontein) for the membranes and to evaluate cleaning of the fouled membrane. The study has indicated that the AFN anionic membrane can be fouled by the pretreated ISWL. However, it was demonstrated that it should be possible to clean the fouled membranes with mechanical cleaning, polarity reversal and cleaning of the membrane with salt solution (3%) at high pH (pH 11,5).

11. ELECTRODIALYSIS TREATMENT FOR THE DESALINATION/CONCENTRATION OF THE PRETREATED INDUSTRIAL SOLID WASTE LEACHATE

Electrodialysis treatment of the pretreated ISWL (Holfontein) in a laboratory scale ED unit (AFN and CMX membrane) has shown that the TDS of the leachate could be reduced from 118 485 mg/ℓ to 17 236 mg/ℓ (85,5% TDS removal). Therefore, an excellent removal of TDS could be obtained with ED treatment of the leachate. Approximately 49% (54 800 to 28 100 mg/ℓ) of the COD could also be removed from the leachate. Water recovery was 62,5%. This implies that the brine volume comprises approximately 38% of the treated leachate. The salt concentration of the brine is high (215 980 mg/ℓ) and the brine should be further treated for safe disposal. Analysis indicated that there was some reduction in the ion-exchange capacity and gel water content of the anion-exchange membranes after ED treatment. This indicates some degree of membrane fouling. However, the extent of the membrane fouling does not appear to be too serious. It was also shown that the desalinated stream (ED product) was far less toxic than the untreated and pretreated leachate and ED brine. A 16 time dilution of the ED product will be required to avoid acute effects on the aquatic life if the ED product is discharged into the water environment. An 800 times dilution will be required in the case of the ED brine and less than a 159 times dilution on the pretreated leachate (ash and chemicals).

12. PERFORMANCE OF ED ON PILOT-SCALE FOR THE TREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE

12.1 Pilot-scale Performance

Treatment of the pretreated leachate from Holfontein in an ED pilot plant (batch treatment) has shown that the TDS of the leachate could be reduced from 104 260 mg/ℓ to 11 903 mg/ℓ (88,6% removal). COD removal was 48,6% (55 400 to 28 500 mg/ℓ) and water recovery was 47,5%. This implies that the brine will comprise approximately 50% of the treated leachate. The TDS of the brine is high (169 805 mg/ℓ), and the brine should be further treated in an evaporator prior to final disposal. This will ensure that a high water recovery and low brine volume can be obtained. It was also shown that the ED product (19 350 mg/ℓ) could be further desalinated to 645 mg/ℓ with RO (96,7% removal). COD was removed from 25 500 to 935 mg/ℓ (96,3% removal). Therefore, a very good quality water could be produced with combination of ED and RO treatment of the leachate. The RO brine which should comprise approximately 20% of the RO feed should also be treated in an evaporator, to increase water recovery and to reduce the brine volume. It also appears that it should be possible to control fouling of the RO membranes with chemical cleaning.

Feed-and-bleed ED pilot tests have shown that the TDS of the leachate could be reduced from 116 235 mg/ℓ to 2 435 mg/ℓ (5 stage ED). Brine volume comprised approximately 41%

of the treated leachate. Water recovery should be more than 90%. It also appears that membrane fouling should not be a serious problem.

Significant quantities of magnesium, potassium, aluminium, iron, manganese and titanium can leach from the sludge produced after chemical treatment of the leachate prior to desalination. However, the concentration of these elements in a leachate is not considered to be dangerous where the sludge is disposed of in a landfill. Organics at low concentrations can also leach from the sludge. These organics are also not considered to be a problem where the sludge is disposed of in a landfill.

12.2 Economics

The capital cost of an 80 kℓ/d (feed) ED plant is estimated at R15,5 million (approximately R23,3 million for a 140 kℓ/d (feed) plant). Operational costs including electrical energy for ion transport, membrane replacement costs (1 year lifetime), chemical pretreatment and pumping costs are estimated at R188,24/kℓ (1 year membrane lifetime). Operational costs are estimated at R93,55/kℓ for a three year membrane lifetime. Cleaning chemical costs are not included in the operational costs.

13. CHARACTERISATION OF THE MUNICIPAL SOLID WASTE LEACHATE

The electrical conductivity of the Bisasar Road leachate were determined as 1 291 and 1 650 mS/m in two cases. The TDS were determined as 7 070 mg/ℓ in one case. The COD of the Bisasar Road leachate were 2 427 and 2 000 mg/ℓ in two cases. Both the conductivity (250 mS/m) and the COD (75 mg/ℓ) of the leachate do not comply to the requirements of the general discharge standard. The BOD/COD ratio was only 0,13 in one case and 0,48 in another case. The low ratio suggests that it will be difficult to biodegrade the organics in the leachate, while the higher ratio suggests that the leachate should be readily biodegradable.

The ammonia-nitrogen concentrations of the Bisasar Road leachate are high (1 271 and 980 mg/ℓ in two cases). The chloride concentrations are also high (1 790 and 2 625 mg/ℓ) Both the ammonia-nitrogen (general discharges standard of <10 mg/ℓ to the water environment) and chloride (discharges requirement of <1 000 mg/ℓ to the sewer) concentrations are higher than the requirements of the discharge standards. The sodium (897 and 1 620 mg/ℓ) and potassium concentrations (1 022 and 1 150 mg/ℓ) are also high.

The heavy metals concentration (Cr, 0,17 mg/ℓ, Pb, 0,126 mg/ℓ; Ni, 0,2 mg/ℓ) were higher in one case than in another case (Cr, 0,05 mg/ℓ; Pb <0,2 mg/ℓ; Ni, 0,09 mg/ℓ). The lead concentration does not comply to the general standard (0,1 mg/ℓ) for discharge to the water environment. The barium (0,495 mg/ℓ), strontium (1,09 mg/ℓ), iron (3,16 mg/ℓ) and manganese (0,382 mg/ℓ) concentrations are also high. These concentrations can affect the performance of a membrane process adversely if not properly controlled during treatment.

The chemical composition of the Bisasar Road leachate show that the leachate should be treated prior to discharge into the water environment. The leachate, however, should comply to the quality requirements for discharge to the Durban Metro Sewer if the chloride concentration could be reduced to less than 1 000 mg/ℓ.

The Mariannhill leachate which is a weaker leachate than the Bisasar Road leachate also does not comply to the discharge quality requirements to the water environment and to sewer and should be treated prior to discharge.

The salinity, COD and ammonium-nitrogen concentrations of other leachates in the Durban area are also too high for discharge to the water environment. Some of the heavy metal concentrations are also too high. The chloride concentrations of most of the leachates are too high for discharge to sewer (>1 000 mg/ℓ Cl). Consequently, most of the MSW leachates should be desalinated prior to discharge to either the water environment and the sewer system.

14. FOULING POTENTIAL OF THE MUNICIPAL SOLID WASTE LEACHATE FOR TUBULAR CELLULOSE ACETATE AND POLYAMIDE MEMBRANES AND MEMBRANE CLEANING STRATEGIES

Batch RO tests have indicated that the leachate from Bisasar Road will foul tubular cellulose acetate and polyamide membranes. However, it was shown in both cases that it should be possible to control membrane fouling. Preliminary results have indicated that the clean water flux (CWF) in the case of the cellulose acetate membranes should be restored with chemical cleaning (acid and STPP and EDTA cleaning solution). Less fouling was encountered with the polyamide membranes. The CWF was much higher after treatment than in the case of the cellulose acetate membranes. It also appears that it should be possible to restore the CWF with chemical cleaning (SLS and EDTA cleaning solutions).

15. PERFORMANCE OF TUBULAR REVERSE OSMOSIS FOR THE CONCENTRATION/DESALINATION OF THE MUNICIPAL SOLID WASTE LEACHATE

The performance of tubular reverse osmosis (cellulose acetate and polyamide membranes) for the desalination/concentration of the leachate was evaluated in a pilot scale RO unit in the batch mode of operation. The initial permeate flux was approximately 600 ℓ/m^2.d (cellulose acetate membranes) and the flux decreased as a function of percentage water recovery as a result of the increased osmotic pressure of the feed at higher water recoveries (<40%). Almost identical flux results were obtained with two runs that were conducted (up to approximately 70% water recovery). The initial and CWF at the end of the runs were almost the same. This shows that it should be possible to control membrane fouling with flow reversal and sponge ball cleaning.

The TDS of the leachate could be reduced from 8 975 to 348 mg/ℓ (96,1% removal). Therefore, an excellent quality water could be produced with RO treatment of the leachate. Ammonia-nitrogen, however, was only reduced from 882 to 82 mg/ℓ (90,7% removal). Therefore, ammonia-nitrogen removal was not that good. However, the removal of the other ions like chloride (92,4%), sulphate (99,5%), calcium (98,8%), magnesium (99,7%), potassium (94,9%), sodium (96,4%) and COD (97,7%) were excellent. The quality of the RO product with the exception of ammonia-nitrogen complies to the discharge quality requirements (water environment and sewer).

The initial permeate flux in the case of the polyamide membrane was approximately 1 200 ℓ/m^2.d and the flux also decreased as a function of percentage water recovery as before. The initial and CWF at the end of the runs were approximately the same. Therefore, it again appears that membrane fouling should not be a problem with the treatment of this type of effluent if flow reversal and sponge ball cleaning are applied.

The permeate flux through the polyamide membranes module was significantly higher than through the cellulose acetate membrane module (1 200 ℓ/m^2.d to 600 ℓ/m^2.d for polyamide membranes at end of run and 600 ℓ/m^2.d to 350 ℓ/m^2.d for cellulose acetate membranes at end of run, 70% water recovery). Therefore, more product water can be produced with the polyamide membrane module (0,181 m^2) than with the cellulose acetate membrane module (1,75 m^2).

Higher TDS removals were obtained with the polyamide membranes (97,9%) than with the cellulose acetate membranes (96,1%). Conductivity removals were 96,9% for the polyamide and 93,2% for the cellulose acetate membranes. A similar ammonia-nitrogen removal was obtained with the polyamide membranes (980 to 81 mg/ℓ, 91,7% removal) than with the cellulose acetate membranes (882 to 82 mg/ℓ, 90,7%). Between 98% to 100% removals of chloride, sulphate, calcium, magnesium, potassium and sodium were obtained. Better lead, nickel and phenol removals were also obtained with the polyamide membranes. Therefore, it appears that the polyamide membranes should perform better for the desalination of the leachate than the cellulose acetate membranes.

16. DEMONSTRATION OF TUBULAR AND SPIRAL WRAP REVERSE OSMOSIS ON PILOT SCALE AT THE BISASAR ROAD WASTE DISPOSAL SITE FOR THE TREATMENT OF MUNICIPAL SOLID WASTE LEACHATE

16.1 Tubular Cellulose Acetate Membranes

It appears that it should be possible to control membrane fouling with regular acid (phosphoric) and chemical (EDTA and SLS and/or STPP and EDTA) cleaning. The CWF and the permeate fluxes remained at approximately 500 and 300 ℓ/m^2.d, respectively, after about 500 hours of operation (feed-and-bleed).

The permeate flux after approximately 500 hours of operation (batch tests) was somewhat lower than the initial permeate flux (batch test). The initial permeate flux started at about 600 ℓ/m^2.d and was about 350 ℓ/m^2.d at 75% water recovery. The permeate flux after 500 hours of operation started at about 480 ℓ/m^2.d and was about 350 ℓ/m^2.d at 70% recovery. Permeate flux was about 20% lower after 500 hours of operation. This, however, can be expected to occur in the last module of an RO treatment train, as was simulated with the feed-and-bleed tests, where membrane fouling is more likely to occur as a result of higher brine concentration.

The TDS and conductivity removals were only 77,1 and 64,2%, respectively, after 500 hours of operation (batch test). The TDS and conductivity removals were 96,1 and 93,2%, respectively, on a fresh membrane surface (batch test). Therefore, a significant reduction in salinity removal has occurred as a result of membrane fouling.

16.2 Tubular Polyamide Membranes

It appears that it should be possible to control membrane fouling with regular acid (hydrochloric) and chemical (Ultrasil 10) cleaning. The CWF and the permeate fluxes remained at approximately 500 and 200 ℓ/m^2.d, respectively, after about 500 hours of operation (feed-and-bleed).

The permeate flux after approximately 500 hours of operation (batch test) was also somewhat lower than the initial permeate flux (batch test). The initial permeate flux started at about 1 250 ℓ/m^2.d and was about 600 ℓ/m^2.d at approximately 70% water recovery. The permeate flux after 500 hours of operation started at 1 000 ℓ/m^2.d and was about 400 ℓ/m^2.d at 70% water recovery. Therefore, permeate flux was also about 20% lower after 500 hours of operation as was the case with the cellulose acetate membranes.

The TDS and conductivity removals were 96,6 and 93,4%, respectively, after 500 hours of operation (batch test). The TDS and conductivity removals were 97,9% and 96,9%, respectively, on a fresh membrane surface (batch test). Therefore, the reduction in salinity removal after 500 hours of operation was significantly less than was the case with the cellulose acetate membranes.

16.3 Spiral Wrap Membranes

Membrane fouling was experienced during treatment of the Bisasar Road MSWL with the spiral membrane system. Permeate flux started at 838 ℓ/m^2.d (CWF 989 ℓ/m^2.d) and was 711 ℓ/m^2.d when the run was terminated after approximately 32 hours of operation (CWF 893 ℓ/m^2.d). Membrane fouling occurred. However, it was possible to control membrane fouling with chemical cleaning. Preservation of the membranes in SMBS solution increased the CWF further to 1 011 ℓ/m^2.d. Therefore, the RO tests over an approximately 32 hour

period have shown that it should be possible to control membrane fouling with regular chemical cleaning of the membranes. However, longer term tests should be conducted to develop a proper membrane cleaning strategy when using spiral wrap membranes for the treatment of leachate

An excellent quality product water could be produced. TDS was reduced from 4 982 mg/ℓ in the RO feed to only 45 mg/ℓ in the RO product (99,1% removal). Conductivity was removed from 11,3 mS/cm to 0,49 mS/cm in the RO product (95,7% removal). Almost 100% of the COD (1 400 mg/ℓ) was removed. Ammonia-nitrogen was removed from 589 mg/ℓ in the RO feed to 47 mg/ℓ (92,0% removal) in the RO permeate. Chloride was removed from 1 535 mg/ℓ to 14,0 mg/ℓ (99,1% removal). Therefore, the RO product complies to the discharge quality requirements to the water environment (except for ammonia-nitrogen) and to the sewer system.

16.4 Economics

The estimated capital and operational costs to treat 250 kℓ/d of MSWL from Bisasar Road with different RO plants are as follows : -

	Capital (MR)		Operational (R/kℓ)
Tubular RO (cellulose acetate)	1,95		11,45
Tubular RO (polyamide)	8,1[(1)]	6,5[(2)]	16,24
Spiral RO	0,56		3,51
Disc tube RO	6,2		26,65

[(1)]Total plant from overseas supplier
[(2)]Only membranes, modules, manifolds, from overseas supplier.

Overseas RO plants are significantly more expensive than local RO plants as a result of the exchange rate. Therefore, good business opportunities exist for local suppliers of RO plants in South Africa and elsewhere.

17. ODOUR CONTROL

Technologies are available that can be used for odour control at landfill sites. Some of these technologies (spraying of counteractants) have been successfully applied in South Africa. Consequently, these technologies should be applied where appropriate.

18. GUIDELINES FOR LEACHATE CONTROL

Guidelines for the control of leachate in South Africa are available in the second edition (1998) of the Waste Management Series known as the ‘Minimum Requirements’.

Leachate must undergo some type of treatment and disposal after it has been removed from the landfill. The most common methods of managing leachate are :

- Discharge to a sewage treatment works (physical, chemical, biological treatment)
- On-site treatment followed by discharge (evaporation / crystallisation / biological / activated carbon; electrodialysis; reverse osmosis; oxidation; reed beds, etc.), and
- Recirculation back into the landfill.

19. RECOMMENDATIONS

It was demonstrated in this study on a difficult to treat industrial solid waste leachate that ED should be a suitable technology for the desalination / concentration of high TDS and high organic containing leachates prior to disposal to the sewer system. This is a new process application of the ED process. However, it is expected that the process should perform better on lower TDS industrial solid waste leachates (20 000 to 50 000 mg/ℓ) as has already been shown. The developed process, however, should be further optimised in terms of : -

(a) Membrane lifetime
(b) Removal of organics from the ED product with a membrane reactor
(c) Further brine volume reduction by adding less water to the brine during treatment or by treatment of the brine with an evaporator or membrane distillation, followed by crystallistion
(d) Current density to reduce membrane area
(e) Economics of the process.

Optimisation studies should be conducted by implementation of the ED process on small-scale (pilot scale) for the treatment of a hazardous leachate. It is suggested that such a study be conducted at the Aloes landfill site in Port Elizabeth or another similar hazardous leachate. The information obtained from such a study should form the basis for the application of the ED process for the treatment of other similar industrial solid waste leachates and high brine concentration industrial effluents.

The anionic ED membranes used in the investigation showed excellent membrane fouling characteristics. However, other anionic membranes are also available which might have improved characteristics for the treatment of industrial solid waste leachates in terms of membrane lifetime and these membranes should also be evaluated for the treatment of industrial solid waste leachates.

The desalinated ED product water may contain recalcitrant organic compounds. However, it should be possible to further degrade these high molecular mass compounds in the ED product in a membrane reactor using ultrafiltration or nanofiltration membranes. It is recommended that this process should be further investigated because the process should be able

to degrade organics that otherwise could not have been degraded in a conventional biological system.

The ED brine can be put back into the landfill or it can be further concentrated with conventional evaporation technologies. Alternatively, the ED brine can be concentrated with membrane distillation followed by crystallisation. This is a new approach to brine treatment and it is recommended that this new approach be studied further.

It was demonstrated in this study that tubular RO should be a suitable technology for the desalination / concentration of a municipal solid waste leachates. Process design criteria for a full-scale application have been derived from pilot tests. Either polyamide or cellulose acetate RO membranes should be suitable for this application, depending on the required quality of the treated leachate. Therefore, it is recommended that this technology should now be implemented in South Africa so that the necessary confidence could be obtained to apply this technology for water pollution control and leachate volume reduction.

It was shown in this study that the treatment of RO brine with evaporation / drying technology is an attractive technology to reduce brine volume and to increase water recovery. This is an established technology in Germany and the cost of brine treatment with this process should be determined with the objective of possibly applying this technology in South Africa.

Preliminary test work with a spiral wrap RO system for the treatment of a municipal solid waste leachate has shown that this RO system with its magnetic field around the membrane module for scale / fouling control, shows promise for the treatment of the leachate. Pretreatment of the leachate only consisted out of sand and cartridge filtration. Preliminary indications are that membrane fouling should be controlled with chemical cleaning. However, a longer term study will be required to confirm it. It is also not clear what the effect of the magnetic field around the membrane is on the controlling of membrane fouling. Therefore, it is recommended that a longer-term pilot study should be conducted with this RO system so that the necessary information could be obtained. This is a new application of this RO process for the treatment of municipal solid waste leachate. Demonstration of the control of membrane fouling with this process can open up a new field for the application of spiral wrap membranes for leachate treatment.

Biological treatment of municipal solid waste leachate may be necessary prior to RO desalination in certain cases. However, very little knowledge is available in South Africa regarding membrane bioreactors for the treatment of municipal solid waste leachate. Difficult degradable organics should be more readily biodegradable with a membrane reactor. Therefore, it is recommended that laboratory or pilot-scale studies be conducted so that the necessary experience and process know-how can be developed.

The evaporation / crystallisation / biological / activated carbon technology that has been piloted by Enviroserv for the treatment of the Holfontein-type leachate with its very high TDS and organic concentrations, appear to be suitable for the treatment of this leachate. It is therefore recommended that the performance of a full-scale application of this technology be closely monitored so that the process design criteria used for the process could be implemented for other similar industrial solid waste leachates.

ACKNOWLEDGEMENTS

The research in this report emanated from a project funded by the Water Research Commission entitled : 'Treatment of Landfill Leachate from Hazardous and Municipal Solid Waste'.

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

Dr G Offringa	Water Reseach Commission (Chair person)
Mr D J Marais	Water Reseach Commission (Secretary)
Dr D A Baldwin	Environmental & Chemical Consultants
Mr R A Jones	Keyplan (Pty) Ltd.
Ms E Gombault	Enviroserv Holdings
Ms S Alcock	Enviroserv Holdings
Ms G Winter	Ekurhuleni Metropolitan Council
Mr T Aab	DWAF, Water Quality Management
Ms A Maurizi	Springs City Council
Mr D Turner	Durban Metro Council
Mr L J Strachan	Durban Metro Council

The financing of the project by the Water Research Commission and Enviroserv Holdings and the contribution of the members of the Steering Committee are gratefully acknowledged.

GLOSSARY OF ABBREVIATIONS

AA	Acetic Acid
Alum	Aluminium Sulphate
AOX	Adsorbed Organic Halogens
ATP	Adenine Triphosphate
BOD	Biological Oxygen Demand
BTEX	Benzene Toluene Ethyl Xylenes
CC	Calcium Carbonate
CF	Concentration Factor
C_{FV}	Volumetric Concentration Factor
COD	Chemical Oxygen Demand
cp	cell pair
CWF	Clean Water Flux
DM	Dutch Mark
DMA	Durban Metropolitan Area
DT	Disc Tube
DWAF	Department of Water Affairs & Forestry
EC	Electrical Conductivity
ED	Electrodialysis
EDTA	Ethylenediaminetetra acetic acid
EPA	Environmental Protection Agency
GDACE	Gauteng Department of Agriculture and Environment
ge	Gram Equivalents
h	hour
HPRO	High Pressure RO
ISW	Industrial Solid Waste
ISWL	Industrial Solid Waste Leachate
kD	Kilo Dalton
kℓ	kilolitre
LC_0	No Effect Concentration
LC_{10}	Minimum Effect Concentration
LC_{50}	Concentration at which 50% of the organisms die
m^3	cubic metre
MLSS	Mixed Liquor Suspended Solids
MSW	Municipal Solid Waste
MSWL	Municipal Solid Waste Leachate
NF	Nanofiltration
NH_3-N	Ammonia-Nitrogen
PAC	Poly-Aluminium Chloride
PLC	Process Logic Controller

RO	Reverse Osmosis
rpm	Revolutions per minute
s	Seconds
SBR	Sequential Batch Reactor
SLS	Sodium Laurel Sulphate
SMBS	Sodium Meta Bisulfite
STPP	Sodium Tripoly Phosphate
TCLP	Toxicity Characteristic Leaching Procedure
TDS	Total Dissolved Solids
TKN	Total Kjeldahl Nitrogen
tfc	thin film composite
TOC	Total Organic Carbon
TOX	Total Organic Halogens
UF	Ultrafiltration
USEPA	United States Environmental Protection Agency
Δt	Membrane Permselectivity

TABLE OF CONTENTS

---oOo---

LIST OF TABLES

LIST OF FIGURES

1. INTRODUCTION AND AIMS

1.1 General Background

Continuing industrial and commercial growth in many countries around the world in the past decades have been accompanied by rapid increases in Municipal Solid Waste (MSW) and Industrial Solid Waste (ISW) generation (Lin and Chang, 2000). At present, landfilling is the most popular way of solid waste disposal and landfilling will continue to be the primary means of MSW and ISW disposal in future. Besides scarcity of available landfill sites in certain regions, a large amount of leachate (originating from water which has percolated through emplaced refuse) generated from a landfill site poses a major problem of landfill disposal of MSW and ISW. Proper treatment of the leachate has therefore been a challenging problem confronting the local authorities.

Landfill leachate has been generally known as a high-strength waste water that is most difficult to deal with. This is due primarily to its large variability of organic and inorganic and heavy metal contents, strongly depending on the age and type of solid waste of a landfill site (Knox and Jones, 1979). Satisfactory treatment of leachate is thus no easy task. The most popular treatment of landfill leachate in the past was anaerobic digestion or the aerobic activated sludge method (Lema *et al.,* 1988). These methods were known to be inadequate in handling such a difficult treatment task. In the more recent decades, search for alternative treatment methods had focussed on various sophisticated technologies. These included advanced biological, chemical and physical treatment methods. Ying *et al.* (1986), Robinson (2000) and Strachan (2000) studied leachate treatment using the sequencing batch reactor (SBR) method. Percival *et al.* (1997) studied aerobic treatment of a high-strength leachate preceded by ion-exchange and lime addition to effect inorganic removal prior to biological treatment. Imai *et al.* (1993, 1995, 1998) developed an efficient biological activated carbon fluidized bed process. Chemical oxidation using strong oxidising agents, such as Fenton's reagent, photo-assisted H_2O_2, ozone or UV-Vis light, was reported by Gau and Chang (1996), Kim *et al.* (1997), Steensen (1997) and Ince (1998). Membrane processes have received considerable attention in the past decades, as reported by many investigators (Chian and de Walle, 1976; Bjorkman and Wechsler, 1983; Keenen *et al,* 1983; Krug and McDougal, 1989; Kingman and Nutini; 1991; Bilstad and Madland, 1992; Pirbazari *et al.,* 1996; Chianese *et al.* 1999, Holz *et al*., 1993; Rautenbach and Linn, 1996; Peters, 1998). In fact, Gierlich and Kolbach (1998) indicated that membrane technology has been widely practiced in many European countries in dealing with leachate treatment. In two recent investigations, Chiang *et al* (1995) and Cossu *et al* (1998) employed an electrochemical method in decomposing the refractory organic and inorganic components in the leachate. Both investigations found efficient removal of the chemical oxygen demand (COD) and ammonia concentration in the leachate.

The above brief review of the literature has indicated that physical (membrane) and chemical methods appear to offer quite good alternatives to biological treatment. Reverse osmosis should be a very suitable technology for the treatment of hazardous leachates with a

relatively low TDS concentration. However, the salinity concentration of certain industrial leachates may be too high to consider RO for treatment. Consequently, evaporation and electrodialysis technologies are the remaining technologies that should be considered for the treatment of high TDS concentration leachates. Consequently, chemical methods and membrane technology (RO for low TDS leachate and ED for high TDS leachate) were selected to study the treatment of two selected leachates in South Africa.

1.2 South African Background

The minimum requirements for waste disposal to landfill sites according to the Department of Water Affairs and Forestry (DWAF) is that all hazardous waste sites should have a leachate management system. The minimum requirements for the classification, handling and disposal of hazardous waste according to DWAF states that all leachates are hazardous, and DWAF is beginning to push waste disposal companies to manage leachate treatment effectively. Therefore, suitable technologies will be required for the successful treatment of these hazardous leachates.

Two types of hazardous municipal leachates are produced in landfill sites in South Africa. The one is a high TDS (50 to 100 g/ℓ), high organic (10 to 80 g/ℓ COD) concentration leachate containing hazardous chemicals like phenols, sulphides, ammonia-nitrogen, chromium, etc. (Industrial Solid Waste Leachate, ISWL). This ISWL is unique to South Africa, with its relatively low rainfall compared with Europe, and past waste disposal practices. The other leachate is a low TDS (2 to 6 g/ℓ), low organic (2 to 6 g/ℓ COD) concentration leachate (Municipal Solid Waste Leachate, MSWL). The ISWL cannot be discharged into the municipal sewer system, because the high salinity levels, as well as the other hazardous chemicals (phenols, chromium, heavy metals) will adversely affect the biological treatment process. However, the MSWL can, in some cases, be discharged directly into the municipal sewer system without any problems, depending on the dilution water available. Municipal solid waste leachate cannot be discharged into the municipal sewer system where little dilution water is available, and where hazardous chemicals are present in the leachate. Positive water balances at landfill sites can also result in pollution of the water environment. Therefore, the leachates should be treated prior to disposal, to prevent adverse effects on municipal biological treatment processes, and to prevent pollution of the water environment.

Industrial solid waste leachate (50 to 100 g/ℓ TDS) cannot be treated with reverse osmosis (RO) for salinity removal, because the salinity levels are too high. Electrodialysis (ED), however, which can handle high salinity levels, has the potential to desalinate this type of leachate effectively. However, membrane fouling may be a problem. Pretreatment of the ISWL with adsorbents or absorbents (ash) and other physical/chemical (coagulation-flocculation, soda ash, etc.) and membrane methods is, therefore, considered to be important for the removal of most of the membrane foulants prior to ED desalination. Electrodialysis membranes are also available, which are claimed to be more fouling-resistant than conventional electrodialysis membranes, and these membranes were used in the study.

Preliminary work on an ISWL has shown that approximately 50 to 60% of the salinity could easily be removed with electrodialysis desalination (Schoeman and Steyn, 1999). However, it was not possible to remove the remaining salinity, consisting of organic complexes, with electrodialysis and membrane fouling was also experienced. Reverse osmosis, however, has the potential to desalinate the electrodialysis diluate further to very low concentration levels.

It should be possible to desalinate MSWL effectively with RO for direct discharge into the water environment. This will help to prevent water pollution effectively. However, very little information is available in South Africa of the performance of RO for the desalination of MSWL. The newly developed RO system from Grahamtek Systems, which is claimed to have the ability to treat high fouling effluents effectively without chemical pretreatment, was used in the study. Tubular cellulose acetate RO membranes, which are manufactured in South Africa, were also evaluated in the study, together with tubular polyamide RO membranes, which are applied overseas for the treatment of MSWL. The tubular cellulose acetate RO membranes are claimed to be more resistant to membrane fouling than polyamide tubular and conventional spiral wrap RO membranes.

There are approximately eight strong leachate-producing landfill sites in South Africa and many more weak leachate-producing sites. Several industries are also producing strong effluents. Therefore, the process technology that would be developed in this study could find wide application in South Africa.

1.3 Aims

The main aim of this project was to develop ED and RO process technology for the treatment of hazardous industrial solid waste leachate (ISWL) with high TDS and high organic concentrations (Holfontein waste disposal site, Springs). The secondary aim of this project was to develop RO process technology for the treatment of hazardous municipal solid waste leachate (MSWL) with low TDS and low organic concentrations (Bisasar Road waste disposal site, Durban). The different tasks include : -

(a) Concise literature overview.

(b) Characterisation of the leachates.

(c) Determination of the biodegradability of the leachates.

(d) Evaluation of pretreatment of the leachates for the removal of suspended material, organics and inorganics.

(e) Determination of the fouling potential of the leachates for ED and RO membranes.

(f) Evaluation of membrane cleaning strategies of fouled ED and RO membranes with commercially available membrane cleaning agents.

(g) Evaluation of ED and RO performance for the desalination/ concentration of the leachates.

(h) Determination of the preliminary economics of the processes.

(i) Technology transfer through ED and RO pilot tests on site, and determination of the final economics of the processes.

(j) Development of the capacity of Technikon students to operate and maintain laboratory and membrane pilot plants.

2 LITERATURE OVERVIEW

2.1 Introduction

The state of the art of the treatment of landfill leachate will be presented followed by process combinations and operating results (Holz *et al.*, 1993). The purification of landfill leachate with disc tube RO and nanofiltration will then be discussed (Peters, 1998). The handling of the concentrate will be addressed as well as process improvement using nanofiltration.

Hybrid processes involving membranes for the treatment of high organic / inorganic concentration contaminated waste water will also be discussed (Rautenbach and Mellis, 1995). The hybrid process consists out of a bioreactor, ultrafiltration and nanofiltration.

High-pressure reverse osmosis and nanofiltration, a 'zero-discharge' process combination for the treatment of wastewater with severe fouling/scaling potential will then be presented (Rautenbach and Linn (1966). Waste water treatment by membrane processes - new developments in ultrafiltration, nanofiltration and reverse osmosis will be discussed (Rautenbach *et al.*, 1996).

Treatment of RO concentrate will be addressed (three different possibilities will be discussed) (Biehler and Hägele, 1995).

A literature study regarding leachate treatment focussing on high COD values has been conducted by Enviroserv (Thomas *et al*., 1999). Advantages and disadvantages of different technologies will be presented and recommendations will be made.

The bio-enhancement treatment of a hazardous leachate will be presented and it will be shown that a leachate with high COD and high TDS, containing high concentrations of phenolics, can be treated successfully (Ambachtsheer, 2000).

Appropriate biological treatment of landfill leachate with full nitrification and denitrification will be presented (Strachan *et al.*, 2000). Different treatment options for the treatment of the Holfontein leachate will also be presented (Enviroserv, 2001).

2.2 Treatment of Landfill Leachate by Reverse Osmosis

The state of the art of the treatment of landfill leachate is described by Holz *et al.*, 1993.

2.2.1 Introduction

The state of the art for the treatment of landfill leachate is represented by the following process combinations: -

(a) Biological pretreatment including nitrification and denitrification; chemical/physical treatment with flocculants and activated carbon; residue treatment by sludge dewatering as hazardous waste and reactivation of activated carbon.

(b) 2-stage RO process (first stage tubular modules, second stage spiral wound modules); residue treatment by concentrate evaporation, drying and disposal as hazardous waste

(c) biological pretreatment including nitrification and denitrification; RO process (tubular modules); residue treatment by concentrate evaporation, drying and disposal as hazardous waste

(d) biological pretreatment including nitrification and denitrification; chemical oxidation using ozone and/or hydrogen peroxide together with ultraviolet radiation.

2.2.2 Qualities and quantities of leachate

Leachate from sanitary landfills can be divided into four categories : -

(a) leachate from the acidic phase (acidic leachate);
(b) leachate from the methanogenic phase (stabilised leachate)
(c) diluted leachate; and
(d) leachate from the disposal of residues.

Typical composition of different kinds of leachates are as follows : -

Table 2.1 : Average concentrations and quantities of different kinds of leachates

Kind of Leachate	COD mg/ℓ	NH_4-N mg/ℓ	AOX mg/ℓ	Cl mg/ℓ	Conductivity mS/cm	Quantity m^3/(ha.d)
Acidic phase	>15 000	12	2 - 5	3 000	25	2 - 5
Methanogenic phase	<5 000	18	4	3 000	18	5
Diluted leachate	15	500	12	1 000	10	7,5
Future leachate	15	700	2	1 000	15	5

2.2.3 The RO process

The following basic data should be carefully evaluated for the layout of a full-scale RO plant:

(a) leachate quality;
(b) leachate quality influenced by the manner of primary purification;
(c) possible concentration factor to keep the required quality of permeate;
(d) kind of modules, membrane material and cleaning procedures;

(e) pressure, temperature and velocity influencing the permeate flux and the permeate quality;
(f) treatment of the concentrate.

In general, physical pretreatment by sand or drum filters is recommended to preserve pumps, membranes and measuring instruments. The permeate quality is determined by minimum requirements or even lower values. The quality of the permeate can be controlled by the volumetric concentration factor C_{FV}, which is the leachate flow rate divided by the concentrate flow rate. For high concentrations of process limiting compounds like calcium and iron or of critical parameters like NH_4-N and AOX, the physical and economic range of the RO process may be exceeded. The parameter C_{FV} is very important, because it determines the quantity of concentrate produced for a given quantity of leachate. A C_{FV} of 5 may be sufficient for the process to be economic, provided the concentrate can be recirculated back to the landfill site. If the concentrate is to be evaporated, the C_{FV} must be greater than 5 to minimize concentrate treatment costs.

There is a strong interrelation between other basic process data. Modules and membranes control the permeate quality. The lifetime of the membranes and the possibility of cleaning them are very important in leachate treatment. Due to the possible damage of the membrane surface by circulating sponge balls, chemical cleaning was found to be more suitable. The size of the plant, the operating pressure, the temperature and the recirculation rates determine the quality of the permeate and the plants power consumption.

Basic process data and the suitability of the modules and membranes have to be evaluated during tests. These tests should also show if biological pretreatment is necessary or if the purification of raw leachate is a sufficient and economic method.

2.2.4 Operating results

(a) **Breinermoor landfill**

The process combination at the Breinermoor leachate treatment plant consists of biological pretreatment with nitrification and denitrification plus a 1-stage RO with tubular membranes. The daily permeate production at Breinermoor is approximately 120 - 150 m^3.

During the starting phase at the end of 1991 it was discovered that the cleaning procedures had to be adjusted to match the specific leachate conditions. Deposits had gathered on the membranes and could not be chemically removed. With a simple sponge-ball cleaning, however, the original productivity of the tubular membranes (flow/retention) could be restored. Since then the plant has been in operation without any major difficulties. The first set of membranes did not have to be exchanged until the date of the publication.

The plant is run at a concentration factor of 5 to 7 and pressures of 40 to 50 bar. Its operation is fully automized and computer controlled. This also applies to the regular cleaning procedures.

Biological pretreatment and reverse osmosis, come fully up to expectations (Table 2.2). The permeate values fall much below the limiting values so that even with an increased raw leachate concentration there will be no risk of exceeding the limits.

Table 2.2: Chemical composition of raw leachate and leachate discharged from biological pretreatment at the Breinermoor landfill plant.

Parameter		Raw Leachate (Inflow)	Biological Pretreatment (discharge)	Permeate (discharge RO)	Limiting Values	
					Local Requirements	Future Legal Requirements
COD	mg/ℓ	520 - 800	480 - 600	<15	200	200
BOD_5	mg/ℓ	20 - 50	<15	not detectable	20	20
TKN	mg/ℓ	160 - 360	35 - 80	15 - 25	—	—
NH_4-N	mg/ℓ	80 - 250	0,1 - 40	0,015 - 1,4	50	10
NO_3-N	mg/ℓ	5 - 150	190 - 350	6 - 40	—	—
NO_2-N	mg/ℓ	0,04 - 1,0	0,1 - 2,5	0,015 - 0,7	—	2
Tot. N_{min}	mg/ℓ	80 - 250	190 - 350	5 - 40	95	70
Cl^-	mg/ℓ	780 - 820	590 -760	10 - 20	—	—
AOX	µg/ℓ	400 - 650	400 - 650	<10	500	500
pH	(-)	6,9 - 7,8	6,5 - 7,7	4,5 - 6,3	6 - 8,5	—
Conduct	mS/m	4,5 - 7,0	5,0 - 5,3	0,14 - 0,4	—	—
Fish tox.	(-)	16	2,0 - 4,0	1 - 2	2	2

A 3 000 m^3 leachate storage tank guarantees for an undisturbed operation of the plant. The vitreous enamelled steel tank serves as a buffer for varying quantities and concentrations of landfill leachate. In case of plant malfunctions a 20-day quantity of leachate may be collected in this tank.

For the further treatment of the RO concentrate an evaporation and drying facility will be installed. Until its completion the concentrate will be taken back to the bottom-sealed landfill site. This arrangement is limited to a period of five years.

(b) **Bassum land fill site**

A mobile container plant is being operated on the Bassum landfill site on a rental basis. The plant is fully automized and has a capacity of 60 m^3 per day. It is supervised by only one operator. The container is equipped with a 2-stage RO with tubular and spiral wound modules. The inflow and discharge values obtained at a concentration factor of approximately 4,5 to 5 is shown in Table 2.3.

(c) **Lampertheim leachate treatment plant**

The Lampertheim leachate is pretreated (denitrification / nitrification) in a single-basin unit and cleaned with a 3-stage RO (200 m^3/d). The discharge values are listed in Table 2.4.

The RO concentrate is evaporated in a 4-stage evaporation facility and dried in a fluid bed granulator to obtain dust-free granulates suitable for final storage. For final storage the granulates are automatically filled into standard drums and taken for underground disposal.

Table 2.3: Raw leachate inflow values and permeate discharge values at the Bassum RO plant

Parameter		Raw Leachate	Permeate II
pH		7,56 - 8,68	3,72 - 4,63
Conductivity	µS/cm	7 840 - 17 130	15 - 33
COD	mg/ℓ	1 805 - 3 926	<15
BOD_5	mg/ℓ	96 - 1 413	0,9 - 2,4
AOX	µg/ℓ	835 - 2 150	<10 - 47
Total-N	mg/ℓ	810 - 1 729	1,6 - 3,8
Ammonium-N	mg/ℓ	470 - 1 066	0,68 - 1,3
Chloride	mg/ℓ	1 200 - 2 203	<1 - 1,8
Fe II	mg/ℓ	7 - 34	<0,005 - 0,05
Fe (total)	mg/ℓ	6,3 - 36	<0,01 - 0,07
Magnesium	mg/ℓ	70 - 210	0,02 - <0,1
Calcium	mg/ℓ	82 - 190	0,11 - <0,5
Lead	mg/ℓ	0,013 - 0,098	<0,001 - 0,02
Chromium	mg/ℓ	0,16 - 0,6	<0,001 - 0,17
Copper	mg/ℓ	0,002 - 0,079	<0,001 - 0,01
Zinc	mg/ℓ	<0,01 - 0,41	<0,01 - 0,42
Cadmium	mg/ℓ	0,0008 - 0,0071	<0,0002
Mercury	mg/ℓ	<0,0002 - 0,0003	<0,0002
Nickel	mg/ℓ	0,11 - 0,27	<0,001 - 0,011

Table 2.4: Raw leachate inflow values and permeate discharge values at the Lampertheim reverse osmosis plant.

Parameter		Raw Leachate (inflow)	Permeate 1	Permeate II	Permeate III (discharge)
pH		7,21 - 8,24	4,09 - 6,51	3,21 - 5,91	3,5 - 6,29
Conduct.	µS/cm	5 950 - 11 900	220 - 520	20 - 90	9 - 40
COD	mg/ℓ	913 - 4 212	<15 - 64	<15	<15
BOD_5	mg/ℓ	76 - 518	<2	<2	<2
NH_4-N	mg/ℓ	176 - 553	5,7 - 26,7	0,9 - 5,4	0,08 - 1,35
NO_3-N	mg/ℓ	29,1 - 120	<0,5 - 11,0	<0,5 - 4,1	<0,5 - 2
AOX	mg/ℓ	1,3 - 3,3	not measured	not measured	0,001 - 0,002
Cd	µg/ℓ	<0,1	<0,1	<0,1	<0,01
Cr	µg/ℓ	<10	<10	<10	<10
Cu	µg/ℓ	<10	<10	<10	<10
Hg	µg/ℓ	<0,1	<0.1	<0,1	<0,1
Nl	µg/ℓ	<10	<10	<10	<10
Pb	µg/ℓ	<10	<10	<10	<10
Zn	µg/ℓ	4 623 - 7 967	<10	<10	<10

2.2.5 Suitable process combination for small and medium-sized leachate quantities

Further treatment of RO-concentrate with a comparably expensive evaporation facility ahead of the drying stage is only recommended for the treatment of substantial quantities of leachate. For small and medium-sized leachate quantities it is more economical to send the concentrate straight to a fluidized bed granulator. This especially applies if sufficient thermal energy is available from landfill gas utilisation (waste energy from thermal power stations). Energy savings such as those achieved with multi-staged evaporation facilities would then be of secondary importance.

The ideal process combination for the treatment of small and medium-sized leachate quantities is biological pretreatment and removal of solids by ultrafiltration. Sludge separation by ultrafiltration allows to obtain a high content of solids in the biological reactor. This makes the installation of small volume reactors possible.

The process combination for the treatment of small and medium-sized leachate quantities is biological pretreatment/reverse osmosis/fluidized bed granulator.

Specific costs of approximately 70 DM/m^3 (capital costs, operating costs, personnel costs) are estimated for a leachate treatment plant with undiluted leachate quantities of approximately 100 m^3/d and the following specifications :

- COD approximately 2 500 mg/ℓ
- NH_4-N approximately 1 500 mg/ℓ
- AOX approximately 3 mg/ℓ

2.2.6 Summary

Reverse osmosis plants in operation have proven the suitability of this process for the treatment of landfill leachate and similar waste waters. This, however, does not only apply to the unsurpassed discharge qualities, but also to the high availability of the installations (in general >90%).

Reverse osmosis plants may be operated with or without biological pretreatment. In general a multi-staged RO is required if a biological pretreatment is not installed. With biological pretreatment a 1-stage RO is in most cases sufficient.

With the new drying technique of evaporation and fluidized bed granulation a solution has finally been found to the much discussed issue of how to properly treat RO-concentrate.

2.3 Purification of Landfill Leachate with Reverse Osmosis and Nanofiltration

2.3.1 Components dissolved in leachate

The purification of landfill leachate with RO and NF has been described by Peters (1998). An evaluation of the data of leachate from more than 150 landfills in Germany, and from a few in Spain, shows that the amount of components dissolved in leachate from different kinds of landfills covers the range from 2 to 15 g/ℓ. From this the fraction of organic components is considered to cover a range between 0,1 and 3 g/ℓ. That is much smaller than the inorganic part, indicated to have a range from 1,6 to 14,3 g/ℓ, including ammonia with values between 0,3 to 2 g/ℓ. Thus, between 80 and 95% of the components dissolved in landfill leachate correspond to inorganic material, and only between 5 and 20% are of organic origin.

Since such contaminants are most not appropriate for treatment by conventional biological processes, new regulations tend to limit the discharge of such complex waste to municipal sewers. Even by combining biological treatment with adsorption by active carbon or with the oxidation of part of the dissolved organic material using ozone or other oxidizing agents, only partial destruction of contaminants will be achieved. It will not reach the purification needed to fully reduce the negative impact of landfill leachate on the environment (Peters, 1966). One aspect is the so-called ‘hard COD’ that is not biodegradable and cannot be destroyed or adsorbed and thus will remain in the water discharged after being treated with the mentioned processes, causing problems in the future as a consequence of the effects of long time accumulation.

Therefore, more effective methods of treatment of this material should be developed. The use of RO either as a main step in a landfill leachate treatment chain or as a single step has shown to be a very successful means of achieving full purification.

2.3.2 Reverse osmosis for the purification of landfill leachate

Due to the ability of modern high-rejection reverse osmosis membranes to retain both organic and inorganic contaminants dissolved in water at rejection rates of 98-99%, RO is also useful for purifying liquid waste such as landfill leachate and for helping to solve the growing problem of water pollution. The permeate contains only very low levels of inorganic and organic contaminants. These meet potable water standards and discharge of this water to the next river or aquifer contributes to maintenance of the natural equilibrium as this leachate was originally mainly clean rain water.

As the RO membrane is operating like a well defined barrier, the purification process itself can be controlled continuously and with a high degree of security by simple and precise measurement of the electric conductivity.

Because of the high rejection rate for each kind of contaminant dissolved in the feed water a high flexibility against changes of the concentration of the compounds in landfill leachate is given. Therefore, the permeate produced has always the expected high quality, as this is based on a reproducible high purification efficiency.

However, in addition to requiring highly resistant membranes, the treatment of landfill leachate with RO demands the use of open channel module systems that can be cleaned with high efficiency with regard to scaling, fouling and especially biofouling. Therefore, tubular modules were the first medium used in the early RO system for the purification of landfill leachate starting in 1984. An alternative was introduced to this market in 1988. The disc-tube-module (DT module) has been installed since then with great success (Peters, 1995). By the year 1997, plants equipped with this DT-module represented more than 80% of the total capacity installed for the purification of landfill leachate by RO.

The successful operation of RO in the plant of the municipal waste landfill of Ihlenberg (former VEB Deponie Schönberg) near the city of Lübeck in Germany, the most modern and largest multi-stage plant that has been realized up to this time for landfill leachate purification, demonstrates the possibilities of modern membrane technology. Some results are given as example in Table 2.5.

This reverse osmosis system with a capacity of 36 m^3/h has been in operation without any problem since the 15th of December of 1989 with one change of membranes until the time of the publication.

With two reverse osmosis stages, the average rejection rates for salts and organic contaminants are about 99%. Depending on the salt content of the feed water and the operation time between the cleaning cycles, the operating pressure ranges between 36 and 60 bar at ambient temperature. The specific permeate flux was calculated to be approximately 15 ℓ/m^2.h.

Table 2.5 : Typical performance in leachate purification

Parameter	Leachate	Permeate 1	Permeate 2	Rejection %
pH value	7,7	6,8	6,6	
El. conduct, µS/cm	17 250	382	20	99,9
COD, mg/ℓ O_2/ℓ	1 797	15	<15	>99,2
Ammonium, mg/ℓ	366	9,8	0,66	99,9
Chloride, mg/ℓ	2 830	48,4	1,9	99,9
Sodium, mg/ℓ	4 180	55,9	2,5	99,9
Heavy metals, mg/ℓ	0,25	<0,005	<0,005	>98

Similar results are reported from other plants, for example for the landfill Kolenfeld near Hannover, with start up in February 1990. This plant, operated at 1,8 m^3/h, has an availability of over 90%. The electric conductivity of the feed is reported to be 15 000 to 16 000 µS/cm, the rejection rate always more than 98%, for COD 99%. New membranes

were installed in April 1993, after more than 3 years in operation because of a decreasing permeate flux.

These data from long terms experience have also been confirmed by the results of the other more than 120 systems that are in operation on different landfills up to now and by the data collected during numerous tests with pilot plants on technical scale all over Europe, North America and in some countries in the Far East.

2.3.3 Handling of leachate concentrate

The purification of landfill leachate helps to avoid further contamination of the resources like groundwater and surface water. But beside the ecological aspect - the minimization of the burden on the environment - also has to be taken into consideration the commercial feasibility, i.e. the affordability. In this regard membrane filtration has proved to be a justifiable and economic solution in most of the cases, even when the overall costs for the purification are compared with other approaches for the treatment of landfill leachate.

This evaluation includes the handling of the concentrate produced in the RO plants with 75 to 80% recovery rate, that in the past has been considered to be a very cost intensive step. That applies in fact for few plants in operation, that are using evaporation and drying followed by deposition of the dry residues in a special landfill. However, today other possibilities are gaining importance. They have been developed considering the best technology available and at the same time ecological and economical requirements. These are : -

(a) transport of the concentrate to an incineration plant equipped for the burning of liquid hazardous waste;
(b) the solidification of the concentrate with different materials, like fly ash (Kenner and Peters, 1995) or sludges from waste water treatment plants (Raphel *et al.*, 1994), and disposal of this kind of dry residue on the landfill itself;
(c) controlled reinjection of the concentrate into changing areas of the landfill in order to improve the biochemical degradation process in the waste itself and accelerate the immobilisation of the organic matter.

This solution should be considered as a first approach, as a landfill usually can be compared to a bioreactor that under optimal operating conditions will produce valuable landfill gas, in this case accelerating at the same time the desired immobilisation of organic components (Pohlard, 1996).

If the system for the controlled return of the concentrate to the landfill is designed according to the needs and the respective conditions on site, no changes for the concentration of pollutants in the leachate being pumped from the landfill to the purification plants are to be expected. This has been demonstrated on different landfills, in one case since 1986 (Henigin, 1995).

Nevertheless, the decision about the concentration factor of the leachate purification system should be based on evaluation criteria that include all kind of economical and ecological aspects specific for each landfill site in order to find an optimised solution. In some cases a one stage unit could be sufficient, in others a combination of processes with the highest possible recovery rate should be selected.

2.3.4 Volume reduction of the concentrate

Steady improvement of membrane technology has resulted in a high pressure RO system based on the DT-module with operation pressure in the range of 120 bar and an adapted process to reduce certain salt fractions by controlled precipitation. With these developments, the limits for the recovery rate in landfill leachate - given among others by the osmotic pressure - have been overcome and the concentration factor for the organic and inorganic matter dissolved in the landfill leachate was doubled This means an increase of the permeate recovery from about 80% - related to a concentration of 5 to 90% recovery with a concentration factor of 10 for the contaminants retained by membrane. Thus, the limit for the electric conductivity in the concentrate of a RO plant was increased from 50 000 to 60 000 µS/cm to the range of 100 000 to 120 000 µS/cm (Peters, 1995).

Due to the high volume reduction relation to this increase of pure water recovery, this technology allows the need for subsequent evaporation steps to be eliminated. After being processed with high pressure RO, the concentrate can then be fed directly into a dryer or a solidification device, or be burned. Such high pressure reverse osmosis systems are in use, actually at about 25 plants.

On the Ihlenberg landfill site high pressure RO has been working since January1992 with a salt rejection of more than 99%, for example from 73 280 µS/cm in the feed to 434 µS/cm in the permeate. The average specific energy demand for this concentrate stage is 14 kWh/m^3 permeate, whereas the leachate stage with 80% recovery in front of this system is consuming less than 5 kWh/m^3. The continuous growth of this landfill and the accompanying increase in leachate required the expansion of the purification capacity. On December 17, 1993 the next RO plant was brought on line with an integrated high-pressure stage (manufacturer PALL ROCHEM) that increased the input capacity to another 48 m^3/h of raw leachate (Kenner and Peters, 1995), being possible under favourable conditions with a permeate recovery up to 90%.

2.3.5 Process improvement with nanofiltration

Even better permeate recovery rates can be reached by installing a combination of nanofiltration (NF) and fractioned removal of solids together with RO and high pressure RO. Nanofiltration allows material dissolved in water to be separated into monovalent and bivalent ions. Consequently, the high rejection rate for sulphate ions and for dissolved organic matter together with very low rejection for chloride and sodium reduces the volume

of concentrate. Some example for rejection rates for different components dissolved in landfill leachate are shown in Table 2.6 (Rautenbach and Linn, 1995).

For this application, very specific NF membranes have to be selected and the module must be suitable designed to optimise the interaction of flow parameters such as feed flow velocity, pressure drop, efficient membrane cleaning, insensitivity to micro-particles. Also, a good cost/performance ratio must be achieved. The requirements for successful and continuous separation of heavily loaded waste water are met by the DTF-module, a flat channel module consisting of only a few components in which open-channel construction is combined with narrow gap techniques.

Table 2.6 : Example for rejection rates for nanofiltration

Parameter	Feed	Permeate	Rejection %
BOD_5 mg O_2/ℓ	480	280	41,62
COD, mg O_2/ℓ	17 000	700	95,88
Ammonia, mg/ℓ	3 350	1 420	57,61
Sulphate, mg/ℓ	31 200	2 345	92,48
Chloride, mg/ℓ	12 760	17 730	-38,95
Calcium, mg/ℓ	2 670	187	93,00
Magnesium, mg/ℓ	1 030	72,7	92,94
Sodium, mg/ℓ	10 900	5 010	54,04
pH-value	6,3	6,4	—
El. conduct, µS/cm	61	43	29,5

A system operated with 8,5 kWh/m^3 of totally produced permeate with a water recovery rate of 97% is one example of the extremely low overall power consumption of this hybrid process using reverse osmosis, NF and fractionated removal of solids.

The fact that a permeate recovery rate of 95 to 97,5% is operating standard today shows that the combination of RO with NF and crystallization is the basis for an economical process for the purification of landfill leachate.

Based on the technology explained above, companies have introduced an “own and operate” service, where clients simply pay a price per m^3 of leachate treated without capital risk and with minimal operational involvement. Also, it is possible to operate equipment for short term emergency situations, caused by heavy precipitation and capacity bottle-necks.

2.3.6 Summary

The results obtained during the operation of an increasing number of plants under very different conditions prove that RO is a very effective instrument for the purification of landfill leachate if all design criteria and requirements specific for landfill leachate have been taken

into consideration, and if an adapted module system as well as correlated technologies are used. This includes high pressure RO, with operating pressure up to 120 bar and/or NF in combination with a controlled crystallization process, that allows permeate recovery rates of more than 95%.

The elimination of the negative impact of landfill leachate on the environment can be achieved with membrane filtration due to the dramatic minimization of residual waste to be processed or immobilized and due to the high quality of the purified water discharged back to nature. The combination of processes designed for this purpose is one example for a sustainable environmentally friendly development.

2.4 Hybrid Processes Involving Membranes for the Treatment of Highly Organic/Inorganic Contaminated Water

Hybrid processes involving membranes for the treatment of high organic / inorganic concentration contaminated water have been described by Rautenbach and Mellis (1995).

2.4.1 Combination of bioreactor, ultrafiltration and nanofiltration

Conventional biological treatment of effluents containing large amounts of recalcitrants requires high residence times, i.e., a voluminous biological treatment process. In such cases, a combination of NF with biological treatment can lead to surprising results. Recalcitrant organics are mostly larger molecules which are rejected by NF membranes whereas the biologically degraded substances and salts will permeate.

By recycling the NF concentrate to the bioreactor, the *residence time* and *concentration* of the recalcitrants are substantially increased without increasing the volume of the bioreactor. As a consequence, the reaction rate, i.e., the rate for biodegradation of recalcitrants, is increased. According to the monod-equation, it increases with concentration until stopped by the eventual increase of toxic substances. Depending on the composition of the waste water to be treated, the unavoidable discharge of reactor effluent with excess sludge can be sufficient for a stable unpoisoned operation of the system on a high concentration level.

Experiments were carried out with dumpsite leachate on a pilot-plant scale. The results are summarized in Table 2.7. By recycling the concentrate of the NF and operating the biological treatment process on a COD level 5 - 10 times above feed level, the elimination rate was increased by 9 - 17%.

During operation up to 6 months, no negative reactions of the biological process were observed. Since the nitrifying bacteria nitrosomas and nitrobacter are very sensitive to toxic components, nitrification is an excellent indicator for the proper operation of the biosystem. In this case, nitrification was always very reliable and complete (nitrite concentration always <2 ppm) during the pilot plant operation.

Table 2.7 : COD elimination rate of 'straight-through' operated biological process vs biological process with nanofiltration concentrate recycle.

COD elimination rate %	Dump site			
	C	D	A	B
Reference stage	58	74	62	45
Recycling stage	75	86	71	56

Where the bleed with excess sludge is not sufficient for stable operation, a partial reduction of COD by either chemical oxidation or adsorption is recommended. Compared to an installation behind a conventional biological treatment process, there are the following advantages : -

- Since the stages are fed with the concentrate of the NF, they are operating at high concentrations, i.e., very effectively.

- The operation of the chemical oxidation or adsorption is not determined by the low concentrations which must be guaranteed before discharge.

- With chemical oxidation, *partial oxidation* of nondegradable components to biodegradable substances is possible since the effluent of the oxidation is recycled to the bioreactor.

After the successful completion of the pilot experiment, the first large-scale installation was under construction (Berg dumpsite).

2.4.2 Dumpsite leachate treatment by high pressure reverse osmosis

In December 1989 the leachate treatment of the Schönberg dumpsite in Germany was commissioned. It consisted of two-stage RO with evaporation of the RO concentrate and its final solidification with fly ash from the power plant. The first stage of the RO unit is equipped with DT-disc modules and the second stage with spiral-wound elements. With an installed membrane area of 1 147 m^2 in the first stage and 768 m^2 in the second stage, it is the largest leachate RO treatment plant in Germany.

In operation, the evaporator had a very low availability as a consequence of heavy fouling/scaling of the plate heat exchanger and, subsequently, the high number of cleaning cycles. A replacement of this evaporator by high-pressure RO proved to be very successful.

The concentrate of the first RO stage operating at 60 bar is concentrated to such high values by the high-pressure stage that solidification by fly ash is possible without further treatment (Table 2.8).

The high-pressure RO stage is subdivided into blocks operating at 80, 120 and 200 bar. All stages have meanwhile been operated for more than 10 000 h without membrane

replacement. Because of the high NaCl concentration in the concentrate, the flux of the 200 bar stage is rather low but still economical (Table 2.9).

Rejection with respect to inorganics (conductivity) and organics is high in all stages as well as the availability (Table 2.10), which is, however, somewhat lower for the 200 bar units. This lower availability of the 200 bar unit is due to automatic shutdowns at a feed temperature of 40 °C which can easily occur because of the high rate of dissipated energy at 200 bar.

Table 2.8 : Comparison of leachate, concentrate and permeate at the Schönberg dumpsite

Parameter	Unit	Leachate	Concentrate high-pressure RO	RO permeate Stage 1
Evaporation residue	mg/ℓ	15 790	102 000	155
Conductivity	µS/cm	20 450	102 000	250
Ammonia	mg/ℓ	513	3 120	7
Sodium	mg/ℓ	3 787	30 667	30
Calcium	mg/ℓ	185	1 133	2
Potassium	mg/ℓ	1 126	—	12
Chloride	mg/ℓ	4 681	27 168	60
Sulphate	mg/ℓ	3 765	29 514	12

Table 2.9 : Schönberg RO plant, operational results in 1992.

Pressure in stage (bar)	Permeate flux (ℓ/m^2h)	Concentration factor (—)	Conductivity feed (µS/cm)	Rejection conductivity (%)
60	11 - 13	3,2	17 - 23	98 - 99
80	9 - 11	3,8	40 - 60	98 - 99
120	6 - 8	6,1	60 - 80	98 - 99
200	4 - 7	6,7	100 - 120	98 - 99

Table 2.10 : Schönberg RO plant, operational results in 1992.

Pressure in stage (bar)	Consumption of cleaning agent (ℓ/m^3 permeate)	Availability (%)	Energy consumption (kWh/m^3 permeate)	Specific processing costs (DM/m^3 permeate)
60	0,19	95	4,87	18,57
80	1,06	86	16,05	57,82
120	1,42	87	16,05	57,82
200	2,10	73	35,46	133,79

The specific power consumption of the 60 bar units is only 5 kWh/m^3 permeate. This is an excellent figure compared to tubular systems for leachate treatment with about 8-12 kWh/m^3 permeate. The power consumption of the 200 bar units (35 kWh/m^3 permeate) is relatively high. It must be emphasized, however, that it operates in a concentration range where until now only evaporators could be employed. Compared to the power consumption of the thermodynamically very efficient evaporator with mechanical vapour compression of 50 - 60 kWh/m^3 distillate, the consumption of the high-pressure RO is still low.

The specific costs listed in Table 2.10 were calculated for a 3 year depreciation period. Averaged, these costs result in a figure of 36 DM/m^3 permeate at an overall concentration factor of *CF = 6.7* for the entire RO plant. For a depreciation time period of 10 years, the specific overall costs would be about 25 DM/m^3 permeate.

2.5 High-pressure Reverse Osmosis and Nanofiltration, a 'Zero Discharge' Process Combination for the Treatment of Wastewater with Severe Fouling/Scaling Potential

A zero-discharge process combination for the treatment of wastewater with severe fouling / scaling potential has been described by Rautenbach and Linn (1996).

2.5.1 Implementation of nanofiltration

Chlorides are a major part of the inorganic components of leachates. Due to the high rejection by RO membranes, they are almost completely in the concentrate with the negative effect of a high osmotic pressure and, in addition, the disposal problem. A positive effect for the operation of a RO process is the increased solubility of scalants such as $CaSO_4$ in the presence of chlorides.

The implementation of a NF stage into the treatment process can extend the limits of RO set by scaling and/or osmotic pressure considerably. One of the features of NF membranes is their significantly different selectivity for mono- and bivalent anions. Whereas the rejection of chlorides is very low — about 30% for NaCl in the case of an aqueous binary solution — the rejection is about 98% for $CaSO_4$ in the case of aqueous binary solution. Even negative rejection coefficients for NaCl can be observed in the presence of SO_4^{2-} ions, depending on the SO_4^{2-} concentration.

The concentrate of the 120 bar RO stage is treated by a combination of NF/crystallization. At moderate transmembrane pressure differences of 20 - 50 bar, the NF unit produces a permeate containing mainly chlorides. This permeate can be concentrated further without danger of scaling by a 200 bar HPRO or discharged into surface waters in cases where the discharge of chlorides is not restricted (this might require a two-stage NF cascade in order to meet the limiting figures regarding COD).

Essential for the process is the operation of the NF/crystalliser cycle in the range of super-saturation with respect to calcium sulphate and at high concentrations of organics. For this reason, the success of the process depends to a large extent on a proper module design; the modules must be insensitive to fouling and the presence of crystals. Such modules have been developed by Rochem, Germany. Stacks of rectangular membrane cushions and matching spacer plates are arranged in series in the pressure vessel in such a way that stagnant areas are avoided and internal friction losses are minimized. Feed flow is strictly parallel to the vessel axis; the velocity in the feedside module channels is about 1,5 m/s.

In cooperation with the Rochem Company and the Institut für Verfarhrenstechnik, RWTH, Aachen, the process has been installed and tested on as technical scale at the Ihlenberg dumpsite. A NF stage consisting of four blocks with nine modules each and a total membrane area of 180 m^2 has been added to the existing RO-HPRO combination and commissioned in September 1994. The unit is designed for the treatment of about 4 m^3/h RO concentrate of a concentration factor of:

$$CF_{V\,RO} = \left[\frac{V_F}{V_R} \right]_{RO} = 10$$

The colour of the NF feed is dark while the colour of the NF permeate is water white. Depending on the concentration factor of the 60 and 120 bar RO stage, the NF stage achieves a further concentration of

$$CF_{V\,NF} = \left[\frac{V_F}{V_R} \right]_{NF} = 10 \div 20$$

Sludge consisting of organics, precipitated inorganics and water is discontinuously withdrawn from the bottom of the crystalliser/sedimentation tank. With the added concentrate of the subsequent 200 bar HPRO stage, this concentrate is disposed of at Ihlenberg by solidification with fly ash and final storage at the dumpsite. It should be emphasized that the process scheme realized in Ihlenberg is an extreme case due to the high salt concentration of the leachate. In many other cases, a process configuration consisting of a 60 bar RO stage, a NF and a final 120 bar RO stage will be sufficient for water recovery rates of 95% or more. Cleaning of the NF modules consists of flushing with feed at zero transmembrane pressure difference for 30 s every hour and an alkaline cleaning every 250 - 300 h.

The results obtained in the operation period September - December 1994 are as follows. After start-up, an average permeate flow rate of 3,6 m^3/h was achieved at transmembrane pressure differences of 20 bar initially, increasing to 40 bar before cleaning. The rejection rate for sulphates was 92 - 95%, for the dissolved solids (conductivity) 20 - 35%. Table 2.11

lists the data for the relevant parameters. Interesting are the high rejection of organics (COD) of about 92% and the unexpected high rejection of about 50% for NH_3-NH_4^+. Similar data have been obtained for this process at the Halle-Lochau dumpsite. In both cases the chloride concentration of the leachate is unusually high. Accordingly, even better results with respect to cleaning cycles, overall water recovery and specific energy consumption can be expected in other cases.

At the Ihlenberg dumpsite (Table 2.12), the specific energy consumption of the NF and the subsequent 200 bar RO stage is 32 kWh/m^3 permeate. This figure is a direct consequence of a typical leachate consumption of the Ihlenberg dumpsite. However, with a figure of 8,5 kWh/m^3 of totally produced permeate and an overall water recovery rate of 97%, the overall specific power consumption of the process is extremely low compared to other processes. This process can still be improved. By the implementation of a hydrocyclone or a filtration step, the present discontinuous operation of the crystallization/sedimentation can be eliminated and by recycling of the sludge filtrate increase the water recovery rate to even higher figures than 97%.

This process, with respect to water is an almost zero-discharge process, is presently being developed and studied in East Germany.

Table 2.11 : Analysis of the dumpsite leachate and rejection rates of the nanofiltration.

	Feed	Permeate	Rejection rate, %
Ihlenberg:			
BOD, mg/ℓ O_2	480	280	41,67
COD, mg/ℓ O_2	17 000	700	95,88
Ammonia, mg/ℓ	3 350	1 420	57,61
Sulphate, mg/ℓ	31 200	2 345	92,48
Chloride, mg/ℓ	12 760	17 730	-38,95
Calcium, mg/ℓ	2 670	187	93,00
Magnesium, mg/ℓ	1 030	72,7	92,94
Sodium, mg/ℓ	10 900	5 010	54,04
pH	6,3	6,4	—
Conductivity, mS/cm	31,2	33,9	- 8,65
Halle-Lochau:			
BOD, mg/ℓ O_2	591,4	239,6	59,59
COD, mg/ℓ O_2	6 400	600	90,63
Ammonia, mg/ℓ	2 260	1 020	54,87
Sulphate, mg/ℓ	NV	NV	—
Chloride, mg/ℓ	29 780	35 629	-19,64
Calcium, mg/ℓ	243	31	87,24
Magnesium, mg/ℓ	952	133	86,03
Sodium, mg/ℓ	41 000	20 390	50,27
pH	6,23	6,31	—
Conductivity, mS/cm	—	—	—

Table 2.12 : Specific energy consumption of the Ihlenberg leachate treatment plant (capacity: 50 m^3/h dumpsite leachate).

Reverse osmosis/high pressure reverse osmosis		
Part of plant	Permeate flow rate, m^3/h	Specific energy consumption, kWh/m^3 permeate
60 bar RO HP-RO	35 11	4 17
Nanofiltration/high pressure reverse osmosis		
NF } RO }	2,7	32
Total plant :		
Water recovery, %		97,40
Specific energy consumption, kWh/m^3 perm.		8,5

2.5.2 Summary

The treatment of highly polluted water by RO is a reliable and economic operation and can be considered as state of the art. The process can produce water of any required quality — if necessary in a cascaded operation.

A major problem of waste water treatment is the water recovery rate, which should be near to 100%, realized in a simple and energy-efficient process combination. As demonstrated by long-time experiments in pilot plants and on a technical scale, this can be achieved by the addition of NF and 200 bar high pressure RO to the 60/120 bar RO stage. The integration of the simple mechanical unit operation crystalliser/hydrocyclone/filtration promises an almost zero discharge process.

2.6 Waste Water Treatment by Membrane Processes - New Development in Ultrafiltration, Nanofiltration and Reverse Osmosis

Wastewater treatment by membrane processes, new developments in ultrafiltration, nanofiltration and reverse osmosis has been described by Rautenbach *et al*. (1996).

2.6.1 Reverse osmosis/nanofiltration/high pressure reverse osmosis for a zero-discharge treatment of dumpsite leachate

Nanofiltration is a still relatively new pressure driven membrane process. Regarding driving force and rejection of organic components in aqueous solutions, NF is somewhere between RO and ultrafiltration (Table 2.13).

Table 2.13 : Driving force and selectivity of reverse osmosis, nanofiltration and ultrafiltration.

	Pressure (driving force), bar	Cut-off g/mol
Reverse osmosis	30 - 200	<200
Nanofiltration	10 - 40	200 - 1 000
Ultrafiltration	2 - 10	1 000 - 100 000

Like RO, NF is based on solution-diffusion as the major transport mechanism. But contrary to RO, NF membranes contain fixed (negatively) charged functional groups. As a consequence, the selectivity of NF for monovalent and bivalent anions is significantly different. Table 2.14 shows results of experiments with binary aqueous solutions. Typically, the rejection for chlorides and nitrates is about 50%, but for sulphates 96 - 98%.

In a number of cases this different rejection of mono and bivalent anions permits the realisation of the Donnan Effect: In a multicomponent system containing among others sulphate and chloride ions, the chloride rejection in shifted towards negative figures with increasing concentration of the sulphate ions - chlorides are forced into the permeate with increasing sulphate concentration.

With respect to dissolved organic components, the difference between RO and NF is only of quantitative nature; Whereas components of a molecular mass of about 150 kg/kmol are almost totally rejected by reverse osmosis, the 'cut-off' of NF is above 300 kg/kmol.

Table 2.14: Characteristics of reverse osmosis and nanofiltration membranes

Reverse Osmosis		Nanofiltration	
Membrane	Reject. rate, %	Membrane	Reject. rate, %
Chloride (monovalent)			
FilmTec FT30SW	99,1	FilmTec NF45	55
Desal 3S	99,2	Desal 5K	50
Toray UTC80	99,4	Toray UTC60	60
Nitto NTR759H	99,5	Nitto NTR7450	50
Sulphate (bivalent)	<99,8	FilmTec NF45	98,5
FilmTec FT30SW		Desal 5K	96,4 - 98

Essentially based on the special feature of NF - the significantly different rejection of mono- and bivalent anions - a new process was developed for dumpsite leachate treatment.

Dumpsite leachate is a direct consequence of rainfall. The average leachate production of a dumpsite in western Europe is about 5 m^3/ha.d. Leachate contains a rather complex mixture of components but mainly inorganics.

The flow diagram of the treatment process consists of: A 2-stage RO operating at 60 bar and 120 bar in the first stages, and 60 bar in the second stages. The concentrate of the 120 bar unit is treated by a combination of NF/crystallisation. At moderate transmembrane pres-

sure differences of 20 - 50 bar the NF unit produces a permeate containing mainly chlorides. This permeate can be concentrated further without danger of scaling by a 200 bar HPRO.

Essential for the process is the operation of the NF/crystalliser cycle in the range of supersaturation with respect to calcium sulphate and at high concentrations of organics. For this reason the success of the process depends to a large extent on a proper module design - the modules must be insensitive to fouling and the presence of crystals. Such modules have been developed by Rochem, Germany. Stacks of rectangular membrane cushions and matching spacer plates are arranged in series in the pressure vessel in such a way that stagnant areas are avoided and internal friction losses are minimized. Feed flow is strictly parallel to the vessel axis, the velocity in the feedside module channels is about 1,5 m/s.

In cooperation of the Rochem company and the Institute für Verfahrenstechnik, RWTH Aachen, the process has been installed and tested on a technical scale at the dumpsite Ihlenberg. A NF stage consisting of 4 blocks with 9 modules each and total membrane area of 180 m^2 has been added to the existing RO-HPRO combination and commissioned in September 1994. The unit is designed for the treatment of about 4 m^3/h RO concentrate of a concentration factor of :

$$CF_{V\,RO} = (V_F / V_R)_{RO} = 10$$

Depending on the concentration factor of the 60 and 120 bar RO-stage, the NF-stage achieves a further concentration

$$CF_{V\ NF} = (V_F / V_R)_{NF} = 10 \div 20$$

A sludge consisting of organics, precipitated inorganics and water is discontinuously withdrawn from the bottom of the crystalliser / sedimentation tank.

With the added concentrate of the subsequent 200 bar HPRO stage this concentrate is disposed of at Ihlenberg by solidification with fly ash and final storage at the dumpsite.

Cleaning of the NF-modules consists of flushing with feed at zero transmembrane pressure difference for 30 seconds every hour and an alkaline cleaning every 250 - 300 hours.

The results obtained in the operation period September / December 94 are as follows: - After startup, an average permeate flow rate of 3,6 m^3/h was achieved at transmembrane pressure differences of, initially 20 bar, increasing to 40 bar before cleaning.

The rejection rate for sulphates was 92 - 95%, for the dissolved solids (conductivity) 20-30%. Table 2.15 lists the data for other relevant parameters. Interesting are the high rejection of organics (COD) of about 96% and the unexpected high rejection of about 58% for NH_3/NH_4^+.

At the dumpsite Ihlenberg (Table 2.16) the specific energy consumption of the NF and the subsequent 200 bar RO stage is 32 kWh/m^3 permeate. This figure is a direct consequence of the untypical leachate composition of the Ihlenberg dumpsite.

But with figures of 8,5 kWh/m^3 of totally produced permeate and an overall water recovery rate of 97%, the overall specific power consumption of the process is extremely low compared to other processes.

Table 2.15: Analysis of the dumpsite leachate and the rejection rates of nanofiltration

Parameter	Feed	Permeate	Reject. rate
BOD, mg/ℓ O_2	480	280	41,67
COD, mg/ℓ O_2	17 000	700	95,88
Ammonia, mg/ℓ	3 350	1 420	57,61
Sulphate, mg/ℓ	31 200	2 345	92,48
Chloride, mg/ℓ	12 760	17 730	-38,95
Calcium, mg/ℓ	2 670	187	93,00
Magnesium, mg/ℓ	1 030	72,7	92,94
Sodium, mg/ℓ	10 900	5 010	54,04
pH	6,3	6,4	—
Conductivity, Ms/cm	61	43	29,5

Table 2.16: Specific energy consumption of the Ihlenberg leachate treatment plant (capacity: 50 m^3/h dumpsite leachate)

Part of plant	Permeate flow rate m^3/h	Specific energy consumption	
		Process stage kWh/m^3 permeate	Total plant kWh/m^3 feed
RO/high pressure RO			
60-bar RO	35	4	2,8
HP-RO	11	17	3,74
NF/high pressure RO			
NF	2,7	32	1,73
RO			
Water recovery: 97,40%	Total plant :		8,27

This process, with respect to water is an almost zero-discharge process, is presently developed and studied in Germany.

2.6.2 Summary

The two examples discussed in this paper illustrate the steadily increasing impact of membrane processes on modern wastewater treatment and, vice versa, how the requirements of individual separation problems intensified research and development.

An especially rapid development can be presently noticed for ultrafiltration/micro-filtration and for RO/NF i.e. processes which are commonly considered as rather well developed.

In ultrafiltration, research and development concentrate on module design aiming at low specific energy consumption and — simultaneously — low investment costs with the special aim of replacing sedimentation/sandfiltration in municipal sewage treatment plants.

In RO /NF, research and development concentrate on shifting the limits of processes to very high water recoveries, i.e. the development of 'almost zero discharge' processes. As shown in the second example, this is also strongly related to module development.

2.7 Treatment Processes of Sanitary Landfill Leachates

The treatment of RO concentrate in the treatment of landfill leachate has been described by Biehler and Hägele (1995).

2.7.1 Reverse osmosis (concentrate)

In the course of the treatment of landfill leachates by reverse osmosis (single and multi-level) a concentrate of 15-25% of the amount of the raw leachates accumulates, the difference being determined by the chosen concentration factor CFv. In principle, there are three possibilities for the disposal of concentrate : -

- a return to the landfill
- evaporation and drying-out
- incineration (for example, in a waste incineration plant)

(a) **Return to the landfill**

In an investigation of the return of concentrates to the landfill, the following points are listed as possible negative influences upon the processes in the landfill body: -

- the raising of the quantity of leachate by the quantity of the concentrate. Depending on the storage intensity of the landfill body, this results in an increase of leachate outflow.

- an increase in concentration, especially of the content's inorganic substances.

It is assumed that the waste body, as an anaerobic bio-reactor, degrades most of the organic substances of the concentrate's content.

The increase in the quantity of run-off can, for instance, be limited by prompt surface sealing or the covering of sites which are already filled. This would permit a return beyond the

operating face. However, when the storage capacity of the body of refuse is reached, almost 100% of precipitation and concentrate percolate through the landfill, thereby leaching up to a 100% greater inflow concentration. A return of landfill leachates after the final filling is therefore not advisable.

The potential effects on the quality of leachates (for instance, an increase in conductivity) are to be taken into consideration when planning the installation. In a sample calculation, it was stated that a 15% higher inflow and a 35% increase in concentration result in 15 - 20% higher processing costs per m^3 of leachate.

In the Gaggenau-Oberweier landfill of the Rastatt district a return of the concentrate has been carried out since 1968. In this the leachate is led into the body of refuse by way of swallow-holes — 4 to 5-metre deep excavations, which have subsequently been filled with gravel — with a perforated tube in the centre. The proportion of concentrate in the quantity of leachate entering the landfill body is almost 10%.

Summarizing the management experiences of the period 1986 - 89, the problem of encrustations in the return duct was only described, which can be avoided by the selection of a larger performance cross-section (DN 50) and the rinsing of the duct during periods when it stands idle. During an inspection of the plant, it was mentioned that in the area surrounding the swallow-holes, the quality of landfill gas was reduced by increased hydrogen sulphide concentrations. The reason for this might be the lowering of the pH value by sulphuric acid. No upward concentration of the raw leachate was observed during the 3-year observation period. There are no indications available as to the behaviour of the concentrate in the swallow-holes.

(b) Evaporation and drying-out

If substances which have been precipitated by earlier processes can be made acceptable for inclusion in landfills, they must be transformed into solids from their initial presence in water solution. A further reduction in volume is to avoid unnecessary costs for transport and storage. This leads to the aim of removing most of the remaining water as well. The separation of water is best achieved by evaporation by a crystallization evaporator. In principle, such a procedure can also be applied to raw leachate. However, since the energy input constitutes a major cost factor in this, application is only recommended for previously enriched wastewater (for instance, reverse osmosis) or for uncommonly salty leachate (for example in the Schwabach hazardous waste landfill).

Apart from extensive laboratory experiments, prolonged practical experience is available only in the case of Schwabach. In the course of the work at the plant there have been frequent problems with dumping and corrosion. Another troublesome factor is the unwanted development of foam. Vaporization technology has, however, meanwhile moved on. Thanks to new materials and the application of the fluidized bed technique, greater reliability and economy can now be expected.

Depending on the existing salt concentration, the initial liquid is being concentrated 5 to 15-fold, up to the point when the evaporation of water leaves a thick paste. The solids content has then reached 50 to 80%. The product might even be ready for storage in suitable containers (such as 'Big Bags').

If a centrifuge is added, the dry mass may even attain 90%. Thermal drying is capable of lowering the water content to 1%. The new raw material is solid. The substance consists of irregular to powdery fragments (thin film drying). A relatively new process is fluidized bed drying. Apart from expecting lower running costs, it commends itself because the solid material is present in a granulated form. The uniformly shaped spheres guarantee easy handling and storability of the residual substance. In addition, there is the possibility of coating the new formed grain with a water-repellent layer. Again, it is unfortunately that so far no experience in practice is available. A fundamental problem with evaporation/drying is the simultaneous evaporation of substances in the leachate which are more volatile than water. They return in the condensate. If this occurs to an intolerable extent, further treatments must be envisaged. The water must either be stripped of every possibly problematic substance in advance or the distillate must be passed through a filter (activated charcoal, for instance).

(c) Incineration

There has hitherto been no investigation of the burning of landfill leachate concentrates in a refuse incineration plant. In principle, it makes sense from an energy viewpoint to achieve a substantial reduction of volume before incineration. Due to the — in part — very high AOX contents, the question of emission burdens arose above all else, and thus the necessity of having the concentrate only incinerated in hazardous waste plants.

2.8 Literature Investigation Regarding Leachate Treatment Focussing on High COD Values (Enviroserv)

The treatment of leachate focussing on high COD values has been summarized by Thomas *et al.* (1999).

2.8.1 Method of Leachate Treatment

Table 2.17 below summarises the leachate treatment methods discussed in the document.

Table 2.17: Summary of leachate treatment methods

Method of treatment	Treatment type	Disadvantages	Advantages	Range of COD	Relative Cost
Membrane bioreactor (MBR)	Biological	• Variations in flow and quality • Effectiveness for very high COD? • Not totally effective for all leachate constituents and must therefore be used in combination with other processes	• Compact biological treatment system • Due to high oxygen transfer efficiencies, high volumetric rates can be achieved • Cost effective option • Pollutants are degraded. • Good capacity in the reduction of the organic load • Avoids the construction of a sedimentation tank therefore eliminates the problems of management of the tank. • The use of MBR with evaporation showed a significant decrease in the COD	< 4 000 mg/ℓ Process can eliminate more than 70% of the COD	High
Activated Sludge (co-disposal)	Biological	• Odorous • Not totally effective for all leachate constituents and must therefore be used in combination with other processes.	• Managed and gradual release of landfill emissions. • Affordable and appropriate technology for developing countries. • Promotion of anaerobic digestion activity in the landfill as a result of increased moisture regimes.	< 4000 mg/ℓ COD removal can be 90 - 98%	Medium
Activated carbon	Biological	• Not totally effective for all leachate constituents and must therefore be used in combination with other processes. • Cost of carbon regeneration high. • Logistics • Emissions control for regeneration process (and licensing). • Not as effective for small molecule compounds. • Less effective for non-organic compounds.	• It can be recycled and therefore there are no concentrated waste products to be disposed of at the site. • The organic compounds that are adsorbed onto the carbon are destroyed in the re-activation process due to intense heat. • Methanol is poorly adsorbed and chlorinated aromatic compounds are very well adsorbed.	< 4000 mg/ℓ Removes up to 50% of the COD	Medium
Chemical oxidation	Physico-chemical	• Not totally effective for all leachate constituents and must therefore be used in combination with other processes. • Air emissions. • Odour. • Expensive on materials.	• It is better than chlorine as it is un-affected by the presence of ammonia. • It is better than precipitants, flocculants, activated carbon and reverse osmosis because there is no residue to disposed of. • The high oxidation potential of ozone allows it to oxidise recalcitrant compounds. • Molecular ozone reacts selectively with organic compounds, the main direct reaction being electrophilic addition. • Increases in ozone dose produce greater reductions in COD in the leachate following ozonation.	Used for high COD levels. Potential to reduce COD levels by between 35 - 48%.	Medium

Method of treatment	Treatment type	Disadvantages	Advantages	Range of COD	Relative Cost
Reverse osmosis	Other	• Not totally effective for all leachate constituents and must therefore be used in combination with other processes. • Very expensive. • Not all compounds are removed.	• Simultaneous removal of salinity and the organic pollutant • Downsized automatic operation. • Stable effluent quality. • Established technology.	—	Very High
Mediaflex™ filter	New Technology	• Limitations of the hydraulic load	• Membranes can be adapted to the type of contaminants to be eliminated.	—	High

2.8.2 Recent Developments in Leachate Treatment Technology

(a) New evaporation treatment

The new evaporation treatment plant in Helsinki, Finland, claimed to use less energy and has been found cheaper to build than conventional evaporation plants. Studies have shown high purification rates despite low capital and operating costs. The new system uses mechanical vapour recompression and the 'falling film' principle. High quality polymeric materials have been incorporated into the heat transfer surface to give large areas for heat transfer at a much lower cost than typical metal surfaces of similar size. This is the key factor in the low energy consumption.

(b) Mediaflex^{TM} filter

The Mediaflex^{TM} filter is a multi-layered process for dealing with liquid effluents. Mediaflex^{TM} filters can be organised in a line configuration, filter A deals mainly with organic matter (BOD and COD), while filter B removes metals, colourants, metals, grease oil and phenols. The process can be combined into a single multi-layered filter unit incorporating both filters A and B. It is also possible to build an aeration system into filter A, so as to improve its biological treatment capacity.

Advantages of such a system include the fact that membranes can be adapted to the type of contaminants to be eliminated.

Disadvantages include limitations of the hydraulic load

2.8.3 Summary

The various methods of leachate treatment discussed above, may be combined in various modes with other standard chemical engineering unit processes, providing a huge range of variable options are available from which a solution must be engineered to optimise the balance between cost and quality. Below are some examples of combinations of these processes that have been shown to be successful in previous studies (Kapetanios et al., 1995).

- Chemical oxidation - Biological oxidation - Chemical precipitation
- Biological oxidation - Chemical oxidation - Biological oxidation - Chemical precipitation
- Chemical precipitation - Chemical oxidation - Biological oxidation
- Chemical precipitation - Biological oxidation - Chemical oxidation - Biological oxidation.

In researching the leachate treatment requirements for Holfontein in some detail it has become apparent that the solution may be found not only in leachate treatment technology but in chemical process technology as well. Alternatives include : -

1. *The Phenosolvan process patented by Lürgi and employed by Sasol for the extraction of phenols from water and other process streams.* This certainly shows promise and has been demonstrated to be successful and is obtainable either through Sasol or directly from Lürgi. The Sasol process provides for the removal of meta para cresols from the phenolic tar acid stream. The pH and ammonia levels are important considerations. Unfortunately, the costs will tend to be high and may make thus unaffordable but they are certainly worth looking at.

2. *Activated Carbon or related chemical dosage technology:* Having briefly scanned the chemical engineering sector of South Africa, there appears to be good promise of finding the solution to the problem doing inexpensive practical and grassroots research, using a stir tank of the leachate and adding various chemical treatments to research effectiveness. The could include activated carbon with varying granule sizes and a number of other possibilities.

It was recommended that the above two as well as a number of other chemical engineering and process options be investigated further, with possible wider application than just for one landfill.

In addition, further investigations are recommended into the most appropriate method of leachate treatment. The following strategy could be employed :

1. Problem definition

   This phase has already been completed and involved:

   - Determination of the quantity and quality of leachate
   - Determination of the size of the problem, i.e. measure the flow rates of the stream, etc.
   - Conduct a literature survey.

2. Problem investigation

   - Laboratory scale investigation.
   - Bench scale investigation to identify processes suitable for reducing the waste strength. Such testing could include carbon isotherms, ammonia stripping at an elevated pH, toxicity screen using ATP and biological treatment using both aerobic and anaerobic cultures (Kang *et al.,* 1990).
   - Larger scale trials (if necessary).
   - Process Engineering and Cost Estimation.

3. Solution Implementation

Environmental companies, in association with local and international associates in the waste management and chemical process fields would be able to further research and develop any of the technologies listed above.

2.9 Leachate Minimization by Reverse Osmosis

Leachate treatment by RO was discussed by Bilstad and Madland (1992).

Leachates from chemical and domestic landfills are defined as hazardous wastewater. Quantitative and qualitative control of leachate can be performed by membrane separation of the total produced leachate volume. Dissolved and suspended solids in the leachate are removed from the major portion of the feed water phase and either returned to the landfill or further treated. The particle free permeate meets the effluent requirements for direct discharge to virtually any water course.

An untreated leachate flow is concentrated thirteen times by tubular type reverse osmosis. The separation efficiencies are 99% for iron, copper, chromium and zinc. For suspended solids the removal is 100%.

2.10 Bio-Enhanced Treatment of Hazardous Landfill Leachate

The bio-enhanced treatment of a hazardous leachate (Holfontein) was described by Ambachtsheer, 1999.

Leachate originating from a hazardous landfill site (Holfontein) has to be treated on site prior to discharge to sewer. As part of the treatment process the organic components must be removed, prior to desalination, by a biological process. In order to determine the treatability of the leachate a pilot study was conducted using a 1 000 litre reactor operating as a sequential batch reactor (SBR). The character of the leachate is not conducive to microbial degradation due to high salt concentrations (8%) and various toxic compounds such as phenol concentrations as high as 9 700 mg/ℓ. In order to enable the microorganisms to function under those inhibiting conditions a proprietary biosurfactant technology was incorporated in the biological process. This enhanced the microbial degradation capability of the process removing 82% of the chemical oxygen demand (108 500 mg/ℓ to 19 710 mg/ℓ), 96% of the ammonia (2 575 mg/ℓ to 48 mg/ℓ) and 99% of the phenol (9 663 mg/ℓ to 15 mg/ℓ) with a hydraulic residence time of 4 days.

2.11 Appropriate Biological Treatment of Landfill Leachates with Full Nitrification and Denitrification

Appropriate biological treatment of landfill leachates with full nitrification and denitrification has been described by Strachan *et al*, (2000).

Landfill leachate, the liquid mixture, which comprises numerous contaminants produced as rainfall percolates through decomposing domestic and industrial wastes, can be hundreds of times stronger than domestic sewage. Landfill leachate may typically display COD levels in excess of 50 000 mg/ℓ, and concentrations of ammoniacal-nitrogen in excess of 4 000 mg N/ℓ. Landfill legislation, design and construction have undergone enforced evolution in recent years, as South Africa has rapidly moved towards the World's developing standards. Current legislation controlling landfill development in South Africa was *kick-started* by the implementation of the Government's 'Minimum Requirements' document in 1994, which was revised in 1998, demanding that landfills are designed, constructed and operated to Government-guided levels. It has been estimated that South Africa's surface water resources may no longer meet the demands of the entire population within the next 20 years, and the utilisation of ground water reserved is increasingly becoming the norm. It is therefore essential that these ground water reserves be well protected by legislation, especially in regions of high rainfall where the risk of contamination from landfill leachate is significantly greater, such as the Durban Metropolitan Area (the DMA).

The work by Strachan *et al.* presents a case study of typical requirements for landfill leachate treatment in South Africa, which until very recently was rarely even considered in landfill design in the country. There have been a few *dabblings* into treatment of leachate in South Africa, but treatment is reportedly ineffective in sustainably achieving legislated discharge standards. Landfill developments in the Durban Metropolitan Area (DMA), an area encompassing some 3,5 million inhabitants, are discussed. A leachate treatment pilot plant is described that has been established and commissioned at the Bisasar Road Landfill Site in Durban, to assess the treatability of landfill leachates throughout the DMA. Complete biological nitrification and denitrification of landfill leachates has been achieved inside the same SBR unit, and the viability of establishing full-scale treatment plants at landfill sites has been assessed. Levels of some 2 000 mg/ℓ of ammoniacal-nitrogen has shown to be completely and consistently nitrified to nitrate-N, and in turn released as harmless nitrogen gas, resulting in only negligible levels of nitrogenous components in the treated effluents. Sludge build-ups are minimal, and sludge waste has proven to be an infrequent occurrence.

A significant benefit from the treatment systems is the minimal adjustment that are required to pH-levels. The utilisation of methanol for the denitrification processes, which has been utilised to supplement carbon *food source* demands, has shown to be up to twice the generally accepted theoretical ratio of 3:1. Typical nitrification rates, for successful treatment have shown to range up to 0,040 kg NH_4-N/kg MLSS.day, whilst for COD reductions, removal rates of up to 0,15 kg COD/kg MLSS.day have resulted. Pilot plant design and commissioning details are provided in the paper, along with operating data and analytical results for raw leachates and treated effluents. Future and ongoing research work involves the assessment of denitrification processes utilising waste molasses, and the feasibility of further *polishing* treatment, aimed in particular at the removal of residual COD levels, using constructed wetlands.

2.12 Holfontein Leachate Treatment

Enviroserv Waste Management (Pty) Limited has recently presented the following information to DWAF regarding actions to treat the ISWL at Holfontein (Enviroserv Waste Management (Pty) Limited, 2001).

2.12.1 Overview

Enviroserv (Pty) Ltd. are investigating the possibility of installing a plant to treat the leachate from their Holfontein waste disposal site. The aim is to purify the leachate to a level where it can be disposed of to the adjacent river i.e. to river water quality.

Approximately 270 000 kℓ of leachate are currently stored in dams on the Holfontein site. Fresh leachate is being produced from the various waste disposal cells and together with rain water amounts to between 2 000 and 3 000 kℓ per month. The proposed leachate treatment plant is designed to process the fresh leachate as well as to handle the stored leachate so that the leachate dams are empty by the year 2023. The design capacity of the plant is 3 800 kℓ/month.

The quality of the leachate at Holfontein is unique. It has much higher organics (as measured by chemical oxygen demand (COD)) as well as inorganics (as measured by electrical conductivity) than leachates elsewhere in South Africa or overseas. It also contains significant concentrations of phenols that are toxic to biological treatment organisms. Tried and proven purification processes used elsewhere in the world for leachate purification, such as chemical treatment, membrane processes, biological treatment, ozonation, etc. as well as some proprietary filtration / separation technologies, proved ineffective when tested in the laboratory or on site. It became apparent from this early work that no process, on its own, would be able to achieve river water quality and that a combination of processes would be required. It also appeared that the only process that could achieve the initial purification was evaporation / crystallisation. The condensate produced by the evaporator / crystalliser would then require further purification. Two alternative process routes to achieve this were successfully tested at pilot plant scale. The first involved treating the condensate with activated carbon. The second involved aerobic biological treatment followed by activated carbon.

The next stage of the project will include discussions and approval of the project by the relevant authorities and approval of the project by Enviroserv Waste Management. This will then be followed by the engineering, procurement, construction and commissioning of the full-scale plant.

2.12.2 Design basis for the plant

The plant is designed to handle all the leachate arising from the waste disposal cells plus emptying the leachate dams over a 20 year period. Extensive water balance modelling was

undertaken by Jones & Wagner, a consulting engineering company. Various scenarios were modelled covering aspects such as average versus 75 percentile rainfall, average evaporation rate and emptying the dams over periods of 10, 15 or 20 years. The modelling work took into account the building of the various cells to cater for the expected waste disposal demands.

A number of conservative assumptions were made in the model regarding the leachate arisings. With these in mind, the scenario of average rainfall, average evaporation and 20 year life was proposed as the design basis. This scenario required the treatment plant to have a capacity of 3 800 kℓ/month.

The final product water from the plant will conform to the specifications for water that may be discharged to river as set in the draft amendment to the regulations produced by the Department of Water and Forestry (DWAF) and the Gauteng Department of Agriculture, Conservation and Environment (GDACE).

2.12.3 Summary of the R&D work

(a) Early test work

A number of laboratory and other small-scale tests were carried out in 1999 to test the application of the various treatment options on raw leachate. These are summarised in Table 2.18 together with the conclusions.

The general conclusions from this early test work were :

- Evaporation / crystallisation is the best technology for the first step in the process. It does not, however, on its own, produce water of the required quality. Further treatment of the condensate produced by the evaporator / crystalliser is required.

- Almost no biological activity occurs in raw leachate due to the toxic nature of the leachate. To obtain significant biological activity requires substantial dilution of the feed.

Table 2.18: Various treatment options to treat the Holfontein leachate

Technology Supplier	Technology	Conclusion
WREN	Evaporation / crystallisation	Substantial reduction in COD and dissolved salts
Air Products	Biological treatment with oxygen enhancement	No COD reduction
Afrox	Reduction of COD by oxygen / ozone	Inadequate COD reduction Costly No reduction in dissolved salts
Amitek	Biological treatment - liquid live organisms	Inadequate COD reduction No reduction in dissolved salts
Rone	Catalytic ozonation	Inadequate COD reduction No reduction in dissolved salts
FPO	Reduction of COD by enzymatic oxidation plus chemical treatment	Inadequate reduction in COD and dissolved salts Full process not tested
Alpha Biotek	Biological treatment (activated sludge)	Inadequate reduction in COD and dissolved salts
Induclean	Filter Media	Moderate reduction in COD Costly Some reduction in dissolved salts
CSIR	Electrodialysis preceded by treatment with ash	Inadequate reduction in COD and dissolved salts. Some membrane fouling.

(Note: The ED process is proposed by some manufacturers in Japan for the desalination of leachates (Urase, 2001)).

(b) Pilot plant work

An evaporator / crystalliser pilot plant was hired from The WREN Group and put into operation in October 1999. This plant consisted of a steam heater, an evaporator / crystalliser vessel, a circulating pump, a water-cooled condenser and associated control system. Steam was provided by a small package boiler and cooling water by a small package cooling tower. The pilot plant was rated at about 120 kg/h of condensate. Leachate is moderate corrosive and various materials of construction, including titanium tubes for the steam heater, were incorporated into the pilot plant to test their suitability.

A number of trials were carried out on the pilot plant spanning the period October 1999 to June 2001.

The first was to prove, in concept, that an evaporator / crystalliser plant was able to reduce COD levels in the leachate substantially. This showed that with leachate containing 80 000 mg/ℓ COD, the condensate would contain about 14 000 mg/ℓ COD. Dissolved salts in the leachate of 100 000 mg/ℓ were reduced to about 300 mg/ℓ.

In the second trial, the effect of raising the pH of the leachate by adding caustic soda was examined. This showed that by raising the pH to 10,5 the COD in the

condensate would be reduced to about 6 000 mg/ℓ COD. If the pH was raised even further - to 12,5 - the COD concentration came down to about 1 300 mg/ℓ.

Two biological treatment pilot plants were taken into operation in January 2001. The first consisted of an anaerobic treatment plant and the second of an aerobic treatment plant. Both plants were rated for 200 ℓ/d of feed. The two plants were run in parallel on the condensate from the evaporator / crystalliser pilot plant produced with a feed pH of 10,5. Almost no COD reduction was found in the anaerobic plant despite numerous changes to the operating conditions. The aerobic plant, in contrast, reduced the COD content of the condensate from 3 000 mg/ℓ to about 400 mg/ℓ. A 1:1 dilution with fresh water was used in this experiment. In practice, product water would be used to dilute the feed condensate. The product from the aerobic plant was then treated with activated carbon in a fixed bed column. This test is ongoing and analytical results are awaited.

A set of activated carbon columns were installed on the evaporator / crystalliser condensate stream in May 2001. Operation of these columns on condensate from the evaporator / crystalliser running with pH 12,5 leachate showed that they were able to reduce the COD of the condensate to about the required level of 65 mg/ℓ.

Running the evaporator / crystalliser pilot plant at elevated pH levels produced condensate with high (1 000 mg/ℓ) levels of dissolved ammonia. This ammonia was not removed adequately in either the activated carbon columns or the aerobic biological treatment plant. A number of treatment options to reduce the ammonia are being tested. The first is distillation of the final condensate after the activated carbon columns in option 1. The second is an air stripping of the condensate prior to biological treatment. Other options such as air stripping of the leachate feed to the evaporator / crystalliser plant have been investigated but are not sufficiently effective.

While running the pilot plants, environmental analyses were done on the air around the plant. These showed that, provided the pH of the leachate was kept above about 10, the level of contaminants in the air from the plant were within acceptable limits.

2.12.4 Process description for the full-scale plant

At this stage of the development of the project, both processing options are still being considered. A final decision will be made once more detailed estimates have been obtained of the capital and operating costs and the risks and uncertainties are better defined.

The treatment options are as follows : -.

(a) Option 1 : Evaporator / crystalliser plus activated carbon

In this option the leachate is treated with caustic soda to raise the pH to 12,5 before being fed to the evaporator / crystalliser. The condensate from the evaporator /

crystalliser then passes through a series of activated carbon columns and then to an ammonia removal column. The concentrate from the evaporator / crystalliser is disposed of to landfill.

Steam for the evaporator / crystalliser is provided by a coal fired boiler.

(b) Option 2 : Evaporator / crystalliser plus aerobic biological treatment plus activated carbon

Like option 1, the leachate is treated with caustic soda to raise the pH before being fed to the evaporator / crystalliser. In this case the pH is 10,5. The condensate passes through an air stripping process to reduce the ammonia level. The condensate is then treated in an aerobic treatment plant. The final water is polished with activated carbon. The concentrate from the evaporator / crystalliser is disposed of to landfill.

Steam for the evaporator / crystalliser is provided by a coal fired boiler.

2.12.5 Schedule

Once the pilot plant work is complete, the capital and operating costs of each of the two options will be determined in more detail. A decision will then be made as to which of the two options will be implemented on the full-scale plant.

The next step will be to seek approval from the authorities for the installation of the plant as well as to seek approval from Enviroserv Waste Management to proceed with the project. These activities are expected to take until the end of 2001 to complete.

Once approval to proceed with the project has been obtained, the necessary engineering, procurement and construction activities can start. These are expected to take about 12 months, so that commissioning of the plant is expected late 2002, early 2003.

2.12.6 Pilot plant studies: evaporator/crystalliser

One of the most successful options for treatment of the leachate that could achieve significant reduction in both the salt and organic content was evaporation utilising an evaporator/crystalliser. The product obtained is water plus a concentrate containing the salts and a large part of the organic content. During this process, volatile components can also be transferred into the condensate, i.e.

- *High pH*:
  At pH levels above 11, most of the ammonia-N in the leachate is present as the volatile ammonia and significant amounts transfer with the condensate. On the other hand, cyanide and the volatile fatty acids, e.g. acetic acid, are present as their sodium salts and therefore, do not tend to transfer with the water vapour.

- *Low pH*:
  At acidic pH values cyanide is present in the leachate as the volatile HCN and the fatty acids are in their acid form, whereas ammonia is largely present as ammonium salts. Hydrogen cyanide and the volatile fatty acids appear in the condensate.

- *Phenols:*
  Phenols, tend to steam distil, both at acidic and basic pH values and therefore will occur in the condensate. Slightly lower amounts would be anticipated at high pH.

An evaporatory/crystalliser pilot plant was put into operation in October 1999. The plant consisted of a steam heater, an evaporator/crystalliser vessel, a circulating pump, a water-cooled condenser and the associated control system.

Although the reduction in COD obtained by the evaporator/crystalliser is significant, the presence of small quantities of phenol in the condensate means that a polishing step would be required, if the condensate water is to be discharged to watercourse or utilised on site. Two biological treatment plants were tested from January 2001 - one aerobic and the second anaerobic. The anaerobic plant was found to be ineffective in further reducing the COD, whereas the aerobic plant reduced the COD to 400 mg/ℓ . Aerobic oxidation of phenols tends to be more rapid than anaerobic and is a standard process for phenol removal from water streams.

The average experimental results are shown in Tables 2.19 to 2.22

Table 2.19 : Untreated leachate

Component	Units	Feed Leachate	Concentrate	Condensate
pH		7,8	7	9,2
Conductivity	mS/m	8 500		1 100
TDS	mg/ℓ	130 000	754 000	280
Ammonia as N	mg/ℓ	1 610	480	2 130
Chloride	mg/ℓ	31 200	45 200	260
Sulphate	mg/ℓ	19 300	32 000	91
Calcium	mg/ℓ	890	650	2
Magnesium	mg/ℓ	710	5 140	1
Potassium	mg/ℓ	7 860	33 800	1
Sodium	mg/ℓ	32 300	72 200	27
Iron	mg/ℓ	73	1 1500	0,19
COD	mg/ℓ	78 000	230 000	14 000

Table 2.20 : pH of treated leachate : 10,5

	Units	Raw Leachate	Treated Leachate	Concentrate	Condensate
pH		8,2	10,6	10,7	9,9
COD	mg/ℓ	61 300	59 900	202 000	7 200
Electrical Conductivity	mS/m	>20 000	>20 000	>20 000	240
Ammonia as N	mg/ℓ	1 440	700	600	740
Nitrate as N	mg/ℓ	10	11	26	<1
Phosphate as P	mg/ℓ	14	3,5	25	<0,3
Sodium	mg/ℓ	20 700	22 800	128 000	7

Table 2.21 : pH of treated leachate : 12,5

	Units	Raw Leachate	Treated Leachate	Concentrate	Condensate
pH		8,4	12,8	12,8	10,5
COD	mg/ℓ	63 500	67 400	256 000	1 500
Ammonia as N	mg/ℓ	1 000	600		700
Phosphate as P	mg/ℓ		30	130	2
Phenol	mg/ℓ	2 400	2 600	20 000	1 000

Table 2.22 : Condensate feed and condensate product

Condensate feed	Units	28 Feb - 14 Mar 01	23 Apr - 29 May 01	30 May - 14 Sep 01
pH		9,1	7,4	7,2
Unfiltered COD	mg/ℓ	2 260	2 530	2 330
Filtered COD	mg/ℓ	2 180	2 470	2 210
Suspended Solids	mg/ℓ	40	30	40
Cyanide	mg/ℓ	0,007	0,017	NM
Phenol	mg/ℓ	730	620	720
Dissolved Oxygen	mg/ℓ	1,4	NM	NM
Fats, oils & greases	mg/ℓ	96	60	NM
Ammonia	mg/ℓ	170	460	300
Condensate product	mg/ℓ			
pH	mg/ℓ	8,5	7,5	7,3
Unfiltered COD	mg/ℓ	710	820	450
Filtered COD	mg/ℓ	670	780	400
Suspended Solids	mg/ℓ	125	50	120
Ammonia as N	mg/ℓ	130	440	290
Phenol	mg/ℓ	380	90	<5
Fats, oils & greases	mg/ℓ	90	70	NM
Dissolved Oxygen	mg/ℓ	0,2	NM	NM
Volatile fatty acids	mg/ℓ as AA	75	100	NM
Alkalinity	mg/ℓ as CC	1 370	540	NM
COD reduction	%	70%	70%	80%

Abbreviations: NM: Not measured AA: Acetic acid
CC: Calcium carbonate MLSS: Mix liquid suspended solids

2.12.7 Summary

Two process routes have been developed to convert leachate to water meeting river quality standards. Each of the process routes have been tested at pilot scale and, apart from ammonia removal, have shown that water of the required quality can be produced.

The next step is to select which of the two process routes should be implemented on full-scale and then to seek approval to proceed with the project.

It is believed that the approach followed from the initial desk top work to the laboratory scale work and final piloting of the work on site should result in a long term sustainable solution for the treatment of the leachate. The project has now been submitted for approval to the authorities.

3. EXPERIMENTAL

3.1 Characterisation of Leachates

All chemical and biological analysis on leachate samples were conducted with well established procedures.

3.2 Biodegradability of the Industrial Solid Waste Leachate

3.2.1 Respirometer

A Micro-Oxymax Respirometer (Columbus Instruments) was used to monitor the biodegradability of the waste and treatment products. The automated machine can run 10 samples simultaneously using the same oxygen and carbon dioxide gas sensors. The sensors are calibrated before each run and a comprehensive self-testing programme is performed to verify sensor calibration and check for leaks.

Activated sludge collected from the Daspoort Sewage Works in Pretoria was used for this study. Due to the high COD values of the feed it was decided to perform the test on the full strength and 5x dilutions of the samples in case the measurements of the undiluted samples would fall outside the range of the sensors. The samples were incubated at 20 °C in 250 mℓ test chambers with continuous shaking for 5 days. The oxygen and carbon dioxide levels were measured at 2-h intervals in the chamber headspace and recorded in a data file. The data file was then imported into Excel for analysis.

The reaction mixture included the following components : -

- 45 g sample,
- 2,5 g activated sludge; and
- 100 mg/ℓ N and 110 mg/ℓ P final concentration.

In the negative control the sample was replaced with 0,9% NaCl and in the positive control with 1 g/ℓ glucose.

The following samples were used in the studies : -

- Effluent, Holfontein (received 08/08/2001)
- Untreated feed (19/10/2000, run 4)
- Treated feed (8/12/2000)
- Product (12/12/2000, run 5).

3.2.2 Bio-enhanced treatment of the industrial waste leachate

The treatability study was performed in a sequential batch reactor (SBR) unit (activated sludge with aeration) (Figure 3.1) controlled by a process logic controller (PLC), with a

process volume of 900 litre. Initially the tank was filled with 80% potable water and only about 20% leachate, which allowed for a gradual acclimatisation of the biomass of the leachate. In order to establish a microbial population in the process a 100 litre activated sludge sample obtained from Daspoort, was added to inoculate the reactor. The plant was fed with a catalyst at 0,8 ℓ/h. Phosphate was added in the form of a phosphate fertilizer at approximately 300 mℓ/d.

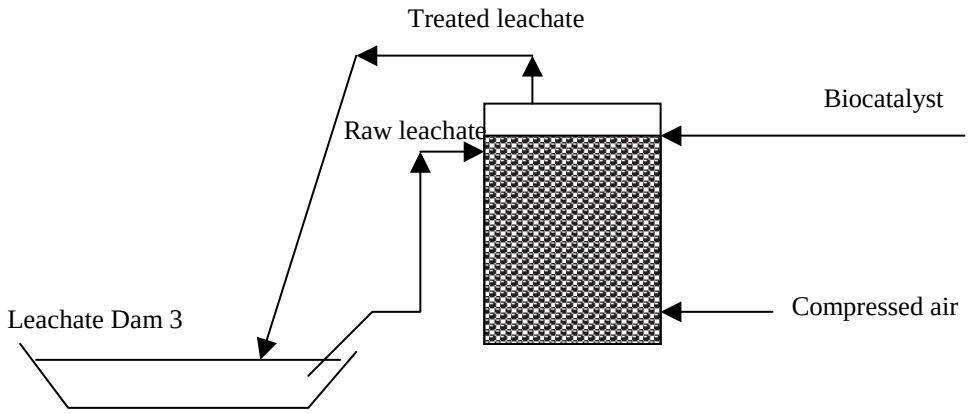


Figure 3.1 : Biocatalyst plant set-up

The PLC was programmed for a treatment cycle of 4 days whereafter the aerator was stopped to allow the activated sludge to settle for 1 hour. After the settling period, 20% of the treated effluent was pumped out and fresh leachate was pumped into the unit and the cycle was then repeated. Samples were taken for COD, pH and conductivity measurements.

3.3 Ash Treatment of the Industrial Solid Waste Leachate for Organics and Inorganics Removal

Leachate samples received during December 1998 and January 1999 were subjected to ash ("Produkstof" and "Oondstof") pretreatment. The December 1998 leachate sample was treated with 'Produkstof", while the January 1999 leachate sample was treated with "Produkstof" and "Oondstof". Leachate samples were also treated with Iscor, Kelvin and Impala and then with Iscor, Lethabo and Sasol ashes.

Ash (200, 400 and 800 g/ℓ) was contacted with the leachate (1 litre) over a 1-hour period while stirring at 100 revolutions per minute. The settled ash volume was measured after 1 and 24 hours. The pH, electrical conductivity, COD, chemical and phenol concentrations were determined before and after treatment.

3.4 Lime, Caustic Soda, Soda-Ash and Ash Pretreatment of the Industrial Solid Waste Leachate for the Removal of Scale-forming Chemicals

The following chemicals were evaluated for the pretreatment of the leachate : -

(a) Lime
(b) Caustic soda
(c) Soda ash

Chemicals were added to 1 litre leachate samples while the samples were stirred at 100 revolutions per minute for one hour after the addition of the chemicals. Sludge volumes were measured after a 30-minute settling period. The pH of the clarified samples was adjusted to approximately 7 with the addition of carbon dioxide. The pH, electrical conductivity, COD and chemical concentration of the leachate were determined before and after pretreatment.

3.5 Coagulation/Flocculation Pretreatment of the Industrial Solid Waste Leachate for the Removal of Suspended Material and Organics

The following flocculants were evaluated for the treatment of the leachate : -

(a) Aluminium sulphate (alum)
(b) Ferric chloride
(c) Poly-aluminium chloride
(d) Magnafloc 1797.

Different flocculant dosages were added to 1 litre leachate samples while rapid mixed at 100 revolutions per minute for one minute. The samples were then slowly mixed at 30 revolutions per minute for approximately 30 minutes and then allowed to settle. (Note: the pH was adjusted to approximately 7 with caustic soda). The electrical conductivity, pH, COD, phenol and suspended solids concentrations of the supernatant were measured. Treated samples were also filtered through Whatman No. 1 filter paper and the COD and suspended solids concentrations were measured. The pH of the leachate sample in the case of ferric chloride was reduced to approximately 4,5 with sulphuric acid to determine the effect of a lower pH on the removal of suspended material.

3.6 Nanofiltration and Ultrafiltration of the Industrial Solid Waste Leachate

The following nanofiltration (NF) and ultrafiltration (UF) membranes were evaluated : -

(a) AFC 40 (PCI-Products)
(b) MPT31 (Koch Membrane Systems)
(c) MPT36 (Koch Membrane Systems)
(d) Polysulphone (UF) (Membratek)

Leachate (12 litre) was circulated through the tubular nanofiltration membranes (0,0261 m^2 area) in the batch mode at feed inlet pressures of approximately 4 000 kPa (Figure 3.2).

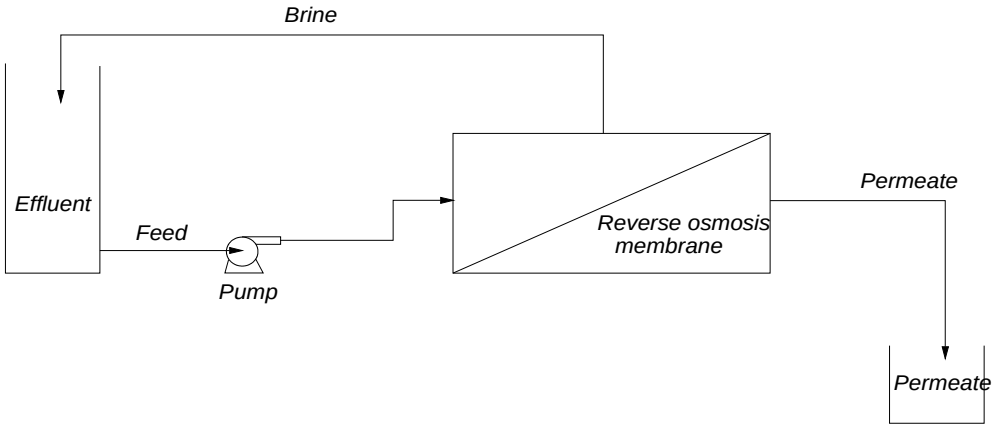


Figure 3.2 : Experimental set-up for batch treatment of the ISWL with NF and UF membranes.

Leachate was circulated through the tubular UF polysulphone membranes (0,261 m^2) in the batch mode at feed inlet pressures of 200 kPa. Permeate was withdrawn and the feed was allowed to concentrate. The run was terminated at a water recovery of approximately 60%.

The following feed inlet pressures were used : -

4 300 kPa for AFC40
3 700 kPa for MPT31
3 700 kPa for MPT36
200 kPa for Polysulphone membranes.

3.7 Determination of the Fouling Potential of the Industrial Solid Waste Leachate for Electrodialysis Membranes

The fouling potential of the leachate for the following membranes were evaluated : -

(a) Selemion AMV and CMV (Asahi Glass)
(b) Tokuyama Soda ACS and CMS membranes
(c) Ionics AR204SZRA and CR67-HMA-412 membranes
(d) Selemion ASV and CSV membranes (Asahi Glass)
(e) Tokuyama Soda AFN and Selemion CMV membranes
(f) Tokuyama Soda AFN and CMX membranes
(g) Tokuyama Soda AXE and CMX membranes

Leachate (5 litre) was circulated (feed 1 ℓ/min; brine 0,5 ℓ/min) through a specially designed ED membrane fouling test cell (membrane area 0,5077 cm^2) under a current density of

20 mA/cm^2 and the voltage drop across the membranes was measured as a function of time (Figure 3.3). An increase in voltage drop across the anion membranes indicates membrane fouling. Membrane resistance before and after ED was also determined.

The fouling potential at different current densities was also evaluated as well as the fouling potential of the most suitable membrane for treatment of the leachate over an extended period of time.

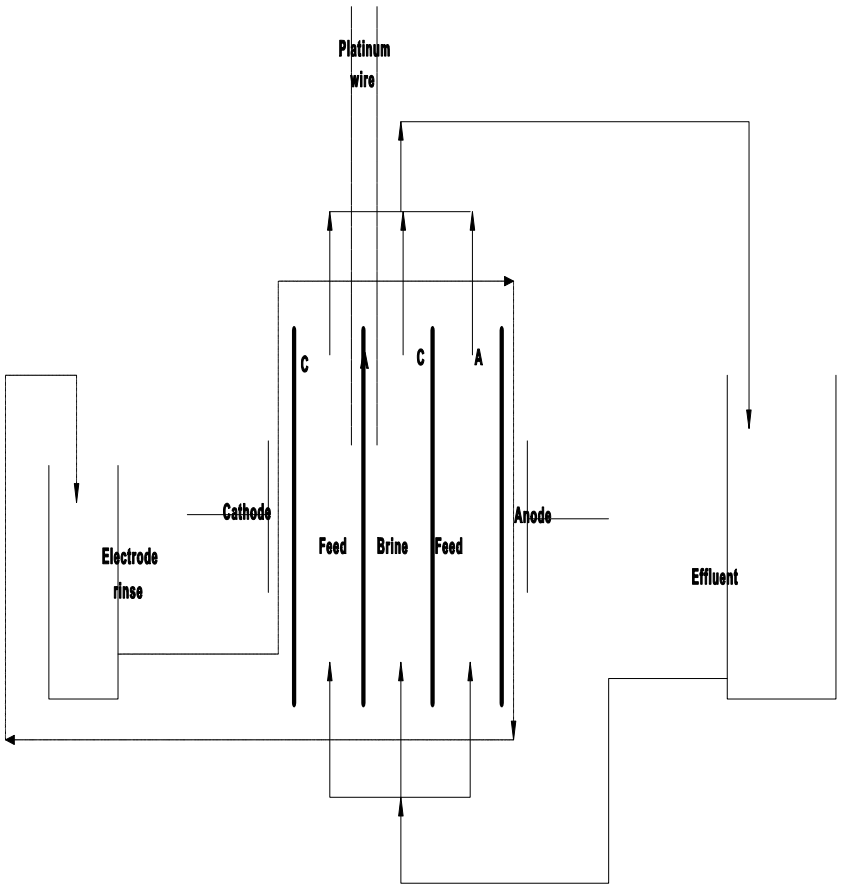


Figure 3.3 : Experimental set-up for ED fouling tests (C: cation; A: anion).

3.8 Evaluation of Membrane Cleaning Strategies of Fouled ED Membranes with Commercially Available Membrane Cleaning Agents .

The anionic membranes were cleaned with 3% NaCl at pH 10,5 (overnight). The membranes were also mechanically and electrically cleaned (polarity reversal).

3.9 Evaluation of ED on Batch Scale for the Desalination/Concentration of the Industrial Solid Waste Leachate

Electrodialysis was conducted in the batch mode using 4 litres pretreated feed (200 g/ℓ Iscor ash, 12 g/ℓ soda ash, 6,9 g/ℓ NaOH, pH approximately 7 with CO_2) and 1 litre brine (pretreated feed) (Figure 3.4). Feed (1 400 mℓ/min) and brine (1 250 mℓ/min) were circulated (linear flow velocity 5,49 cm/s) through the membrane stack (10 cell pairs, 81 cm^2 effective area per membrane) at a cell pair voltage of 0,5 (cell pair voltage measured with platinum wire across 9 cell pairs). One normal sodium chloride in a carbon slurry was used as electrode rinse. The electrical current and the electrical conductivity of the ED feed, ED product and ED brine were measured as a function of time. The initial and final feed and brine volumes were recorded.

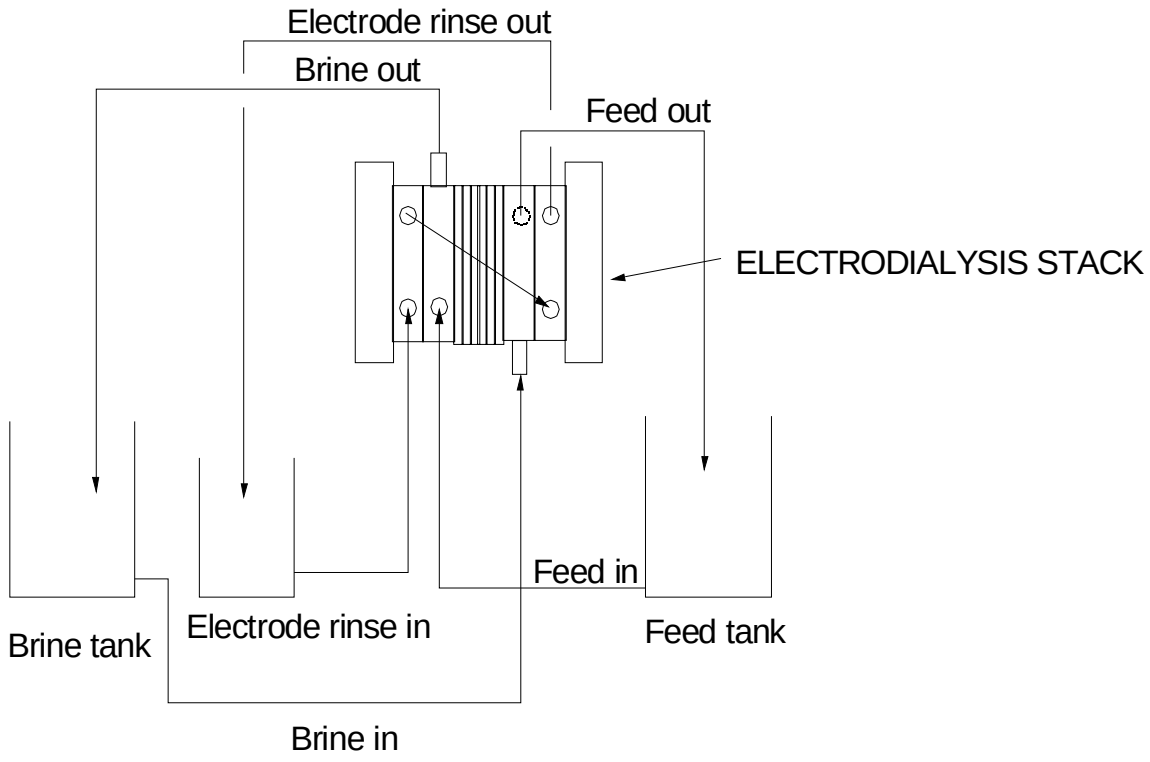


Figure 3.4: Laboratory-scale ED unit operated in the batch mode.

A total of 10 runs were conducted on the pretreated feed. The electrical current, electrical conductivity, cell pair resistance and current density were plotted as a function of time.

Electrodialysis with salt solution (3 000 mg/ℓ NaCl and a sodium chloride solution simulating the concentration of the pretreated feed) was conducted before, during and after the ED runs to determine current efficiency.

Membrane characteristics like membrane resistance, membrane permselectivity, ion-exchange capacity and gel water content were conducted before and after the runs to evaluate membrane fouling.

Electrodialysis runs were also conducted at low pH (approximately 4) and high pH (approximately 10) to determine the effect of pH on ED performance. The chemical composition of the raw leachate, pretreated leachate, ED product and brine was also determined. Polarisation curves were established to determine membrane polarisation characteristics. This was done by changing the voltage and measuring the current and voltage and plotting the current at a function of voltage. The inflection point on the plot line indicates the polarisation point (where water splitting occurs). The toxicity of the raw leachate, pretreated leachate, ED product and brine was also determined (see 3.10).

(a) Current efficiency

Current efficiency was determined from batch ED runs by measuring the initial feed and final desalinated feed volumes and determining the salt equivalents (gram equivalents, ge) removed during electrodialysis. Electrodialysis was conducted with sodium chloride solutions with concentrations of approximately 3 000 mg/ℓ, 88,2 and 94 mS/m, and with pretreated leachate.

The current efficiency was calculated as follows from batch ED tests : -

$$\text{Current efficiency} = \frac{96\,500 \text{ Coulombs / ge x ge removed}}{\text{Coulombs x cell pairs}}$$

Either the sodium removed (in case of NaCl) or the cations or anions removed can be used in the calculation.

(b) Membrane resistance

Membrane resistance was measured between platinum electrodes coated with platinum black in a specially designed membrane resistance measurement cell with a resistance meter. A salt concentration of 0,5 mol/ℓ sodium chloride was used. Membrane resistance was expressed in ohm.cm^2.

(c) Permselectivity

The difference between the counter- and co-ion transport number, Δt, which is called the apparent transport number or membrane permselectivity, was measured as follows (Schoeman, 1992) : -

The potential ($\Delta\Psi m$) of a membrane is usually measured between 0,1/0,2 mol/ℓ or 0,5/1,0 mol/ℓ sodium chloride solution in a specially designed cell with calomel electrodes. The theoretical potential, $\Delta\Psi i$, is calculated from the activities of the two solutions.

Membrane permselectivity, Δt , can then be calculated from these values where $\Delta\Psi m$ is the measured potential and A_s"/A_s' is the ratio of salt activities on both sides of the membrane.

$$\Delta t = \frac{\Delta\Psi m}{\Delta\Psi i}$$

Where $\Delta t = 2t_i - 1$ and

$$\Delta\Psi i = \frac{RT}{F} \ln \frac{A_s''}{A_s'}$$

(d) Ion-exchange capacity

Membrane ion-exchange capacity was determined as follows (Simon and Calmon, 1986) : -

Approximately 3 g dried membrane sample (weighed accurately) was equilibrated with 150 mℓ 1 mol/ℓ hydrochloric acid for 16 hours at room temperature. The membrane sample was rinsed free of chloride. The sample was then treated with 200 mℓ 4% sodium carbonate solution for 2 hours, neutralised to below pH 8,3 with 0,1 mol/ℓ sulphuric acid, potassium chromate (2 mℓ) added and the sample titrated with standardised 0,1 mol/ℓ silver nitrate and the total anion membrane exchange capacity calculated.

(e) Gel water content

The gel water content of the membranes was determined as follows (Simon & Calmon, 1986) : -

Membrane samples (pretreated to their reference form) were blotted dry with filter paper and the mass recorded. The membrane sample was then dried at 165 °C for 16 hours and the dried mass recorded. The gel water content (%) was calculated from the mass loss

3.10 Toxicity Tests

(a) Sample information

The waste (Sample 1) used in the study was collected from Dam 4 at the Holfontein Waste Site. The waste was pre-treated (Sample 2) by adding Iscor ash (200 g/ℓ), soda ash (12 g/ℓ) and caustic soda (7 g/ℓ) to the waste and by bubbling carbon dioxide through the mixture. The pre-treated sample was desalinated (Sample 3) using electrodialysis. During the process, brine (Sample 4) is generated which will, during full-scale operation, be returned to the landfill. All four samples were very dark in colour and had a distinct creosote-like odour.

Moderately hard reconstituted water was used for control testing and dilution of the waste samples (Table 3.1) (Slabbert *et al.*, 1998).

(b) *Daphnia* toxicity test

Toxicity was established by means of a 48-h *Daphnia pulex* lethality test. The test has been recommended for water and effluent toxicity testing in South Africa and was carried out according to procedures established at the Environmentek laboratories and described in the **Guidelines for Toxicity Bioassaying of Waters and Effluents in South Africa** (Slabbert *et al.*, 1998).

Organisms 24 h or less in age were used for toxicity testing. To obtain the necessary number of young for a test, adult females bearing embryos in their brood pouches were removed from the stock cultures 24 h preceding the initiation of a test and placed in beakers containing moderately hard water (Table 3.1) and food suspension (trout chow, alfalfa and yeast). Test conditions are summarized in Table 3.2. Test organisms were transferred to a small intermediary holding beaker and from there to the test beakers.

Table 3.1 : Moderately hard reconstituted water[1]

Reagent added[2] (mg/ℓ)	$NaHCO_3$	96,0
	$CaSO_4.2H_2O$	60,0
	$MgSO_4$	60,0
	KCl	4,0
Nominal water quality range	pH	7,4 - 7,8
	Hardness[3]	80 - 100
	Alkalinity	60 - 70

[1]US EPA (1985) [2]Prepared with Milli-Q water [3]As mg/ℓ $CaCO_3$

Table 3.2 : *Daphnia* test conditions[1].

Temperature	20 ± 1 °C
Light quality	Laboratory illumination
Photoperiod	Approximately 14 h day light
Feeding regime	No feeding
Oxygen concentration	As obtained
pH	As obtained
Size of test beaker	50 mℓ
Volume of test sample	25 mℓ
Number of organisms/beaker	5
Number of replicate beakers	4
Total number of organisms/test	20
Test duration	48 h
Effect measured	Lethality (no movement of body or appendages on gentle prodding)
Interpretation of results	Lethality ≥10% is an indication of toxicity, provided that control lethality <10%

[1]According to US EPA (1985) procedure

(c) Statistical Analysis

A probit computerized statistical programme (US EPA, 1985) was applied to test data to calculate the *Daphnia* LC_{50} (concentration at which 50% of the organisms died) and 95% confidence limits, and the LC_{10} (minimum effect concentration). The LC_0 (no effect concentration) was derived from the concentrations tested.

3.11 Evaluation of ED on Pilot Scale for the Treatment of the Industrial Solid Waste Leachate

Pilot studies were conducted in an ED pilot plant (Figure 3.5). The ED stack contains 75 cell pairs with an active area of 210 cm^2 per membrane. AFN (anionic) and CMX (cationic) membranes from Tokuyama Soda were used for the pilot study. The hazardous waste leachate was pretreated as before, and the supernatant was passed through cross flow polyester membranes for the removal of suspended material prior to ED.

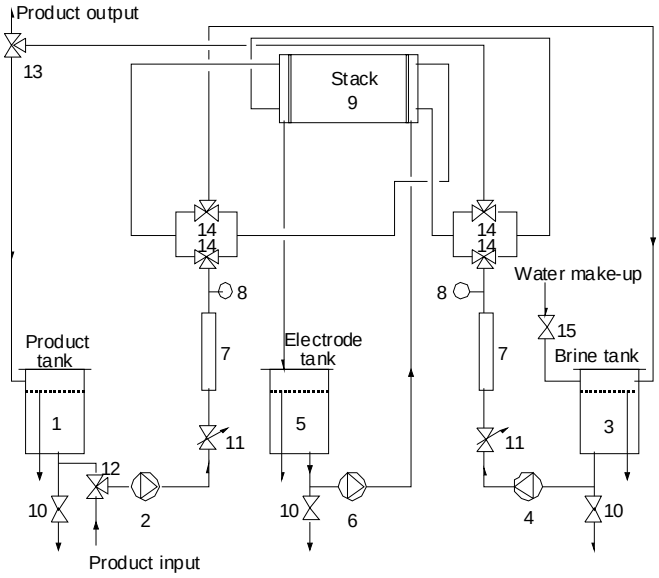


Figure 3.5 : Electrodialysis pilot plant

1:	50 ℓ Polyethelene tank;	9:	Stack with 75 cell pairs;
2:	Product recirculation pump;	10:	Discharge valves;
3:	Brine recirculation tank;	11:	Adjusting valves;
4:	Brine recirculation pump;	12:	Raw product input;
5:	Electrode recirculation tank;	13:	Treated product output;
6:	Electrodes	14:	Reversal valves;
7:	Flow meters;	15:	Make-up valve.
8:	Nanometers;		

Current efficiencies were determined before and after the runs

Electrodialysis was started with 40 litre feed in the feed tank and 10 litre brine (treated feed) in the brine tank. One normal sodium chloride solution was used as electrode rinse solution.

The ED feed and brine solutions were circulated through the ED stack at flow rates of approximately 300 ℓ/h. Electrodialysis was conducted in the batch mode at constant voltage. The electrical current and the electrical conductivity of the ED feed, product and brine were measured as a function of time. The current, conductivity, resistance and current density were plotted as a function of time. The chemical composition of the ED feed, product and brine was also determined. Further desalination of the ED product was investigated with RO.

Electrodialysis runs were also conducted in the feed-and-bleed mode of operation (Figure 3.6) to develop process design criteria for a full-scale application. Electrodialysis was first conducted on the ISWL (1st stage), whereafter ED was conducted on different dilutions (different stages) of the original ISWL to simulate what would happen in a full-scale application. The TDS, loading rate, current efficiency, water transfer and electrical energy consumption were determined after the different ED stages.

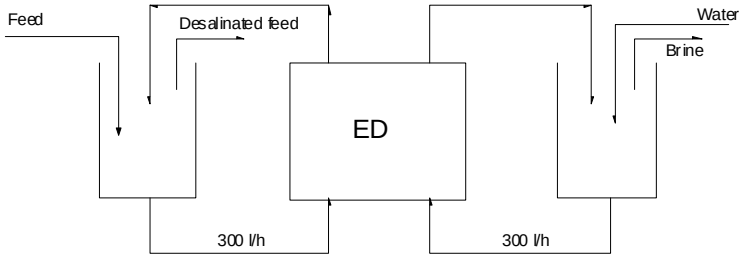


Figure 3.6 : Simplified flow diagram of the feed-and-bleed ED system.

(a) Current efficiency

Membrane current efficiency was determined before the runs were started with 5 g/ℓ sodium chloride feed. The current efficiency was determined as follows : -

$$\text{Current efficiency} = \frac{26{,}8 \times Q \times N}{nI}$$

Where Q : Product flow rate (m^3/h)
N : Salt quantity removed (ge/m^3)
I : Applied current (amperes)
n : Number of cell pairs (75)

The current efficiency was also determined at the end of the runs and after membrane cleaning.

(b) Leachability

The EPA's Standard Method 1311 for the toxicity characteristic leaching procedure (TCLP) was used. This method is described in the 'Minimum Requirements for the Handling, Classification and Disposal of Hazardous Waste, 2nd Edition, 1998, Department of Water Affairs and Forestry.

EPA extraction method 3510B was used to extract the sludge sample into methylene chloride and EPA analytical method 8270C was used to determine the contaminants present in the extract by means of a Varion Saturn 2000 Ion Trap Chromatograph/Mass Spectrometer.

3.12 Determination of the Fouling Potential of the Municipal Solid Waste Leachate for Tubular Cellulose Acetate and Polyamide RO Membranes

The experimental set-up for the determination of the fouling potential of the MSWL for the membranes was similar to the set-up as described under 3.6. (Figure 3.2).

(a) Cellulose acetate membranes (Membratek)

The pH of 12 litre leachate sample was adjusted to pH 6,5 with sulphuric acid. Reverse osmosis was conducted in the batch mode (membrane area 0,0261 m^2) at a feed inlet pressure of 4 000 kPa. The clean water flux (CWF) was measured in the beginning and at the end of the run. The run was terminated at a water recovery of approximately 60%. Permeate flux and water recovery were measured as a function of time.

The membranes were cleaned with 1% sodium tripolyphosphate/0,5% EDTA (pH10,6 with NaOH).

(b) PCI membranes (PCI Memtech)

The same experiment as described above was conducted with PCI AFC 99 membranes (polyamide), membrane area 0,0261 m^2. No pH adjustment of the feed was applied in this case. The membranes were cleaned with 0,2% sodium laurel sulphate/0,1% EDTA (pH 11,7 with NaOH). Permeate flux and water recovery were measured as a function of time.

3.13 Evaluation of Tubular RO for the Desalination/Concentration of Municipal Solid Waste Leachate

(a) Cellulose acetate RO membranes (Membratek)

Approximately 100 litre MSWL (pH adjusted to 6,5 with H_2SO_4) was treated in the batch mode (4 000 kPa inlet pressure) in a RO pilot plant (membrane area 1,75 m^2,

1 module) (see 3.6, Figure 3.2 for batch mode). Reverse osmosis was terminated at a water recovery of approximately 70%. The CWF was measured before and after the run. Permeate flux was measured as a function of water recovery. The chemical composition of the RO feed, product and brine was also determined. (NOTE: Sponge ball cleaning and flow reversal used.)

(b) PCI AFC99 membranes (PCI Memtech)

The same experiment as described under 3.13(a) was conducted with the AFC 99 polyamide RO membranes. No pH adjustment of the RO feed was conducted in this case. The membrane area was 0,81 m^2 (1 module).

3.14 Evaluation of RO on Pilot-Scale for the Treatment of the Municipal Solid Waste Leachate

(a) Feed and bleed RO tests

Reverse osmosis pilot studies were conducted at Bisasar Road Landfill Site in Durban. Reverse osmosis runs were conducted in the feed-and-bleed mode of operation (see Figure 3.7). The membrane area for the cellulose acetate and polyamide membranes (PCI-AFC99) were 1,75 and 0,81 m^2, respectively (one module each). The pH of the feed water to the cellulose acetate and the polyamide membranes were adjusted with hydrochloric acid to pH 6,2 - 6,5 and 7 - 7,2 respectively. An antiscalent, Flocon 260 (12,5 mg/ℓ) was dosed during RO treatment of the leachate with the cellulose acetate membranes while Permatreat 391 (11 mg/ℓ) was dosed during the treatment of the leachate with the polyamide membranes. The water recovery was kept at approximately 70%. The CWF was determined at the start of the runs and then once a day after a water rinse for 30 minutes. The CWF was also determined before and after membrane cleaning. Cleaning of the cellulose acetate membranes was conducted with nitric acid, sodium-tripoly phosphate (STP)/EDTA, citric acid, Ultrasil 50, EDTA/sodium laurel sulphate (SLS) and phosphoric acid solutions. The polyamide membranes were cleaned with hydrochloric acid and Utrasil 10. Sponge ball cleaning (30 min) with flow reversal was used. The RO product flux was measured as a function of time. The chemical composition of the RO feed, product and brine was also determined.

(b) Continuous RO tests

Continuous or once through RO tests were conducted on the leachate with a Grahamtek Systems RO pilot plant (tfc spiral wound sea water membranes model tfc (8" x 40") (2,78 m^2)). Pretreatment of the leachate consisted out of sand filtration (sand and anthracite) and 5 micron cartridge filtration prior to RO treatment. The experimental set-up is shown in Figure 3.8.

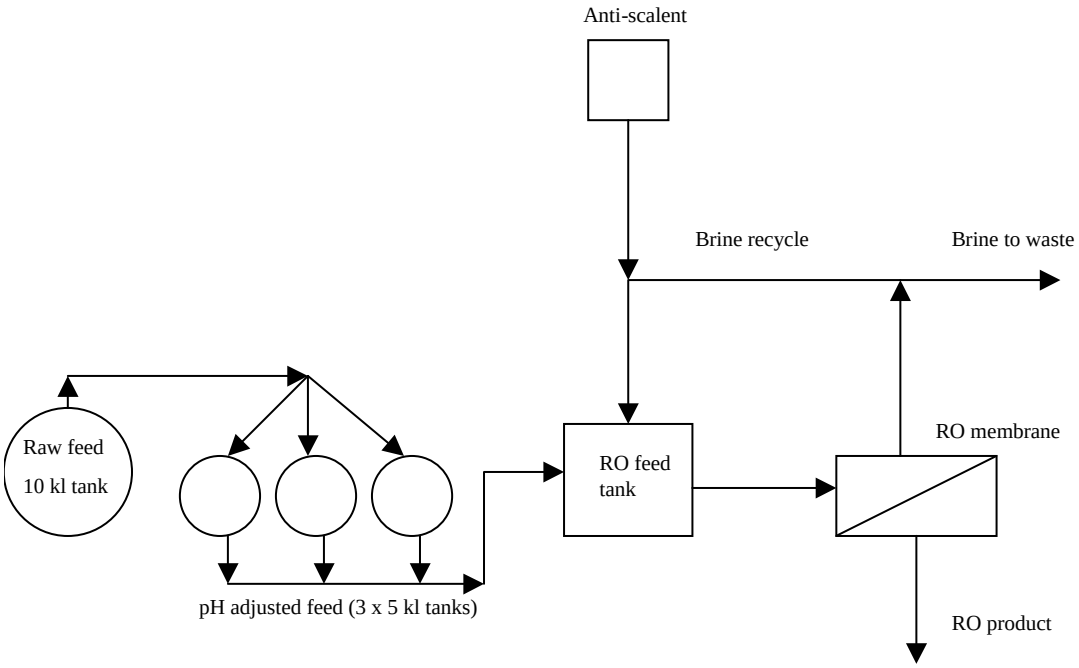


Figure 3.7 : Feed-and-bleed RO experimental set-up.

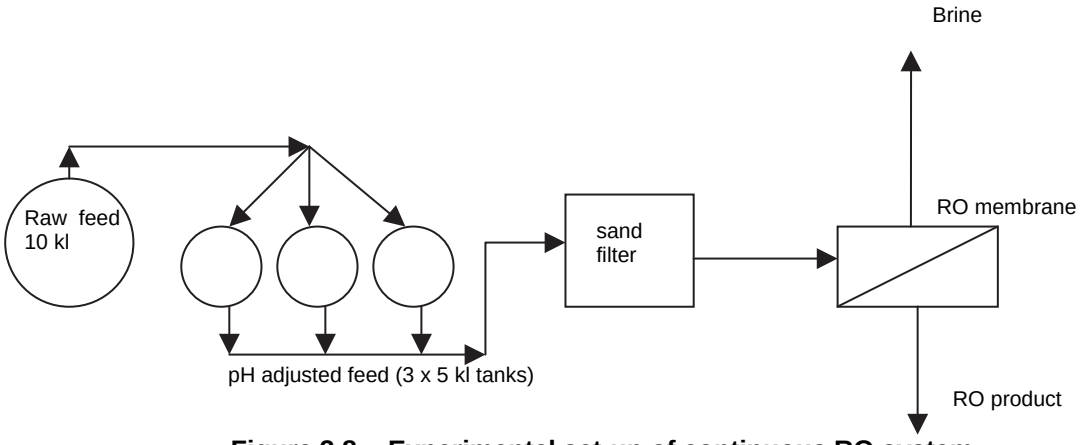


Figure 3.8 : Experimental set-up of continuous RO system.

Reverse osmosis was conducted at a feed inlet pressure of 4 000 kPa. Water recovery was initially approximate 54%. The permeate flux was measured as a function of time. Clean water fluxes were conducted at 3 000 kPa. The membranes were cleaned with acid (HCl pH 2) and Ultrasil 10 (0,25%) to evaluate membrane cleaning on RO performance**.**

4. CHARACTERISATION OF THE INDUSTRIAL SOLID WASTE LEACHATE

The history of the quality of the Holfontein leachate over an eight-year period is shown in Appendix A. Some of these results are summarised in Figures 4.1 to 4.13.

4.1 Electrical Conductivity

The electrical conductivity of the leachate showed an increase from about 11 000 mS/m during November 1992, to approximately 25 000 mS/m during January 1996 (Figure 4.1). A sharp decrease in the electrical conductivity of the leachate was then experienced, and the electrical conductivity remained at between approximately 7 000 (August 1997) and 9 000 mS/m until September 1999. The heavy rains during early 2000 decreased the electrical conductivity to approximately 3 700 mS/m.

4.2 pH

The pH of the leachate varied between approximately 6,7 and 8 from November 1992 to February 2000 (Figure 4.2).

4.3 Chloride and Sulphate

The chloride concentration increased from 10 000 mg/ℓ in November 1992 to approximately 51 000 mg/ℓ in January 1997 (Figure 4.3). It then decreased and was 30 000 mg/ℓ during November 1997 and varied between 23 000 and 37 000 mg/ℓ for the period December 1997 to October 1999. The chloride concentration was significantly lower during February 2000, as a result of the heavy rains that fell in the country.

The sulphate concentration was 10 000 mg/ℓ in November 1992, increased to approximately 32 000 mg/ℓ during September 1995, decreased and remained between approximately 15 000 and 27 000 mg/ℓ from July 1997 to June 1999 (Figure 4.3). It then decreased and was approximately 5 000 mg/ℓ during February 2000.

4.4 Nitrate-Nitrogen

There was almost no nitrate-nitrogen in the leachate from November 1992 to September 1995 (Figure 4.4). The nitrate-nitrogen concentration then increased somewhat, decreased again and reached a high during May 1998. It then decreased rapidly and was almost zero during February 2000.

4.5 Ammonia-Nitrogen

The ammonia-nitrogen concentration was 570 mg/ℓ during November 1992, increased and decreased and reached a high of 3 700 mg/ℓ during July 1999 (Figure 4.5). The ammonia-nitrogen concentration then decreased to almost zero during September 1999 and was approximately 250 mg/ℓ during February 2000.

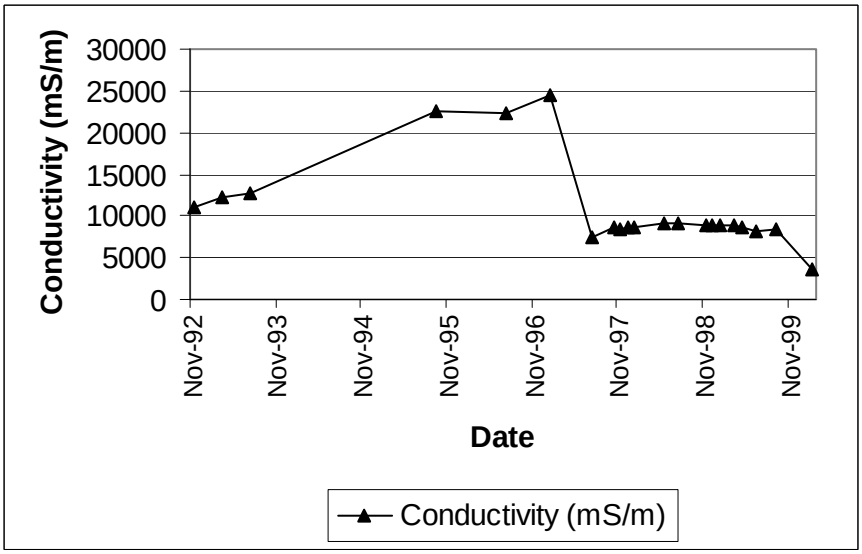


Figure 4.1 : Electrical conductivity of the ISWL over a period of time.

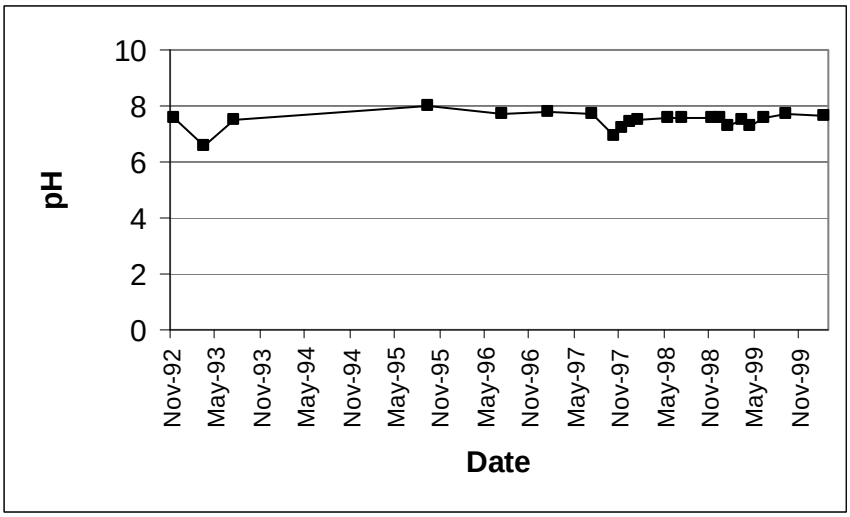


Figure 4.2 : pH of the ISWL over a period of time.

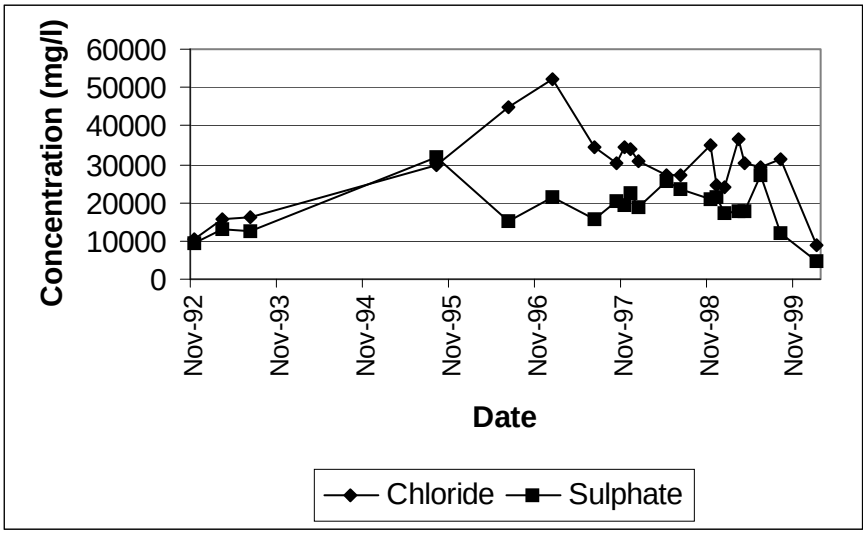


Figure 4.3 : Chloride and sulphate concentrations of the ISWL over a period of time.

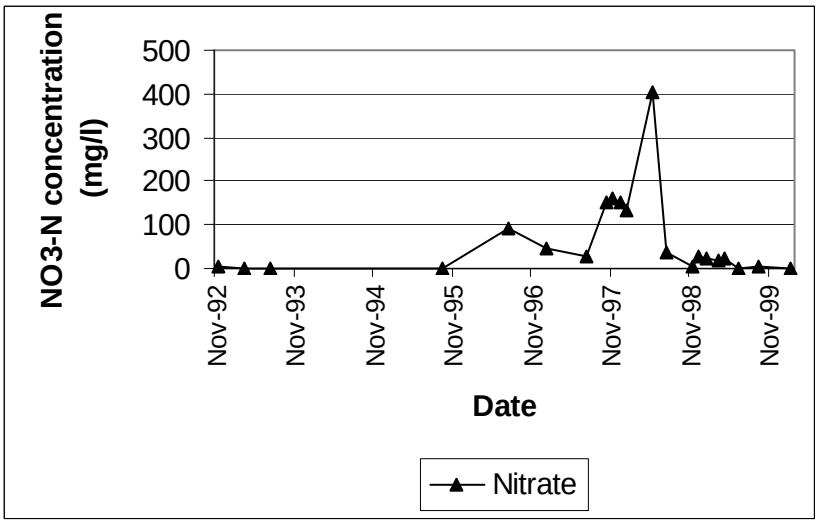


Figure 4.4 : Nitrate-nitrogen concentration of the ISWL over a period of time.

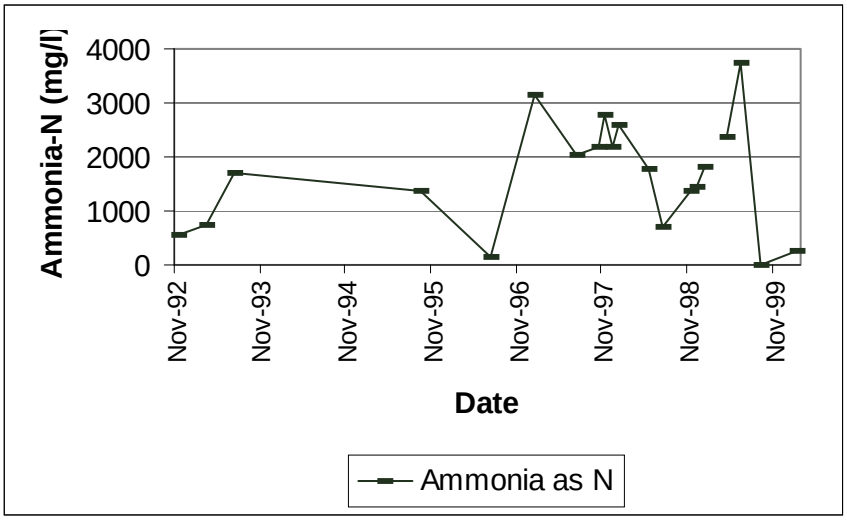


Figure 4. 5 : Ammonia-nitrogen concentration of the ISWL over a period of time.

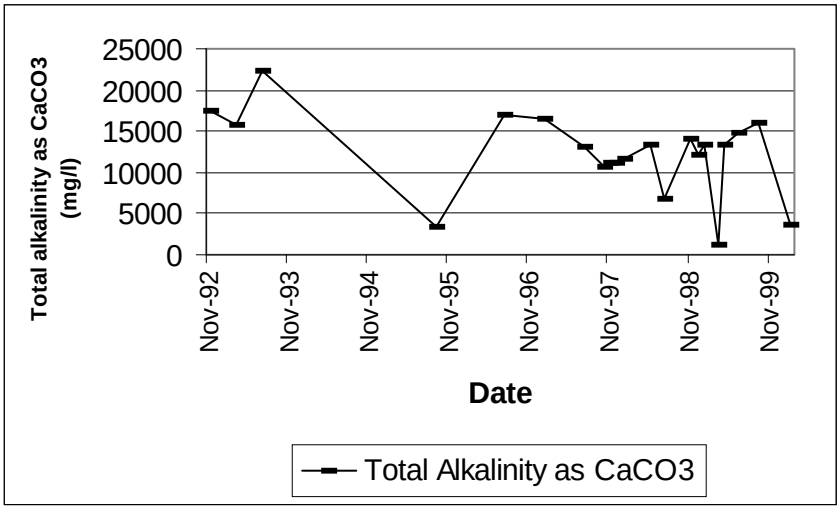


Figure 4. 6 : Total alkalinity of the ISWL over a period of time.

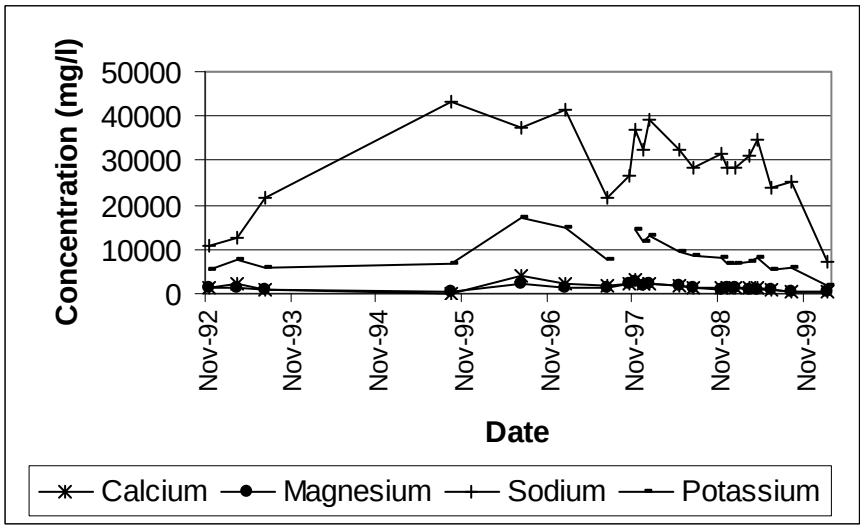


Figure 4.7 : Calcium, magnesium, sodium and potassium concentrations of the ISWL over a period of time.

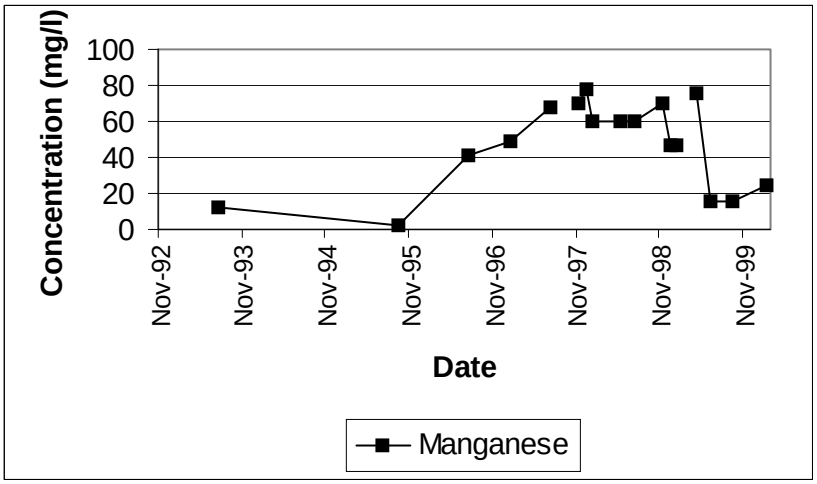


Figure 4.8 : Manganese concentration of the ISWL over a period of time.

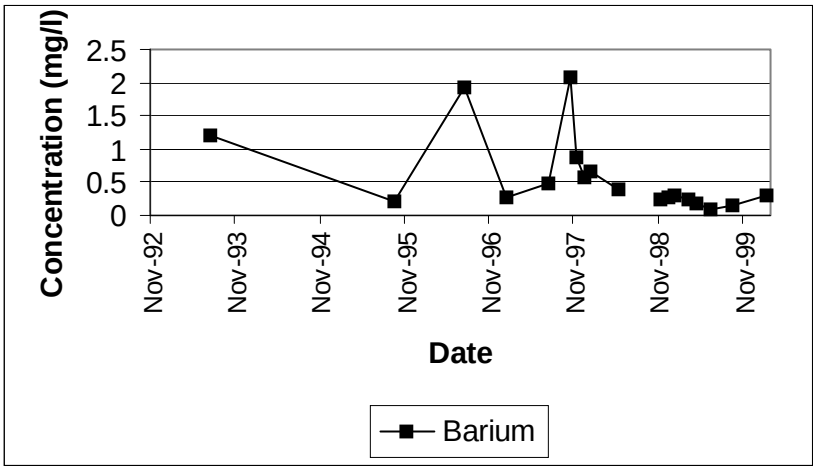


Figure 4.9 : Barium concentration of the ISWL over a period of time.

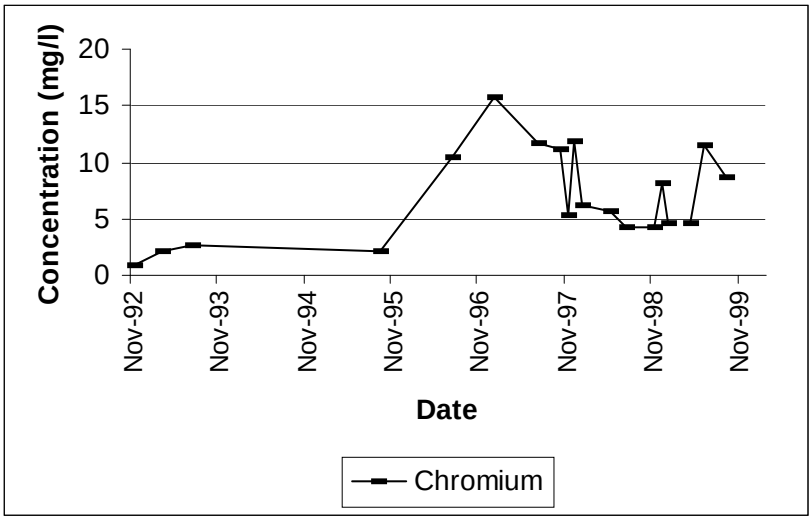


Figure 4.10 : Chromium concentration of the ISWL over a period of time.

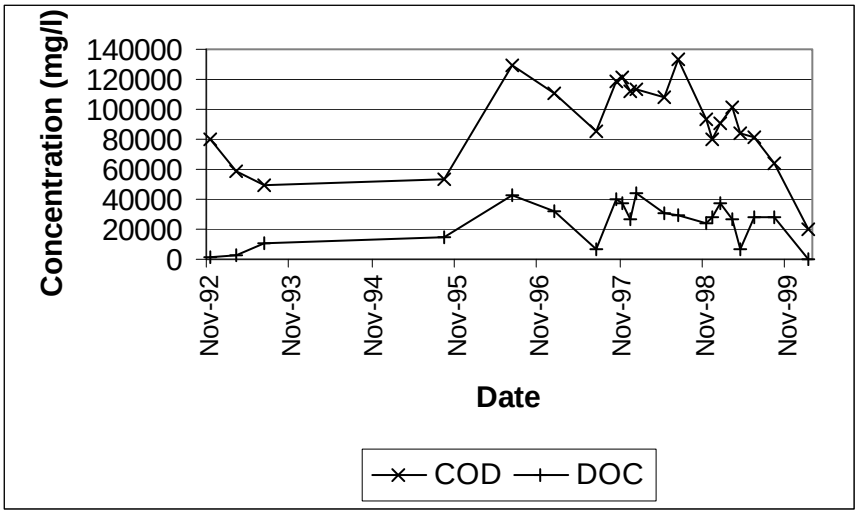


Figure 4.11 : COD and DOC concentrations of the ISWL over a period of time.

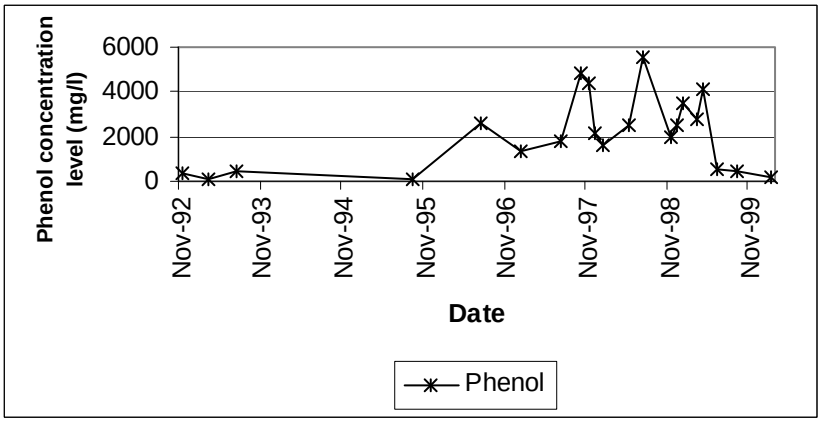


Figure 4.12 : Phenol concentrations of the ISWL over a period of time.

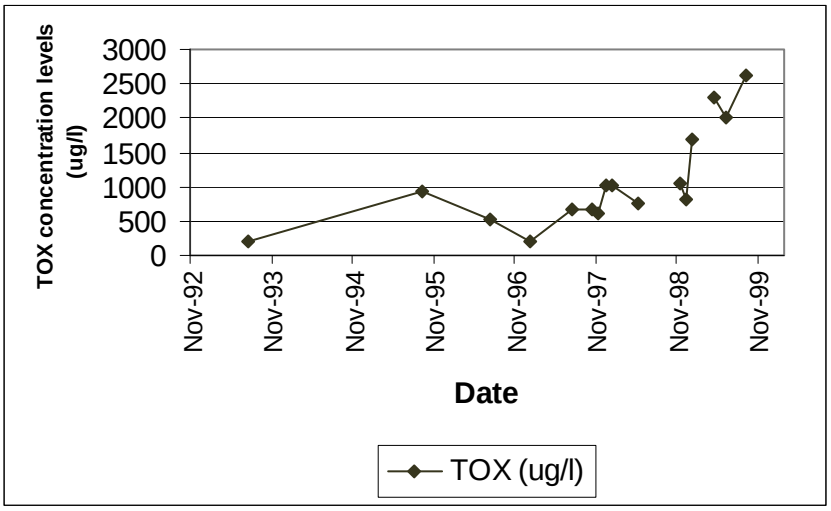


Figure 4.13 : TOX concentration of the ISWL over a period of time.

4.6 Total Alkalinity

The total alkalinity was 17 500 mg/ℓ during November 1992, and the maximum and minimum concentrations over the period until February 2000 were 22 300 and 1 300 mg/ℓ, respectively (Figure 4.6). The total alkalinity was approximately 4 000 mg/ℓ during February 2000.

4.7 Calcium, Magnesium, Sodium and Potassium

The sodium concentration is high (Figure 4.7). The sodium concentration was approximately 11 000 mg/ℓ during November 1992, and it appeared that it reached a high during September 1995. It then decreased somewhat and increased and remained between 21 000 and 40 000 mg/ℓ until November 1999. Very much the same pattern was observed with the chloride concentration. The sodium concentration was only 7 230 mg/l during February 2000.

The potassium concentration over the period was much lower (Figure 4.7). It was approximately 5 500 mg/ℓ during November 1992, increased to a maximum of approximately 17 000 mg/ℓ, and was approximately 1 000 mg/ℓ during February 2000.

The calcium and magnesium concentrations were significantly lower than the sodium and potassium concentrations and were almost the same (Figure 4.7). The calcium and magnesium concentrations were approximately 1 000 mg/ℓ during November 1992, and increased to a maximum of approximately 4 000 (July 1996) and 2 800 mg/ℓ (November 1997), respectively, and decreased to approximately 500 mg/ℓ during February 2000.

4.8 Manganese

The minimum and maximum manganese concentrations over the period from July 1993 (12,5 mg/ℓ) to February 2000 were 2,22 and 77,9 mg/ℓ, respectively (Figure 4.8).

4.9 Barium

The barium concentration during July 1993 was 1,2 mg/ℓ and the minimum and maximum concentrations until February 2000 were 0,1 and 2 mg/ℓ, respectively (Figure 4.9).

4.10 Chromium

The total chromium concentration varied between 0,9 and 15,8 mg/ℓ from November 1992 to September 1999 (Figure 4.10).

4.11 COD and DOC

The COD concentration of the leachate is very high (Figure 4.11). It was 80 000 mg/ℓ during November 1992, decreased and it appeared that it reached a maximum of approximately 133 000 mg/ℓ during July 1998. It then decreased, varied up and down, and the lowest concentration (20 000 mg/ℓ) was obtained during February 2000.

The DOC concentrations over the same period varied between a minimum of approximately 1 800 mg/ℓ and a maximum of approximately 44 000 mg/ℓ.

4.12 Phenols

The phenol concentration of the leachate is also high (Figure 4.12). The phenol concentration during November 1992 was approximately 338 mg/ℓ with minimum and maximum concentrations of 50 and 5 500 mg/ℓ, respectively, until February 2000.

4.13 TOX

The TOX concentration of the leachate is high (Figure 4.13). The minimum and maximum concentrations from July 1993 to October 1999 were approximately 200 and 2 600 µg/ℓ, respectively.

4.14 Quality of the ISWL during the Study Period

A typical example of the quality of the leachate during the study period is shown in Table 4.1.

A cationic and an anionic balance of the major cations and anions present in the leachate, as well as the percentage composition of the major cations and anions in the leachate, are shown in Table 4.2. The major cations in the leachate are sodium (67%), ammonium-

nitrogen (12,5%), potassium (11,0%) and the major anions are chloride (56,4%), sulphate (31,4%) and bicarbonate (12,3%). However, significant quantities of magnesium and calcium are also present.

Table 4.1 : Chemical composition of the Holfontein leachate (Dam 4, aeration dam, Date - 02/06/2000)

Constituents*	Concentration
Conductivity (mS/cm)	87,4
pH	6,99
Chemical oxygen demand (COD)	64 000
BOD_5 as O_2	16 500
BOD_5/COD ratio	0,2578
Total organic Carbon	19 303
Alkalinity as $CaCO_3$	8 971
Total suspended solids	3 778
Total dissolved solids	105 580
Ammonia Nitrogen as N	2 199
Nitrate as N	0,8
Chloride	29 239
Sulphate	21 999
Fluoride	3,2
Cyanide (Total)	75
Silicon as Si	13,3
Calcium	983
Magnesium	858
Sodium	19 400
Potassium	5 440
Barium	0,32
Strontium	3,83
Chromium (Total)	5,27
Copper	0,04
Iron	95,5
Lead	1,63
Manganese	29,1
Nickel	2,1
Zinc	7,04
Arsenic (Inorganic)	7,36
Mercury (Total)	0,005
Phenolic compounds	900
Volatile Fatty Acids	6 578

*Concentration in mg/ℓ , unless stated otherwise. Cd = 0,12 mg/ℓ (23/03/2001)

The organic concentration of the leachate is also high (64 000 mg/ℓ COD) (Table 4.1). It is interesting to note that BOD/COD ratio is low (0,26). Therefore, it will be difficult to biologically degrade the organic material in the leachate. It is also interesting to note that the phenol concentration of the leachate is high (900 mg/ℓ). The relatively high volatile fatty acid content of the leachate (approximately 6 600 mg/ℓ) shows that some of the organic material should be readily biodegradable. Biological biodegradability, however, of the leachate could be inhibited by toxic compounds in the leachate. The concentrations of toxic heavy metals like chromium, lead, nickel, arsenic are reasonably high

Table 4.2 : Cationic and anionic balance and % composition of the major cations and anions present in the strong leachate.

Cations			Anions		
Type	**me***	**%**	**Type**	**me***	**%**
Na	843,5	67,0	Cl	823,6	56,4
NH_4-N	157,1	12,5	SO_4	458,3	31,4
K	139,1	11,0	HCO_3	179,4	12,3
Mg	70,3	5,6			
Ca	49,2	3,9			
	1 259,2	100,0		1 461,3	100,0

* me: milli equivalents ion

5. BIODEGRADABILITY OF THE INDUSTRIAL SOLID WASTE LEACHATE

5.1 Respirometer

The biodegradability of the ISWL from Holfontein, electrodialysis of the pretreated feed, ED product and brine in the presence and absence of biosupplements, are summarised in Table 5.1. The detailed experimental conditions, experimental procedure (see 3.2.1) and results are shown in Appendix B.

The percentage COD removals calculated from the analytically and respirometrically determined COD values, respectively, do not correspond (Table 5.1). The reason for this can probably be ascribed to analytical errors. However, the results should be considered as a qualitative trend that could be expected during biological treatment of the leachate. (NOTE: The calculated COD removals from the oxygen consumption is considered to be more reliable because of the sensitivity of the instrument).

As expected, almost no COD was removed from the leachate (0,23%) in the presence of inactivated sludge (Table 5.1). However, in the presence of activated sludge a small percentage of COD (7,3 to 8%) was removed from the leachate and treated (ash and chemicals) leachate. A significant amount of COD was also removed from the ED product (30%). However, very little COD was removed from the ED brine (0,81%). This is to be expected as a result of the high salt concentration in the ED brine inhibiting the biomass in the activated sludge.

The leachate and ED treated leachate were diluted five times to decrease the toxicity to the activated sludge and the biodegradability tests were repeated. Significant COD removals were obtained on the diluted leachate (22,9% inactivated sludge), diluted leachate (34,9%), diluted treated leachate (40,4%) (26,0% from COD), diluted ED product (34,2%) (54,1% from COD) and brine (34,7%). Therefore, dilution of the leachate improves its biodegradability.

Some COD removals were obtained on the leachate after the addition of biosupplements to the leachate. However, the increase in COD removal after the addition of different biosupplements were not very significant (from 5,4 to 7,8% to 8%, COD calculated from oxygen consumption). However, more significant COD removals were observed with the analytically determined COD values (from 1,8 to 4,7 to 30%). Therefore, it appears from these results that biosupplements should improve the removal of COD from the leachate.

Biodegradability tests on the ED product have also shown that the biosupplements should improve COD removal (from 24,8 to 31,5 to 46% COD, calculated from oxygen consumption) from the ED product.

Table 5.1 : Biodegradability of the leachate, ED treated leachate and samples to which biosupplements had been added.

Leachate Samples	Vol (mℓ)	COD start mg/ℓ	COD end mg/ℓ	COD Removed[1] (%)	COD removed (mg/ℓ)	COD Removed (Calc)[2] (mg/ℓ)	COD Removed (Calc) (%)
Leachate + inactivated sludge	45	57 700	64 250	-11,35	-6 550	133	0,23
Leachate[3]	45	57 700	56 000	2,95	1 700	4 578	7,93
Treated leachate[3]	45	41 550	46 100	-10,95	-4 550	3 044	7,33
ED product[3]	45	20 150	19 350	3,97	800	6 044	30,00
ED brine	45	55 150	51 750	6,17	3 400	444	0,81
Leachate + inactivated sludge (5 x)	45	11 540	7 350	36,31	4 190	2 644	22,91
Diluted leachate (5 x)	45	11 540	8 100	29,81	3 440	4 022	34,85
Diluted treated leachate (5 x)	45	8 310	6 150	25,99	2 160	3 356	40,38
Diluted ED product (5 x)	45	4 030	1 850	54,09	2 180	1 378	34,19
Diluted ED brine (5 x)	45	11 030	9 000	18,40	2 030	3 822	34,65
Biosupplements							
Leachate	40	59 700	58 600	1,84	1 100	3 200	5,36
Leachate + 1004 TX	40	67 950	64 750	4,71	3 200	5 325	7,84
Leachate + 1002 CG	40	69 050	51 600	25,27	17 450	5 400	7,82
Leachate + 1003 FG	40	60 000	42 000	30,00	18 000	4 825	8,04
ED product	40	23 000	17 000	26,09	6 000	5 700	24,78
ED product + 1004 TX	40	29 350	23 300	20,61	6 050	9 250	31,52
ED product + 1002 CG	40	27 400	22 250	18,80	5 150	10 300	37,59
ED product + 1003 FG	40	21 450	22 900	-6,76	-1 450	9 775	45,57

(1) COD removed calculated from analytical results

(2) COD removed calculated from cumulative oxygen consumption (Respirometer)

(3) See 12.2, Table 12.2 (run 4).

The commencement of the biological activity and the percentage COD removals of the different leachate samples (undiluted, diluted, ED product, ED feed, treated feed and brine, with and without supplements), are shown in Tables 5.2(a) and 5.2(b). The biological activity starts first with the ED product and the commencement of the biological activity was delayed with increasing salt concentration (ED product to ED brine) (Table 5.2(a)). The percentage COD removal decreases with increasing activity start time. The biological activity started at the same time (14 h) after dilution of the leachate and the percentage COD removal was somewhat higher (34 to 40%) than on the undiluted leachate samples. The addition of biological supplements decreased the commencement of biological activity and increased COD removal (Table 5.2(b)).

Some of the oxygen consumption and carbon dioxide production results from the respirometer results (Table 5.1) are shown in Appendix B.

Table 5.2a: The effect of dilution on the biodegradability of the leachate samples (feed; pretreated feed, ED product, ED brine).

Sample	Undiluted samples		Diluted samples (5X)	
	Activity start (h)	%COD removal (%)	Activity start (h)	%COD removal (%)
ED Product	14	30	14	34
Treated Feed	24	7,3	14	40
Feed	38	7,9	14	34
Brine	68	0,81	20	34
Autoclaved Sludge + Feed	none	0,23		

Table 5.2b : The effect of biosupplements on the biodegradability of the leachate.

Sample	Feed		ED Product	
	Activity start (h)	%COD removal (%)	Activity start (h)	%COD removal (%)
No Supplements	20	5,4	8	24,8
Textile (1004 TX)	13	7,8	5,5	31,5
Tannery (1002 CG)	13	7,8	5,5	37,6
Abattoir (1003 FG)	13	8	5,5	45,6

5.2 Bio-enhanced Treatment of the Industrial Solid Waste Leachate

Tests were also conducted in an SBR unit (activated sludge with aeration cycle) at the Holfontein site to evaluate the biodegradability of the leachate (see 3.2.2). The plant was fed with a catalyst at 0,8 ℓ/h. Phosphate was added in the form of a phosphate fertilizer at approximately 300 mℓ/d. The unit was operated at a 96 hour cycle. Approximately 250 litre product was removed and replaced with fresh feed.

The operational results are shown in Table 5.3. The results have shown that approximately 20 to 45% of the COD could be removed in the reactor.

Table 5. 3 : COD removal during the treatment of the leachate in an SRB unit.

Date	Feed mg/ℓ	Product mg/ℓ	Conductivity reactor (mS/cm)	COD removals %
19/01/2001	Start reactor			
02/02/2001			27,4	
19/02/2001	70 000	48 000	67,2	31,4
27/02/2001	70 000	38 500	67	45,0
23/03/2001	65 000	52 000	84	20,0
Plant not getting phosphate				
5/7/2001	108 000			
31/7/2001	112 000	65 000		

6. EVALUATION OF ASH PRETREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR ORGANICS AND INORGANICS REMOVAL

6.1 Treatment of the Industrial Solid Waste Leachate with "Produkstof" and "Oondstof"

Leachate samples received during December 1998 and January 199 from Holfontein were subjected to ash ("Produkstof" and "Oondstof") pretreatment. The December 1998 leachate sample was treated with "Produkstof" while the January 1999 leachate sample was treated with "Produkstof" and "Oondstof"

Ash (200, 400 and 800 mg/ℓ) was contacted with the leachate (1 litre) over a 1-hour period while stirring at 100 revolutions per minute. The settled ash volume was measured after 1 and 24 hours. The pH, electrical conductivity, COD and phenol concentrations were determined before and after treatment. Supernatant of the ash treated leachate was subjected to further treatment with soda ash (10 g/ℓ), caustic soda (1,5 g/ℓ) and carbon dioxide prior to preliminary ED tests (Schoeman *et al.*, 1999).

The results are shown in Tables 6.1 to 6.6

Table 6.1 : Chemical composition of the supernatant after ash ("Produkstof") treatment of the leachate (1 hour contact time). (Sample received December 1998).

Constituents*	Ash (g/ℓ)			
	0	200	400	800
pH	7,80	8,35	8,73	8,65
Conductivity (mS/cm)	87,5	82,3	71,4	61,1
COD	68 600	66 700 (2,8)	33 300 (51,5)	23 000 (66,5)
Phenols	1 977	1 126 (43,0)	727 (63,2)	180 (90,9)
Ammonia as N		1 094	879	857
Chloride		23 100	21 950	11 790
Sulphate		14 000	16 000	17 000
Chromium		2,54	1,59	1,80
Manganese	217	220	8,5 (96,1)	8,4 (96,1)
Potassium		5 100	4 900	3 800
Sodium		18 900	18 500	14 400
Magnesium		1 140	795	774
Calcium	1 210	1 590	2 880	1 680

* Concentration in mg/ℓ unless otherwise stated.

( ) : % removal

Sludge volumes increased with increasing ash dosage, and sludge volumes were high (Tables 6.2, 6.4, 6.6). Sludge volume, for example, was 66,7% (400 g/ℓ ash) in the case of

"Produkstof" (Table 6.2). This means that the clarified water volume comprised only 33,3% of the total volume. However, sludge volume was lower in the case of "Oondstof". Sludge volume was only 39,6% (400 g/ℓ ash) in this case (60,4% clarified water) (Table 6.6).

Table 6.2 : Sludge volume after ash ("Produkstof") treatment of the leachate (1 hour contact time). (Sample received December 1998).

Ash (g/ℓ)	Leachate (mℓ)	Total volume (mℓ)	Sludge volume (mℓ)	
			After 1 hour	After 24 hours
0	1 000	1 000	-	-
200	1 000	1 000	400	350
400	1 000	1 200	800	800
800	1 000	1 450	1 450	1 200

Table 6.3 : Chemical composition of the supernatant after ash ("Produkstof") treatment of the leachate (1 hour contact time). (Sample received 25 January, 1999).

Constituents*	Ash (g/ℓ)							
	0	200		400		800		
pH	7,25	7,64		7,67		7,51		
Conductivity (mS/cm)	98,7	104,8		105,4		108,8		
COD	79 800	60 900	(23,7)	33 700	(57,8)	28 700	(64,0)	
Phenols	3 374	2 208	(34,6)	1 260	(62,7)	367	(89,1)	
Ammonia as N	1 550	1 912		1 106	(28,6)	924,1	(40,4)	
Chloride	25 240	29 990		20 520	(18,7)	20 750	(17,8)	
Sulphate	13 000	16 000		12 000		18 000		
Chromium	4,7	1,2	(74,5)	3,3	(29,8)	2,4	(48,9)	
Manganese	240	27,9	(88,4)	24,5	(89,8)	17,8	(92,6)	
Potassium	6 200	5 900	(4,8)	5 900	(4,8)	6 000	(3,2)	
Sodium	22 200	21 100	(5,0)	21 700	(2,3)	22 000	(0)	
Magnesium	1 500	1 600		1 620		1 870		
Calcium	1 240	1 250		1 340		1 430		

* Concentration in mg/ℓ unless otherwise stated.
( ) : % removal

Table 6.4 : Sludge volume after ash ("Produkstof") treatment of the leachate (1 hour contact time). (Sample received 25 January 1999).

Ash (g/ℓ)	Leachate (mℓ)	Total volume (mℓ)	Sludge volume (mℓ)	
			After 1 hour	After 24 hours
0	1 000	1 000	-	-
200	1 000	1 000	400	400
400	1 000	1 200	750	750
800	1 000	1 450	1 450	1 200

Table 6.5 : Chemical composition of the supernatant after ash ("Oondstof") treatment of the leachate (1 hour contact time). (Sample received 25 January, 1999).

Constituents*	Ash (g/ℓ)							
	0	200		400		800		
pH	7,25	8,04		8,10		8,14		
Conductivity (mS/cm)	98,7	97,9		96,1		93,0		
COD	79 800	74 000	(7,3)	71 600	(10,3)	64 700	(18,9)	
Phenols	3 374	3 608		3 408		3 048	(9,6)	
Ammonia as N	1 550	1 663		1 648		1 579		
Chloride	2 5240	24 000	(4,9)	23 320	(7,6)	22 650	(10,3)	
Sulphate	13 000	16 000		19 000		18 000		
Chromium	4,7	4,0	(14,9)	3,7	(21,3)	3,1	(34,0)	
Manganese	240	24,9	(89,6)	19,9	(91,7)	16,2	(93,3)	
Potassium	6 200	6 000	(3,2)	5 900	(4,8)	5 300	(14,5)	
Sodium	22 200	21 100	(5,0)	20 900	(5,9)	19 100	(14,0)	
Magnesium	1 500	1 450	(3,3)	1 410	(6,0)	1 250	(16,7)	
Calcium	1 240	1 200	(3,2)	1 270		1 240		

Concentration in mg/ℓ unless otherwise stated.
( ) : % removal

Table 6.6 : Sludge volume after ash ("Oondstof") treatment of the leachate (1 hour contact time). (Sample received 25 January 1999).

Ash (g/ℓ)	Leachate (mℓ)	Total volume (mℓ)	Sludge volume (mℓ)	
			After 1 hour	After 24 hours
0	1 000	1 000	-	-
200	1 000	1 000	225	225
400	1 000	1 200	475	475
800	1 000	1 450	1 250	1 250

The chemical composition of the supernatant after ash (“Produkstof”) treatment of the December 1998 leachate sample (Table 6.1) showed:

(a) That the pH of the supernatant increased with increasing ash dosage.

(b) That the electrical conductivity of the supernatant decreased with increasing ash dosage.

(c) That the COD and phenol concentrations of the supernatant significantly decreased with increasing ash dosage.

(d) That ammonia-nitrogen of the supernatant somewhat decreased with increasing ash dosage.

(e) That significant reductions in the chloride and potassium and sodium concentrations were experienced with increasing ash dosage.

(f) That excellent manganese removal could be obtained.

(g) That the calcium and sulphate concentrations of the supernatant increased with increasing ash dosage.

The salinity (conductivity) reduction with increasing ash dosage may be ascribed to an increasing ion-exchange capacity of the ash at higher pH levels.

A similar trend in results was experienced with the leachate sample received during January 1999 (Table 6.3). In his case, however, the electrical conductivity of the supernatant increased with increasing ash dosage. This showed that the ash had introduced some cations and anions into the supernatant. It is also interesting to note that the initial pH of the leachate was lower in this case than in the previous case (see Table 6.1.)

Much poorer removals of chemicals were obtained when the leachate was treated with "Oondstof" (Table 6.5). The settling characteristics of the treated leachate, however, were better (Table 6.6).

The initial pretreatment of the ISWL from Holfontein with two different ash samples for the removal of organics and inorganics from the leachate has shown that significantly different results can be obtained with different ash samples. Treatment of the leachate with 'Produkstof' gave significantly better results for the removal of inorganics from the leachate. 'Oondstof', on the other hand, gave lower sludge volumes. Therefore, an ash with the most suitable characteristics for pretreatment of the leachate should be selected. The two ashes used, however, were no longer available and it was decided to conduct further test work on ashes that were readily available in large quantities. Ashes from Iscor, Sasol and from power stations were therefore used in further tests.

6.2 Treatment of the Industrial Solid Waste Leachate with Iscor, Kelvin and Impala Ash

A leachate sample obtained from Holfontein during February 2000 was subjected to ash pretreatment. Different ash samples were used, and organics removal (COD and phenols) were determined as a function of ash dosage (experimental procedure the same as in 6.1). The results are summarised in Tables 6.7, 6.8 and 6.9.

Excellent phenol removals were obtained with Iscor ash (Table 6.7). Phenol removals varied between approximately 89 and 96% in the ash dosage range from 100 to 600 g/ℓ. The COD

removals varied from approximately 13 to 40% over the same dosage range. Poorer phenol and COD removals, however, were obtained with Kelvin (Table 6.8) and Impala ash (Table 6.9). It is further interesting to note that the Iscor ash had increased the pH of the sample (Table 6.7), while there was almost no pH change in the case of the other two samples (Tables 6.8 and 6.9). It is also interesting to note that some salinity was removed in the case of the Kelvin (Table 6.8) and the Impala ash (Table 6.9), while no salinity was removed in the case of the Iscor ash (Table 6.7).

Table 6.7 : Treatment of the leachate with Iscor ash for organics removal (February 2000).

Ash (g/ℓ)	pH	Conductivity (mS/m)	Final volume (mℓ)	Sludge volume $(mℓ)_{30\ min}$	COD (mg/ℓ)	Phenol (mg/ℓ)	Removal (%)	
							COD	Phenol
0	7,63	3 660	1 000	-	18 000	339	-	-
100	9,61	3 580	1 000	200	15 700	28,2	12,78	91,68
200	10,61	3 620	1 100	400	12 600	19,8	30,00	94,16
400	12,18	4 170	1 200	650	11 550	36	35,83	89,38
600	12,21	4 330	1 350	900	11 350	13,6	36,94	95,99

Table 6.8 : Treatment of the leachate with Kelvin ash for organics removal (February 2000).

Ash (g/ℓ)	pH	Conductivity (mS/m)	Final volume (mℓ)	Sludge volume $(mℓ)_{30\ min}$	COD (mg/ℓ)	Phenol (mg/ℓ)	Removal (%)	
							COD	Phenol
0	7,63	3 660	1 000	-	18 000	339	-	-
100	7,57	3 600	1 000	50	-	-	-	-
200	7,63	3 600	1 100	100	-	-	-	-
400	7,6	3 520	1 300	300	18 250	142,4	-1,39	57,99
600	7,56	3 470	1 400	500	21 750	95,2	-20,83	71,92

Table 6.9 : Treatment of the leachate with Impala ash for organics removal (February 2000)

Ash (g/ℓ)	pH	Conductivity (mS/m)	Final volume (ℓ)	Sludge volume $(ℓ)_{30\ min}$	COD (mg/ℓ)	Phenol (mg/ℓ)	Removal (%)	
							COD	Phenol
0	7,63	3 660	1 000	-	18 000	339	-	-
100	7,69	3 610	1 000	75	-	-	-	-
200	7,71	3 580	1 100	100	-	-	-	-
400	7,63	3 510	1 300	350	16 000	171,2	11,11	49,50
600	7,44	3 460	1 400	500	20 000	137,6	-11,11	59,41

6.3 Treatment of the Industrial Solid Waste Leachate with Iscor, Lethabo and Sasol Ash

Treatment of the Holfontein leachate with Iscor, Kelvin and Impala ash as described in section 6.2 was conducted on a more dilute leachate as a result of heavy rains during late 1999. Excellent phenol removals were obtained with Iscor ash. Phenol removals varied between approximately 89 and 96% in the ash dosage range from 100 to 600 g/ℓ. COD removals varied from approximately 13 to 40% over the same dosage range. Poorer phenol and COD removals were obtained with the other two ash samples.

Treatment of the leachate with Iscor, Lethabo and Sasol ash (see 3.3) is shown in Tables 6.10, 6.11 and 6.12 (see Appendix C for composition of ashes). The leachate sample, on which the work was conducted, is considered to be more representative of the quality of leachate that can be expected at Holfontein.

Very good phenol (900 mg/ℓ to 58 mg/ℓ, 93,6% removal) removals were obtained with Iscor ash at an ash dosage of 200 g/ℓ (Table 6.10). COD removal, however, was not that good (64 000 mg/ℓ to 54 700 mg/ℓ, 14,5% removal) at the same dosage level. It further appears that the COD removals increase with increasing ash dosage and that the calcium concentration in the treated leachate increases with increasing ash dosage. No significant removals of magnesium, manganese (except at very high ash dosage), barium, strontium and silica were experienced with increasing ash dosage. Some ammonia and iron removals were experienced with increasing ash dosage. Sludge volume also increases with increasing ash dosage and comprises 27% of the treated volume at an ash dosage of 200 g/ℓ.

Poorer phenol removals were obtained with Lethabo ash (Table 6.11) than with Iscor ash (Table 6.10). COD removals were approximately the same. The ash also introduced some calcium into the treated leachate with increasing ash dosage. Almost no magnesium, manganese and iron were removed from the leachate with increasing ash dosage. Some barium, strontium and silica, however, were introduced into the leachate with increasing ash dosage. Some ammonia was removed from the leachate with increasing ash dosage. This can be ascribed to a higher pH caused by a higher ash dosage. Sludge volume also increases with increasing ash dosage.

Poor phenol removals were obtained with Sasol ash (Table 6.12). The COD removals were about the same as with the previous two ash samples. A significant increase in pH of the leachate was experienced at high ash dosage levels with the result that significant removals of magnesium, manganese, ammonia and iron were experienced. However, significant quantities of calcium were introduced into the leachate with increasing ash dosage. Some strontium was also introduced into the leachate, while the barium concentration remained the same.

Table 6.10 : Treatment of the leachate with ISCOR ash for organics removal (leachate sample 2/6/2000).

ISCOR Ash g/ℓ	Total Vol mℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	pH after CO_2	COD (mg/ℓ)	Ca (mg/ℓ)	Mg (mg/ℓ)	Mn (mg/ℓ)	Ba (mg/ℓ)	Sr (mg/ℓ)	Si (mg/ℓ)	Ammonia (mg/ℓ)	Na (mg/ℓ)	Cl (mg/ℓ)	Phenol (mg/ℓ)	Iron (mg/ℓ)
0	1 000			87,4	6,99		64 000	983	858	29,1	0,32	3,83	13,3	2 199	19 400	29 239	900	95,5
50	1 000	50	5	92,9	8,02	7		1 670	1 141									
100	1 000	150	15	93,3	8,39	7	53 600	2 110	1 410								170	
200	1 100	300	27	93,1	8,87	7	54 700	2 470	1 410	34,3	0,88	6,81	20,1	1 480	21 400	32 367	58	91,6
400	1 300	500	38	92,2	9,21	7	39 100	2 760	1 350					1 271				
800	1 600	750	47	91,1	9,4	7	30 300	2 910	949	13,4	0,69	6,74	11,4	1 312	21 800	24 240	60	37,3
800 + 200	1 100	300	27	90,7	9,68	7	26 900	2 980	900					1 192			26	

Table 6.11 : Treatment of the leachate with LETHABO fly ash for organics removal (leachate sample 2/6/2000).

Ash g/ℓ	Total Vol mℓ	Sludge Vol mℓ	% Sludge	EC Ms/cm	pH	pH after CO_2	COD (mg/ℓ)	Ca (mg/ℓ)	Mg (mg/ℓ)	Mn (mg/ℓ)	Ba (mg/ℓ)	Sr (mg/ℓ)	Si (mg/ℓ)	Ammonia (mg/ℓ)	Na (mg/ℓ)	Cl (mg/ℓ)	Phenol (mg/ℓ)	Iron (mg/ℓ)
0	1 000			87,4	6,99		64 000	983	858	29,1	0,32	3,83	13,3	2 199	19 400	29 239	900	91,6
50	1 000	50	5	90,3	7,37	7	44 200	1 220	944									
100	1 000	100	10	90,2	7,6	7	43 800	1 310	923									
200	1 100	200	18	89,5	7,93	7	26 700	1 610	943	34,8	0,59	7,54	12,2	1 782	21 200	30 365	510	114
400	1 200	400	33	87,3	8,13	7	37 300	1 770	898	30,5				1 698				97,4
800	1 500	750	50	85,5	8,39	7	38 500	1 940	894	22,9	0,64	13,8	15	1 505	19 400	28 287	560	82,8

Table 6.12 : Treatment of the leachate with SASOL fly ash for organics removal (leachate sample 2/6/2000).

Ash g/ℓ	Total Vol mℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	pH after CO_2	COD (mg/ℓ)	Ca (mg/ℓ)	Mg (mg/ℓ)	Mn (mg/ℓ)	Ba (mg/ℓ)	Sr (mg/ℓ)	Si (mg/ℓ)	Ammonia (mg/ℓ)	Na (mg/ℓ)	Cl (mg/ℓ)	Phenol (mg/ℓ)	Iron (mg/ℓ)
0	1 000			87,4	6,99		64 000	983	858	29,1	0,32	3,83	13,3	2 199	19 400	29 239	900	91,6
50	1 000	50	5	87,1	8,09	7,5	36 600	1 920	900	36,1								
100	1 000	175	18	86	8,47	7,5	34 700	2 710	885	33,9				1 500			800	102
200	1 100	300	27	81,2	9,06	7,5	32 600	3 810	809	23,4	0,52	8,78		1 018	19 000	26 024		66,3
400	1 200	600	50	84,4	9,41	7,5	34 700	3 910	875	6,86			9,7	692			1 000	
800	1 300	1 000	77	85,2	10,98	7,5	35 300	4 500	9	0,2	0,3	14,6	5,2	365	20 400	30 019	940	11,6

7. EVALUATION OF LIME, CAUSTIC SODA AND SODA-ASH PRETREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR ORGANICS AND INORGANICS REMOVAL

7.1 Treatment of the Industrial Solid Waste Leachate with Chemicals for the Removal of Organics and Inorganics (sample of 1999)

The following chemicals and combinations of chemicals were used in the preliminary evaluation of the pretreatment of the leachate (Schoeman *et al.*, 1999):

(a) Lime
(b) Soda-ash
(c) Caustic soda
(d) Hydrochloric acid
(e) Aluminium sulphate
(f) Polyelectrolytes (T7, TC19S)

Chemicals were added to 1 litre leachate samples while the samples were stirred at 100 rpm. Stirring was continued for 1 hour after the addition of the chemicals. Sludge volumes were measured after a 30-minute settling period. The pH of the clarified samples was adjusted to approximately 7 with the addition of carbon dioxide. The pH, electrical conductivity, COD and calcium concentrations of the leachate were determined before and after treatment.

The results are shown in Table 7.1.

The calcium and manganese concentrations of the leachate are high. These cations, plus other heavy metals, can cause fouling/scaling of ion-exchange membranes unless removed to low levels. Therefore, preliminary work was conducted to determine to what extent these cations could be removed from the leachate, and what dosages of lime, soda-ash and caustic soda would be required to remove these cations from the leachate to low levels.

The results showed that good removals of both calcium and manganese could be obtained (Table 7.1). However, the best calcium removal was obtained at soda-ash and caustic soda dosages of 10 and 3 g/ℓ, respectively. The manganese concentration, however, was not so low (152 mg/ℓ) in this case. Lower manganese concentrations, however, could be obtained, but the calcium concentration was significantly higher.

Table 7. 1 : Pretreatment of the leachate with lime, caustic soda, soda-ash and carbon dioxide.

Experi-ment	Lime g/ℓ	Soda-ash g/ℓ	Caustic soda g/ℓ	HCl g/ℓ	Alum g/ℓ	TC7 and TC19S mℓ/ℓ	pH	Cond. mS/cm	pH After CO_2	mS/cm After CO_2	COD (mg/ℓ)	Mn (mg/ℓ)	Ca (mg/ℓ)	Sludge $(mℓ)_{30\ min}$
Feed	-	-	-				7,87	87,2	-	-	68 600	217	1 210	-
1	32,2	-	-				10,02	81,4	7,00	85,2	62 500 (8,9)	15,1 (93,0)	3 050 (77,8)	200
2	-	64	-				9,31	95,3	7,10	96,5	54 600 (20,4)	63 (71,0)	269 (77,8)	600
3	-	-	6,08				10,00	83,6	6,96	85,4	62 500 (8,9)	41,4 (80,9)	412 (77,8)	800
4*	10	5	-				8,80	86,1	7,03	84,4	-	45 (79,3)	704 (66,0)	50
5*	-	7,5	--				8,50	87,5	6,98	86,7	-	204 (6,0)	705 (41,7)	300
6*	10	10	-				8,80	85,9	6,93	86,3	64 500 (6,0)	91,4 (57,9)	254 (79,0)	375
7*	-	5	2				8,80	88,0	6,98	88,4	-	192 (11,5)	573 (52,6)	375
8*	-	10	2				8,85	88,1	7,05	89,6	-	134 (38,2)	284 (76,5)	600
9*		10	3				9,06	90,0	7,02	90,9	-	152 (30,0)	167 (86,2)	700
10*						4	7,64				67 600 (1,5)			
11*					2		7,19				67 600 (1,5)			
12*				6,75			4,43				60 200 (12,2)			

* Aeration for 1 hour
( ) % removals

It appears from the results that Iscor ash will be the best ash to use for treatment of the leachate, especially when high phenol removals are required.

It is further interesting to note that:

(a) High sludge volumes were produced.
(b) COD removals were relatively low.
(c) Alum, polyelectrolytes and acid did not remove COD to low levels.

It should be pointed out that no attempt was made to optimise the results in the preliminary investigation as shown in Table 7.1. Lower pretreatment costs should result with the addition of optimised chemical dosages to the leachate.

7.2 Treatment of the Industrial Solid Waste Leachate with Chemicals for the Removal of Organics and Inorganics (sample of February 2000)

Leachate received during February 2000 was similarly treated as under 7.1. The results are shown in Tables 7.2, 7.3 and 7.4.

Excellent calcium, magnesium and manganese removals were obtained with caustic soda treatment of the leachate (Table 7.2). However, poor COD and phenol removals were obtained.

Excellent magnesium and manganese removals were obtained with lime treatment of the leachate (Table 7.3). However, a significant amount of calcium is introduced into the leachate at higher lime dosages. This, of course, is not beneficial for membrane desalination processes. Reasonably good phenol removal was obtained at a high lime dosage. However, poor COD removal was obtained.

Excellent calcium and magnesium removals were obtained with soda-ash treatment of the leachate (Table 7.4).

Sludge volumes increased with increasing chemical dosing level and especially high sludge volume was produced at the highest caustic soda dosage level (50%) (Table 7.2).

Table 7.2 : Pretreatment of the leachate with caustic soda.

NaOH (g/ℓ)	pH	Conduc-tivity (mS/m)	Sludge volume $(ℓ)_{30\ min}$	Ca (mg/ℓ)	Mg (mg/ℓ)	Mn (mg/ℓ)	COD (mg/ℓ)	Phenol (mg/ℓ)	Removal (%)	
									COD	Phenol
0	7,63	3 660	-	461	387	24.5	18 000	339	-	-
0,88	9,01	3 690	-	231 (49,9)	355 (8,3)	1,01 (95,9)				
2,08	10,02	3 640	75	48 (89,6)	286 (26,1)	0,3 (98,8)				
3	11	3 630	200	64 (86,1)	126 (67,4)	0,41 (98,3)	18 750	317	-4,17	6,49
4,12	12	3 850	500	56 (87,9)	1 (99,7)	0,11 (99,4)	17 500	237	2,78	30,09

( ) : % removal

Table 7.3 : Pretreatment of the leachate with lime.

Lime (g/ℓ)	pH	Conduc-tivity (mS/m)	Sludge volume $(ℓ)_{30\ min}$	Ca (mg/ℓ)	Mg (mg/ℓ)	Mn (mg/ℓ)	COD (mg/ℓ)	Phenol (mg/ℓ)	Removal (%)	
									COD	Phenol
0	7,63	3 660	-	461	387	24,5	18 000	339	-	-
3,9	8,99	3 630	-	331 (28,2)	315 (18,6)	2,81 (88,5)				
5	10,02	3 610	50	645	332 (14,2)	0,63 (97,4)				
5,52	11	3 620	100	837	162 (58,1)	0,1 (99,6)	20 000	197	-11,11	41,89
6,42	12,02	3 850	125	1120	1 (99,7)	0,09 (99,6)	17 500	131	2,78	61,36

( ) : % removal

Table 7.4 : Pretreatment of the leachate with soda ash.

Soda ash (mg/ℓ)	pH	Conduc-tivity (mS/m)	Sludge volume $(ℓ)_{30\ min}$	Calcium (mg/ℓ)	Magnesium (mg/ℓ)	Removal (%)	
						Calcium	Magnesium
0	7,63	3 660	-	461	387	-	-
1 000	8,02	3 760	20	61	50	86,7	87,1
2 000	8,27	3 800	50	30	35	93,6	91,0
4 000	8,62	3 910	100				

( ) : % removal

7.3 Treatment of the Industrial Solid Waste Leachate with Chemicals for the Removal of Organics and Inorganics (sample June 2000)

Treatment of leachate samples received during 1999 and February 2000 with chemicals were discussed in the previous paragraphs. The results obtained with chemical treatment of a leachate sample received from Holfontein during June 2000, which is considered to be more representative of the quality of the ISWL, is presented in this section.

The treatment of the leachate with caustic soda, lime, soda ash and lime, and caustic soda and soda ash are shown in Tables 7.5, 7.6, 7.7 and 7.8, respectively (see 3.4).

Excellent magnesium (97,1%, 858 to 25 mg/ℓ), manganese (98,5%, 29,1 to 0,45 mg/ℓ), barium (90,6%, 0,32 to 0,03 mg/ℓ) and iron (86,2%, 91,6 to 12,6 mg/ℓ) removals were obtained with caustic soda treatment at high pH (pH 12,02) (Table 7.5). Good strontium removal was also obtained (60,3%, 3,83 to 1,52 mg/ℓ). It further appears that significant amounts of COD (44,7%, 64 000 to 34 500 mg/ℓ) and phenols (57,8%, 900 to 380 mg/ℓ) can also be removed with caustic soda treatment of the leachate. However, a significant amount of sodium was added to the leachate with caustic addition, while almost no calcium was removed.

Lime treatment of the leachate showed that a significant amounts of manganese (99,2%, 29,1 to 0,24 mg/ℓ), barium (90,6%, 0,32 to 0,03 mg/ℓ) and iron (82,9%, 91,6 to 15,7 mg/ℓ) could be removed from the leachate (Table 7.6). Magnesium removal was not as good as with caustic soda treatment, while some silica was also removed. Significant quantities of calcium, however, were added to the leachate. A significant amount of COD (42,5%, 64 000 to 36 800 mg/ℓ) could be removed with lime treatment, while almost no phenols were removed in this case.

Soda ash treatment of the leachate showed that a significant amount of calcium (83,3% , 983 to 164 mg/ℓ) could be removed from the leachate at a high dosage (Table 7.7). Significant amounts of manganese (96,4%, 29,1 to 4,0 mg/ℓ) and barium (75%, 0,32 to 0,08 mg/ℓ) could also be removed from the leachate with soda ash and lime treatment, while strontium removal (50,7%, 3,83 to 1,89 mg/ℓ) was also obtained. Very little magnesium removal was obtained because the pH was not high enough, while high concentrations of sodium were introduced into the leachate as a result of soda ash treatment. Some COD and phenol removal could also be achieved.

Treatment of the leachate with caustic soda and soda ash showed that excellent removals of calcium (83,1%, 983 to 166 mg/ℓ), magnesium (71,2%, 858 to 247 mg/ℓ) and barium (68,8%, 0,32 to 0,1 mg/ℓ) could be obtained (Table 7.8). Iron and manganese removals were not as good as with lime treatment. A significant amount of sodium was again introduced into the leachate. Very little COD removal was achieved, while it appeared that a significant amount of phenol could be removed from the leachate.

Table 7.5 : Treatment of the leachate with caustic soda (Leachate sample 2/6/2000).

Caustic g/ℓ	Total Vol mℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	pH after CO_2	COD (mg/ℓ)	Ca (mg/ℓ)	Mg (mg/ℓ)	Na (mg/ℓ)	Mn (mg/ℓ)	Ba (mg/ℓ)	Sr (mg/ℓ)	Si (mg/ℓ)	Iron (mg/ℓ)	Phenol (mg/ℓ)
0	1 000			87,4	6,99		64 000	983	858	19 400	29,1	0,32	3,83	13,3	91,6	900
0,68	1 000	-	-	91,4	7,98	7		1 270	1 079		42					
4,4	1 000	-	-	86	9,01	7		870	892	22 900	35,5	0,25	3,41	9	105	740
8,9	1 000	-	-	87,5	10	7	35 400	1 012	746		13,1	0,09	2,54	4,6	58,4	
12,5	1 000	-	-	88,3	11,15	7		842	181	27400	1,05					380
14	1 000	-	-	87,3	12,02	7	46 300	864	25	29100	0,45	0,03	1,52	4,4	12,6	

Table 7.6 : Treatment of the leachate with lime (Leachate sample 2/6/2000).

Lime g/ℓ	Total Vol mℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	pH after CO_2	COD (mg/ℓ)	Ca (mg/ℓ)	Mg (mg/ℓ)	Mn (mg/ℓ)	Ba (mg/ℓ)	Sr (mg/ℓ)	Si (mg/ℓ)	Iron (mg/ℓ)	Phenol (mg/ℓ)
0	1 000			87,4	6,99	7,5	64 000	983	858	29,1	0,32	3,83	13,3	91,6	900
0,9	1 000			89,6	8,01	7,5		1 470							
3,6	1 000	50	5	87,9	8,98	7,5	43 400	3 160	1 210	38,1	0,25	3,54	7	107	820
10,6	1 000	100	10	87,6	10,01	7,5	38 100	3 650	956	2,67	0,05	2,58	5,1	57,9	
16,2	1 000	200	20	90,5	11	7,5	40 000	4 750	567						1 080
26	1 000	350	35	89	11,92	7,5	36 800	3 910	213	0,24	0,03	2,36	2,6	15,7	

Table 7.7: Treatment of leachate with soda ash and lime (Leachate sample 2/6/2000).

Soda Ash g/ℓ	Lime g/ℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	pH after CO_2	COD (mg/ℓ)	Ca (mg/ℓ)	Mg (mg/ℓ)	Na (mg/ℓ)	Mn (mg/ℓ)	Ba (mg/ℓ)	Sr (mg/ℓ)	Si (mg/ℓ)	Fe (mg/ℓ)	Phenol (mg/ℓ)
0				87,4	6,99	-	64 000	983	858	19 400	29,1	0,32	3,83	13,3	95,5	900
5,5		50	5		8,02	7,5	42 400	948	1 034		34,2					
10		100	10	96,9	8,89	7,38		753	930	24 900	7,12	0,077	2,2	5,94	79,3	494
20		350	35	95,6	9,80	7,5		252	877							
43,8		400	40	105,4	9,57	7,56	41 000	164	938	33 300	3,95	0,103	2,23	0,522	79,7	474
2	3,6	50	5		8,84	7,5		1 510	1 460	24 200	36,9					640
4	3,6	100	10		8,78	7,5	43 800	1 580	1 255		22,3					
8	3,6	400	40		8,76	7,5	37 900	511	897	25 400	3,21	0,08	1,49	7,4		700
10	3,6	300	30	92,5	9,77	7,5		376	820							
12	3,6	300	30	97,4	9,71	7,55		367	863	24 500	11,3	0,019	1,64	7,79	73,5	504
15	3,6	300	30	93,4	9,89	7,5	68 500	363	870							
20	3,6	500	50	95,4	8,99	7,5	69 600	281	740							
25	3,6	600	60	99,0	9,39	7,5	72 600	221	736	27 600	10	0,007	0,806	7,61	75,8	504
2	10	50	5	85,5	10,2	7,5	71 000	2 010	929	24 300	7,13	0,19	4,82	9,83		420
4	10	100	10	86,1	10,07	7,5		2 540	1 225							
8	10	200	20	87,5	10,02	7,5		1 720	1 024	26 600	4,01	0,08	1,89	9,31		660
12	10	800	80	85,2	10,06	7,5		910	745							

Table 7.8 : Treatment of the leachate with caustic soda and soda ash (Leachate sample 2/6/2000).

Caustic g/ℓ	pH	Soda ash g/ℓ	EC mS/cm	pH	pH CO_2	COD mg/ℓ	Na mg/ℓ	Ca mg/ℓ	Mg mg/ℓ	Si mg/ℓ	Sr mg/ℓ	Fe mg/ℓ	Mn mg/ℓ	Ba mg/ℓ	Phenol mg/ℓ	Sludge (mℓ)
0	6,99	0	87,4	6,99	-	64 000	19 400	983	858	13,3	3,83	95,5	29,1	0,32	900	-
-	9	4		9,19	7,5	60 400	25 100	1 110	1 010	12,2	6,29	126	44,5	0,48	280	
1,4	10	4		10,03	7,5	69 700	24 400	925	989	6,29	5,37	106	35,6	0,3	300	<100
12	11	4	93,2	10,99	7,5	63 800	27 900	166	247	5,27	3,07	78,1	20,4	0,1	380	300

7.4. Treatment of the Industrial Solid Waste Leachate with Ash and Chemicals for the Removal of Organics and Inorganics

Treatment of the leachate with a combination of chemicals and ash for inorganics and organics removal is shown in Tables 7.9, 7.10 and 7.11. Treatment of the leachate with Sasol ash, caustic soda and soda ash is shown in Table 7.9. Excellent calcium (73,8%, 983 to 258 mg/ℓ), magnesium (95%, 858 to 52,5 mg/ℓ), manganese (98,1%, 29,1 to 0,55 mg/ℓ) and barium (50%, 0,32 to 0,16 mg/ℓ) removals were obtained with an ash dosage of 200 g/ℓ (8 g/ℓ Na_2CO_3 and 9 g/ℓ NaOH). However, much poorer removals were obtained at an ash dosage of 100 g/ℓ (8 g/ℓ Na_2CO_3 and 2,5 g/ℓ NaOH). A significant amount of sodium was also introduced into the leachate. Poor iron removals were also obtained. It also appears that very little organics were removed, although some phenol was removed. Sludge volumes, however, were very high.

Treatment of the leachate with Iscor ash, caustic soda and soda ash is shown in Table 7.10. Excellent calcium (86,8%, 983 to 130 mg/ℓ), manganese (92,1%, 29,1 to 2,3 mg/ℓ) and barium (68,8%, 0,32 to 0,1 mg/ℓ) removals were obtained (200 g/ℓ ash, 12 g/ℓ Na_2CO_3, 6,9 g/ℓ NaOH). Strontium and iron removals, however, were not that good. A significant amount of sodium was also introduced into the leachate. Poor COD removals were also obtained. However, phenol removal (92,7%, 900 to 66 mg/ℓ) was good. Sludge volume was high.

Treatment of the leachate with Lethabo ash, caustic soda and soda ash is shown in Table 7.11. Excellent calcium (90,8%, 983 to 90,6 mg/ℓ), magnesium (97,9%, 858 to 17,8 mg/ℓ), manganese (96,9%, 29,1 to 0,9 mg/ℓ) and barium removals were obtained (200 g/ℓ ash, 8 g/ℓ Na_2CO_3, 11,8 g/ℓ NaOH). However, iron and strontium removals appeared to be poor. COD removal was good, and some phenols were also removed, although not as much as with the Iscor ash and chemicals.

The best removal of phenols was obtained with Iscor ash. It is known that high concentrations of phenols can affect membranes adversely. Therefore, Iscor ash appears to be the best candidate for treatment of the leachate prior to desalination with a membrane process. Therefore, Iscor ash at a dosage of 200 g/ℓ, 12 g/ℓ soda ash and 6,9 g/ℓ caustic soda were selected for pretreatment of the leachate prior to desalination of the leachate with ED. The chemical treatment cost will amount to R20,40/kℓ for soda ash and R22,08/kℓ for caustic soda (Na_2CO_3 at R1,70/kg and NaOH at R3,20/kg).

Table 7.9 : Treatment of the leachate with Sasol fly ash, caustic soda and soda ash (Leachate sample 2/6/2000).

SASOL ash g/ℓ	Soda ash g/ℓ	EC mS/cm	Caustic g/ℓ	pH	pH CO_2	COD mg/ℓ	Ca mg/ℓ	Mg mg/ℓ	Phenol mg/ℓ	Na mg/ℓ	Mn mg/ℓ	Ba mg/ℓ	Sr mg/ℓ	Fe mg/ℓ	Total volume (mℓ)	Sludge (mℓ)
0	0	87,4	0	6,99	-	64 000	983	858	900	19 400	29,1	0,32	3,83	95,5		-
100	8		2,5	10	7,5	52 200	423	853	440	27 100	15,3	0,18	2,68	100	1 100	350
200	8		9	10,99	7,5	65 000	258	42,5	640	29 000	0,55	0,16	2,02		1 200	800

Table 7.10 : Treatment of the leachate with Iscor fly ash, caustic soda and soda ash (Leachate sample 2/6/2000).

ISCOR ash g/ℓ	Soda ash g/ℓ	EC mS/cm	Caustic g/ℓ	pH	pH CO_2	COD mg/ℓ	Ca mg/ℓ	Mg mg/ℓ	Phenol mg/ℓ	Na mg/ℓ	Mn mg/ℓ	Ba mg/ℓ	Sr mg/ℓ	Fe mg/ℓ	Total volume (mℓ)	Sludge (mℓ)
0	0	87,4	0	6,99	-	64 000	983	858	900	19 400	29,1	0,32	3,83	95,5		-
200	4	95,6	6,6	10,01	7,5	58 900	653	536		28 500					1 200	550
200	8		2,9	10,01	7,5	54 700	393	936	66	27 700	13,2	0,2	2,67	84,1	1 200	450
200	12	98,5	6,9	10	7,5	62 600	130	560	66	29 200	2,3	0,1	1,54	31,1	1 200	650
200	16	99,3	7	10	7,5	59 800	114	575		31 900	0,43	0	0,95	30,8	1 200	625
200	8	97,4	11,2	11	7,5	66 500	104	15		25 500	0,99	0,09	1,53	13,0	1200	800

Table 7.11 : Treatment of the leachate with Lethabo fly ash, caustic soda and soda ash (Leachate sample 2/6/2000).

LETHABO ash g/ℓ	Activated carbon mg/ℓ	Soda ash g/ℓ	EC mS/cm	Caustic g/ℓ	pH	pH CO_2	COD mg/ℓ	Phenol mg/ℓ	Ca mg/ℓ	Mg mg/ℓ	Na mg/ℓ	Mn mg/ℓ	Ba mg/ℓ	Sr mg/ℓ	Fe mg/ℓ	Total volume (mℓ)	Sludge (mℓ)
0		0	87,4	0	6,99	-	64 000	900	983	858	19 400	29,1	0,32	3,83	95,5		-
200	0	8		2,6	10,01	7,5	36 600	220	413	843	25 800	16,2	0,18	2,52	95,8	1 200	450
200	0	8	95,5	11,8	11	7,5	71 000		90,6	17,8	30 600	0,9				1 200	675
200	200	8		2,6	10,01	7,5	34 800	540									450
200	400	8		2,6	10,03	7,5	35 000	480									450
200	800	8	89,4	2,6	10,00	7,5	36 900	440								1 200	450

Sludge volumes and water recovery are shown in Table 7.12. Water recovery was 800 mℓ (80%) after vacuum filtration while the sludge volume was only 300 mℓ .

Table 7.12 : Sludge volumes and water recovery.

ISCOR ash (g/ℓ)	Soda ash (g/ℓ)	Caustic soda (g/ℓ)	Leachate (mℓ)	Total volume (mℓ)	Sludge volume $(mℓ)_{1\ hour}$	Water recovery* (mℓ)	Sludge volume* (mℓ)	Wet weight* (g)	Dry weight (g)
200	12	6,9	1 000	1 200	550	800	300	312	219

* Vacuum filtration

Treatment of the leachate with powder activated carbon is shown in Table 7.13. Very little COD could be removed with the activated carbon at the dosages applied. However, some phenols could be removed.

Table 7.13 : Treatment of industrial solid waste leachate with powder activated carbon (Leachate sample 2/6/2000).

Activated carbon mg/ℓ	COD mg/ℓ	Phenol mg/ℓ
0	64 000	900
200		500
400		580
1 000	62 800 (1,9%)	720 (20%)

( ) : % removal

8. EVALUATION OF COAGULATION/FLOCCULATION PRETREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR THE REMOVAL OF SUSPENDED MATERIAL AND ORGANICS

8.1 Treatment of the Industrial Solid Waste Leachate with Flocculants (Sample of February 2000)

Treatment of the leachate received during February 2000 from Holfontein with flocculants are shown in Tables 8.1, 8.2, 8.3 and 8.4 (see 3.5).

Preliminary test work has shown that suspended solids removals could be obtained with the flocculants. However, it is interesting to note that better suspended solids removals were obtained at lower flocculant dosage than at higher dosage. This is most probably due to deflocculation at the higher flocculant dosages. Poor organics removals were obtained.

Almost no COD removal could be obtained. It further appears that sludge volumes increase with increasing flocculant dosage level.

Table 8.1 : Treatment of the leachate with aluminium sulphate (Alum).

Alum (mg/ℓ)	pH	Conduc-tivity (mS/m)	Sludge volume $(ℓ)_{30\ min}$	COD (mg/ℓ)	Suspended solids (mg/ℓ)	Removal (%)	
						COD	Sus. Solids
0	7,63	3 660	-	18 000	680	-	-
200	7,57	3 710	-		220		67,65
400	7,57	3 700	-		266		60,88
800	7,36	3 700	-	18 000	368	0,00	45,88
1 600	6,72	3 680	-	17 000	618	5,56	9,12

Table 8.2 : Treatment of the leachate with ferric chloride.

Ferric (mg/ℓ)	Add NaOH (mg/ℓ)	pH	Conductivity (mS/m)	Sludge volume $(ℓ)_{30\ min}$	COD (mg/ℓ)	Suspended solids (mg/ℓ)	Removal (%)	
							COD	Sus. Solids
0		7,63	3 660	-	18 000	680	-	-
400		6,61	3 730	15		585		13,97
800		6,3	3 740	50	18 750	498	-4,17	26,76
1 600	280	6,11	3 780	50		510		25,00
3 200	1 040	6,09	3 840	50	16 560	579	8,00	14,85

Table 8.3 : Treatment of the leachate with poly-aluminium chloride (PAC).

PAC (mg/ℓ)	pH	Conductivity (mS/m)	Sludge volume $(\ell)_{30\ min}$	COD (mg/ℓ)	Suspended solids (mg/ℓ)	Removal (%)	
						COD	Sus. Solids
0	7,63	3 660	-	18 000	680	-	-
400	7,4	3 640	50		237		65,15
800	7,35	3 640	50	12 250	147	31,94	78,38
1 600	7,27	3 640	150		488		28,24
3 200	7,10	3 640	450	18 250	500	-1,39	26,47

Table 8.4 : Treatment of the leachate with a polyelectrolyte (Magnafloc 1797).

Magnafloc 1797 (mg/ℓ)	pH	Conductivity (mS/m)	Sludge volume $(\ell)_{30\ min}$	COD (mg/ℓ)	Suspended solids (mg/l)	Removal (%)	
						COD	Sus. Solids
0	7,63	3 660	-	18 000	680	-	-
1 000	7,82	3 680	-		688		-1,18
2 000	7,79	3 680	175	18 000	390	0,00	42,65
4 000	7,76	3 660	200		209		69,26
6 000	7,72	3 610	200	20 000	251	-11,11	63,09

8.2 Treatment of the Industrial Solid Waste Leachate with Flocculants (Sample of June 2000)

Treatment of the leachate received during February 2000 with flocculants is shown in section 8.1. Treatment of the leachate received during June 2000 with flocculants is shown in Tables 8.5 to 8.8.

Treatment of the leachate with alum is shown in Table 8.5. Approximately 70% (3 778 to 1 147 mg/ℓ) of the suspended solids could be removed with an alum dosage of 200 mg/ℓ. Poorer suspended solids removals were obtained at the higher dosages. The best suspended solids and COD removals, however, were obtained by filtration of the leachate through Whatman No. 1 filter paper. Suspended solids removal of approximately 73,8% (3 778 to 988 mg/ℓ) were obtained. Poor COD removals were obtained.

Treatment of the leachate with ferric chloride is shown in Table 8.6. Suspended solids removal of approximately 68% (3 778 to 1 212 mg/ℓ) was obtained at a ferric chloride dosage of 200 mg/ℓ. Approximately the same removal was obtained at a dosage of 400 mg/ℓ, while it appeared that somewhat better removals were obtained at higher dosages. It is interesting to note that the best removal of suspended solids (86,3%, 3 778 to 515 mg/ℓ) was obtained when the pH was lowered to 4,5 with acid addition. Phenol removal was 64,4% (900 to 320 mg/ℓ) in this case. COD removals also appear to be better in this case than in the case of alum treatment.

Treatment of the leachate with poly-aluminium chloride is shown in Table 8.7. Suspended solids removal of approximately 82% (3 778 to 698 mg/ℓ) could be obtained with a PAC dosage of 200 mg/ℓ. Suspended solids removal appears to decrease at a very high dosage (1 600 mg/ℓ). Significant amounts of COD were removed at the higher dosages (42,2%, 64 000 to 37 000 mg/ℓ).

Treatment of the leachate with Magnafloc 1797 is shown in Table 8.8. Approximately 63%, (3 778 to 1 405 mg/ℓ) of the suspended solids were removed at a dosage of 200 mg/ℓ flocculant. Suspended solids removal, however, appears to decrease at a dosage of 400 mg/ℓ, while it improved at the higher dosages. Almost no COD removal was obtained.

Suspended solids removal from the leachate with flocculants was not very successful. Therefore, flocculation should be coupled with filtration for better removals of suspended solids prior to ED desalination.

Table 8.5 : Treatment of the leachate with alum (Leachate sample 2/6/2000).

Alum mg/ℓ	pH	Add NaOH mg/ℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	COD mg/ℓ	SS mg/ℓ
0	-	-	-	-	87,4	6,99	64 000	3 778
Filtered Whatman No 1							56 700	988
100	7,07	-	-	-	88,6	7,07		
200	7,07	-	-	-	88,6	7,07	67 000	1 147
400	6,91	-	-	-	88,2	6,91	66 200	1 463
800	6,65	120	-	-	88,2	6,95	67 000	1 031
1 600	6,29	320	-		88	6,99	62 000	1 319

Table 8.6 : Treatment of the leachate with ferric chloride (Leachate sample 2/6/2000).

Ferric mg/ℓ	Sludge Vol mℓ	% Sludge	Sulphuric acid g/ℓ	EC mS/cm	pH	COD mg/ℓ	Phenol mg/ℓ	SS mg/ℓ
0	-	-	-	87,4	6,99	64 000	900	3 778
Filtered Whatman No 1						56 700		988
100	-	-	-	90	6,78			
200	-	-	-	89,5	6,7	67 800		1 212
200			13,5		4,52	57 900	320	515
400	-	-	-	89,3	6,41	36 100		1 213
800	-	-	-	89,2	6	36 600		986
1 600	-	-	-	88,9	5,65	36 000		758
3 200	-	-	11	88,4	4,37	54 400		

Table 8.7 : Treatment of the leachate with poly-aluminium chloride - PAC (Leachate sample 2/6/2000).

PAC mg/ℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	COD mg/ℓ	SS mg/ℓ
0	-	-	87,4	6,99	64 000	3 778
Filtered Whatman No 1					56 700	988
100	-	-	90	6,78	63 800	1274
200	-	-	89,5	6,7	57 800	698
400	-	-	89,3	6,8	44 300	1064
800	-	-	89,2	6,79	37 000	736
1600	-	-	88,9	6,8	37 100	1465

Table 8.8 : Treatment of the leachate with Magnafloc 1797 (Leachate sample 2/6/2000).

Magnafloc mg/ℓ	Sludge Vol mℓ	% Sludge	EC mS/cm	pH	COD mg/ℓ	SS mg/ℓ
0	-	-	87,4	6,99	64 000	3 778
Filtered Whatman No 1					56 700	988
100	-	-				
200	-	-			65 800	1405
400	-	-			68 100	1962
800	-	-			66 100	799
1 600	-	-	87,6	7,01	68 000	763

9. EVALUATION OF NANOFILTRATION FOR THE TREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE

9.1 Introduction

Nanofiltration (AFC40, MPT31 and MPT36) and ultrafiltration membranes (polysulphone) were evaluated for the treatment of the leachate. The objectives of the investigation were to determine to what extent the organics could be separated from the inorganics, and to what extent the monovalent ions could be separated from the divalent ions in the leachate. Another objective was to determine the fouling potential of the leachate for the membranes.

9.2 Treatment of the Leachate with Tubular AFC40 Membranes

Industrial solid waste leachate (12 ℓ) was circulated through a tubular AFC40 membrane (0,0261 m² area) in the batch made at a feed inlet pressure of 4 300 kPa (see 3.6). Permeate was withdrawn and the feed was allowed to concentrate. The run was terminated at a water recovery of approximately 60%. The experimental conditions and results are shown in Table 9.1.

Permeate flux as a function of time and percentage water recovery is shown in Figures 9.1 and 9.2, respectively. Permeate flux was approximately 1 000 ℓ/m².d in the beginning of the run, and decreased as a function of time and percentage water recovery. Permeate flux was approximately 500 ℓ/m².d when the run was terminated.

The clean water flux was determined at approximately 2 000 ℓ/m².d at the end of the run (Figure 9.1). Therefore, membrane fouling had taken place (initial CWF 4 000 ℓ/m².d). However, it was possible to restore the CWF with a dilute caustic clean.

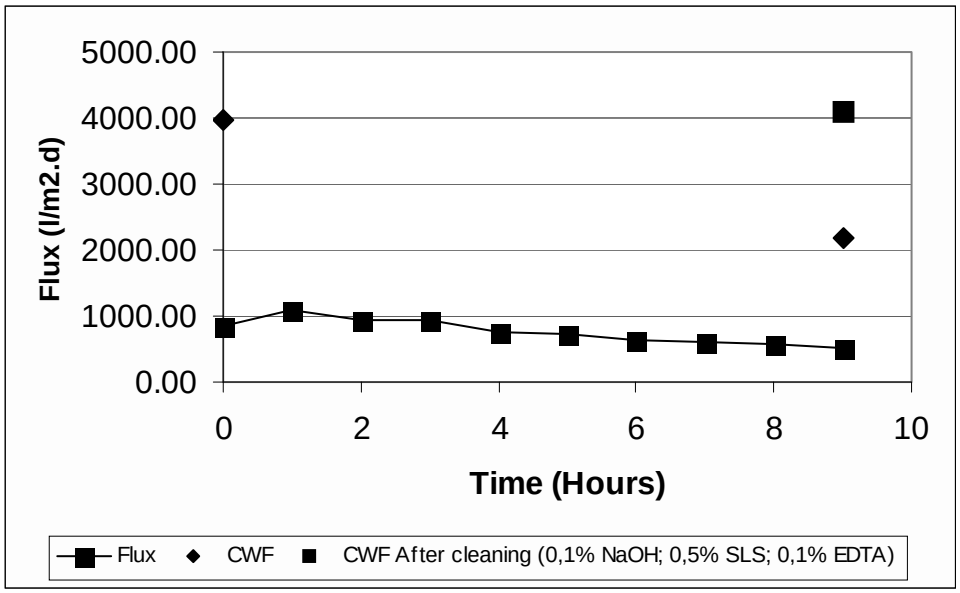


Figure 9.1 : Permeate flux as a function of time.

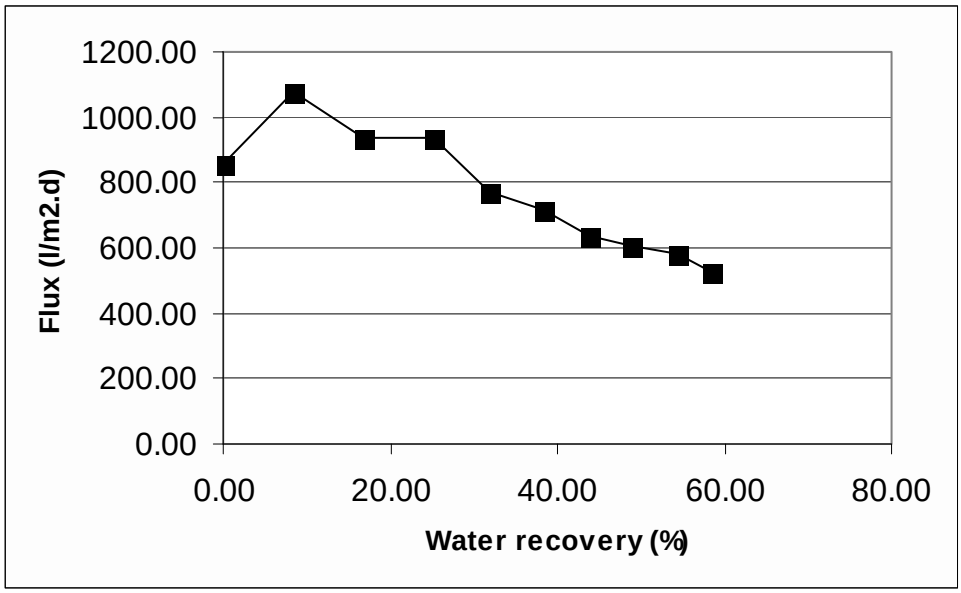


Figure 9.2 : Permeate flux as a function of percentage water recovery.

The electrical conductivity of the feed and permeate is shown in Figure 9.3. The electrical conductivity removal is shown in Figure 9.4.

Very little salinity was retained in the feed by the membrane. The initial conductivity rejection was about 17% and decreased to about 8% when the run was terminated.

The chemical composition of the feed, product and brine is shown in Table 9.2. Most of the organics, as indicated by the COD (24,2%) and TOC (19,3%) in the permeate (product), remained in the brine. However, a significant amount of the phenolics (55,7%) permeate the membrane.

Most of the TDS also remained in the brine (28,4% TDS in permeate). However, significantly more of the monovalent chloride ions (52,5%) were present in the permeate than the divalent sulphate ions (18%). Also, significantly more monovalent sodium ions were present in the permeate (40,4%) than divalent calcium (14,6%) and magnesium (15,0%) ions.

Table 9.1 : Experimental conditions and results for the treatment of the leachate with AFC40 NF membranes.

Time	Pressure in	Pressure out	Temper-ature	Flow rate	Flux	Feed	Product	Volume recovered		Recovery	Rejection
	(kPa)	(kPa)	(°C)	(mℓ/min)	(ℓ/m2.d)	(mS/cm)	(mS/cm)	(ℓ)	(ℓ)	(%)	(%)
0	4 300	4 000	23	72	3 972,41	27,7	5,69	12	CWF	-	79,46
0	4 300	4 000	25	15,5	855,17	88,7	-	0	0	0,00	
1	4 300	4 000	39	19,5	1 075,86	86,6	71,6	1	1	8,33	17,32
2		4 000	42	17	937,93	86,7	76,6	1	2	16,67	11,65
3	4 300	4 000	41,5	17	937,93	87,1	77,6	1,02	3,02	25,17	10,91
4	4 300	4 000	40	14	772,41	88,1	78,5	0,8	3,82	31,83	10,90
5	4 300	4 000	40	13	717,24	88,4	79,8	0,78	4,6	38,33	9,73
6	4 300	4 000	39,5	11,5	634,48	88,1	80	0,67	5,27	43,92	9,19
7	4 300	4 000	39	11	606,90	88,6	81	0,6	5,87	48,92	8,58
8	4 300	4 000	40,5	10,5	579,31	88,8	81,2	0,65	6,52	54,33	8,56
9	4 300	4 000	41	9,5	524,14	89,6	82,4	0,5	7,02	58,50	8,04
9	4 300	4 000	23	39,5	2 179,31						
9	4 300	4 000	23	74	4 082,76	CWF after cleaning (0.1 % NaOH, pH 9,23 (sulphuric acid) + 0,1% EDTA + 0,5% SLS)					

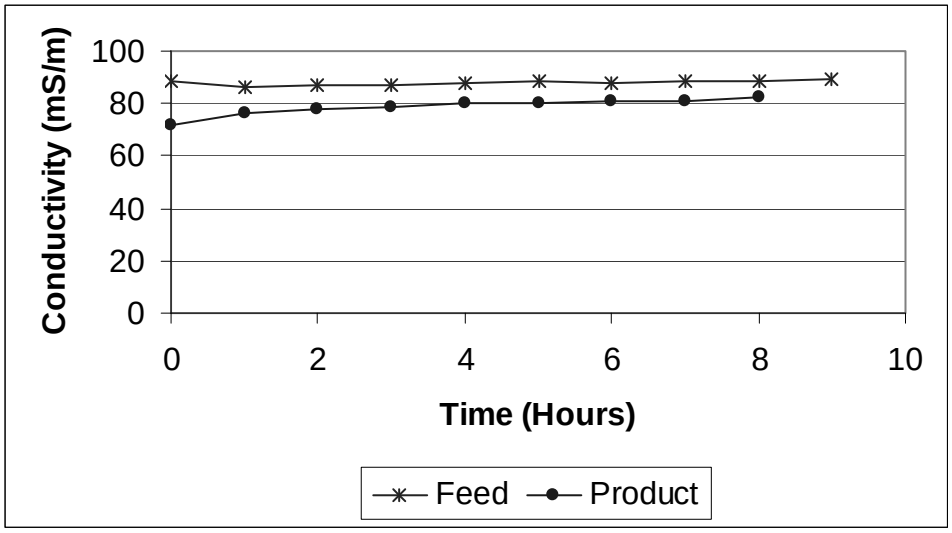


Figure 9.3 : Electrical conductivity of the feed and permeate as a function of time.

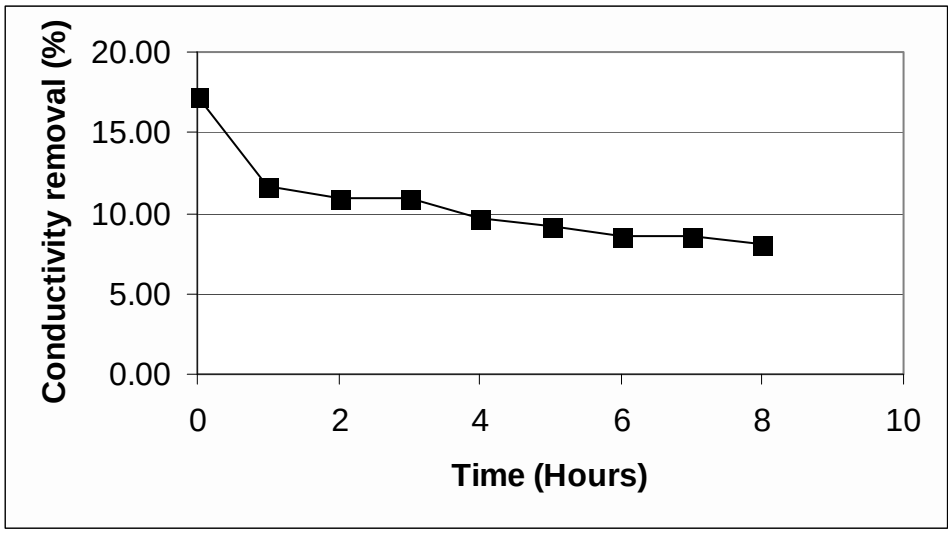


Figure 9. 4 : Electrical conductivity removal as a function of time.

Table 9.2 : Chemical composition of the nanofiltration (AFC40) feed, brine and product (feed, 19/10/2000, untreated).

Constituents*	Feed	Product	Brine	Removal (%)	Feed (g)	Product (g)	Brine (g)	Product/ Feed (%)
Conductivity (mS/cm)	89,8	85,2	86,9	5,12				
pH	7,6	7,35	7,74					
Chemical oxygen demand (COD)	69 000	28 500	103 000	58,70	828,00	200,07	512,94	24,16
Total organic Carbon	18 484	6 086	32 129	67,07	221,81	42,72	160,00	19,26
Alkalinity as $CaCO_3$	10 244	4 593	14 071	55,16	122,93	32,24	70,07	26,23
Total dissolved solids	117 005	56 865	139 560	51,40	1 404,06	399,19	695,01	28,43
Ammonia Nitrogen as N	1 894	1 323	2 460	30,15	22,73	9,29	12,25	40,86
Chloride	29 904	26 810	23 740	10,35	358,85	188,21	118,23	52,45
Sulphate	20 028	6 160	33 070	69,24	240,34	43,24	164,69	17,99
Calcium	1 040	260	1 480	75,00	12,48	1,83	7,37	14,63
Magnesium	993	254	1 310	74,72	11,92	1,78	6,52	14,96
Sodium	23 000	15 900	28 400	30,87	276,00	111,62	141,43	40,44
Potassium	7 090	4 060	7 850	42,74	85,08	28,50	39,09	33,50
Phenolic compounds	679	646	542	4,86	8,15	4,53	2,70	55,66

* Concentration in mg/ℓ, unless stated otherwise

Feed	12 ℓ
Product	7,02 ℓ
Brine	4,98 ℓ

9.3 Treatment of the Industrial Solid Waste Leachate with Tubular MPT31 Membranes

Industrial solid waste leachate was treated as in the previous case (9.2), but at a feed inlet pressure of 3 700 kPa in this case.

The experimental conditions and results are shown in Table 9.3.

Table 9.3 : Experimental conditions and results for the treatment of the leachate with MPT31 NF membranes.

Time	Pressure in	Pressure out	Temperature	Flow rate	Flux	Feed	Product	Volume recovered		Recovery	Rejec-tion
	(kPa)	(kPa)	(°C)	(mℓ/min)	(ℓ/m2.d)	(mS/cm)	(mS/cm)	(ℓ)	(ℓ)	(%)	(%)
0	3 700	3 500	23	74	4 082,76	28,1	12,2	12	CWF	-	
0	3 700	3 500	23,5	12,5	689,66	88,9		0	0	0,00	
1	3 700	3 500	38	15	827,59	88,2	75,5	0,76	0,76	6,33	14,40
2	3 700	3 500	38	13	717,24	87,2	77,9	0,74	1,5	12,50	10,67
3	3 700	3 500	36,5	12,5	689,66	88,2	78,9	0,68	2,18	18,17	10,54
4	3 700	3 500	37,5	11,5	634,48	88,4	79,9	0,69	2,87	23,92	9,62
5	3 700	3 500	36,5	10,5	579,31	88,3	80,4	0,54	3,41	28,42	8,95
6	3 700	3 500	37	10	551,72	88,6	80,9	0,59	4	33,33	8,69
7	3 700	3 500	37,5	9,5	524,14	88,9	81,9	0,5	4,5	37,50	7,87
8	3 700	3 500	41	12,5	689,66	89	83,7	0,74	5,24	43,67	5,96
9	3 700	3 500	40	11,5	634,48	89,1	83,5	0,59	5,83	48,58	6,29
10	3 700	3 500	40,5	10,5	579,31	89,5	84,4	0,52	6,35	52,92	5,70
11	3 700	3 500	42,5	10	551,72	88,9	84,3	0,5	6,85	57,08	5,17
12	3 700	3 500	41	9	496,55	89,4	85,7	0,47	7,32	61,00	4,14
12	3 700	3 500	23	46,5	2 565,52	CWF					
12	3 700	3 500	22,5	72	3 972,41	CWF (After cleaning with 0.5% P3 Ultrasil 91)					
						86,4	82				

The permeate flux as a function of time and percentage water recovery is shown in Figures 9.5 and 9.6, respectively. Permeate flux was high in the beginning (700 to 800 ℓ/m².d) and decreased towards the end of the run (500 ℓ/m².d). The CWF could be restored after cleaning with Ultrasil 91 cleaning agent.

The electrical conductivity of the feed and permeate as a function of time is shown in Figure 9.7. The electrical conductivity removal was 14% in the beginning of the run, and decreased to 4% at the end of the run (Figure 9.8).

The chemical composition of the feed, product and brine is shown in Table 9.4.

Most of the organics again remained in the brine (35,4% COD in product, and 15,9% TOC in product). Significantly more monovalent chloride ions (56,2%) were present in the permeate than divalent sulphate ions (10,3%). The same applied to the monovalent sodium (44%) and calcium ions (31,3%).

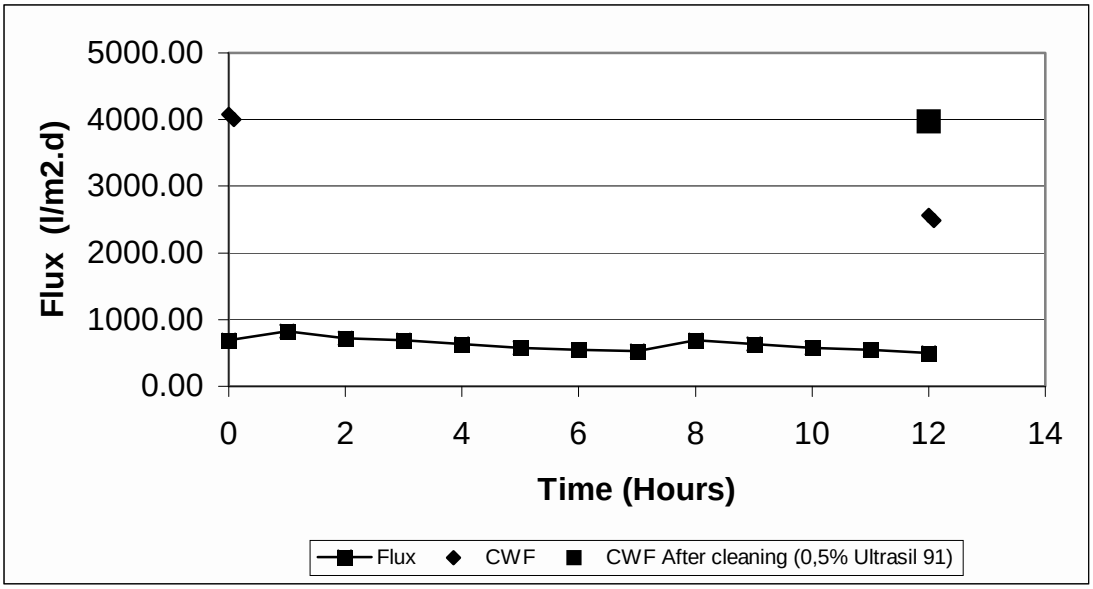


Figure 9.5 : Permeate flux as a function of time.

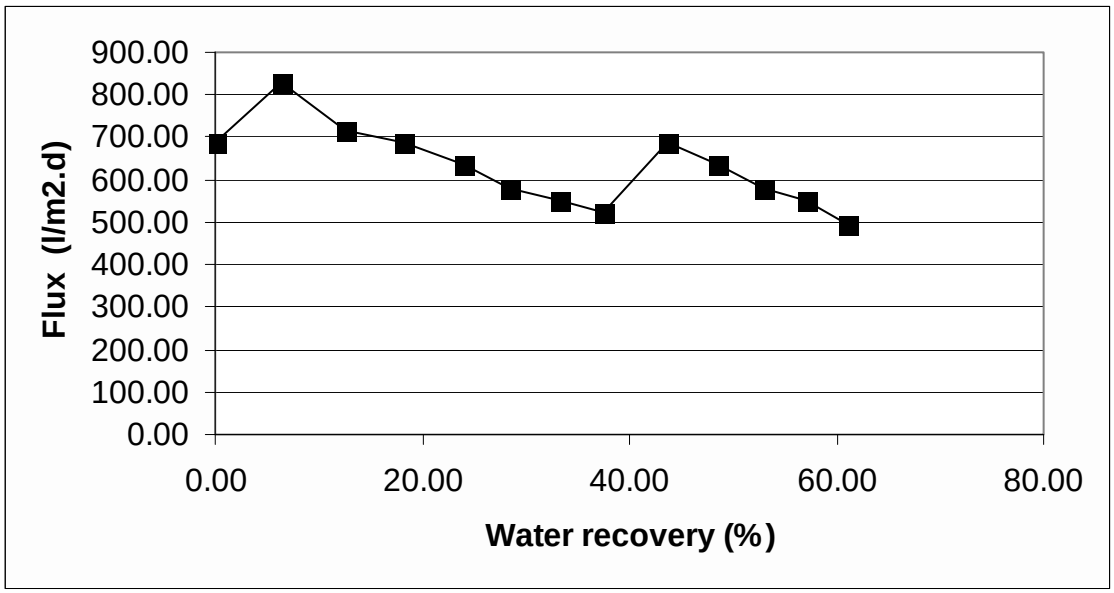


Figure 9.6 : Permeate flux as a function of water recovery.

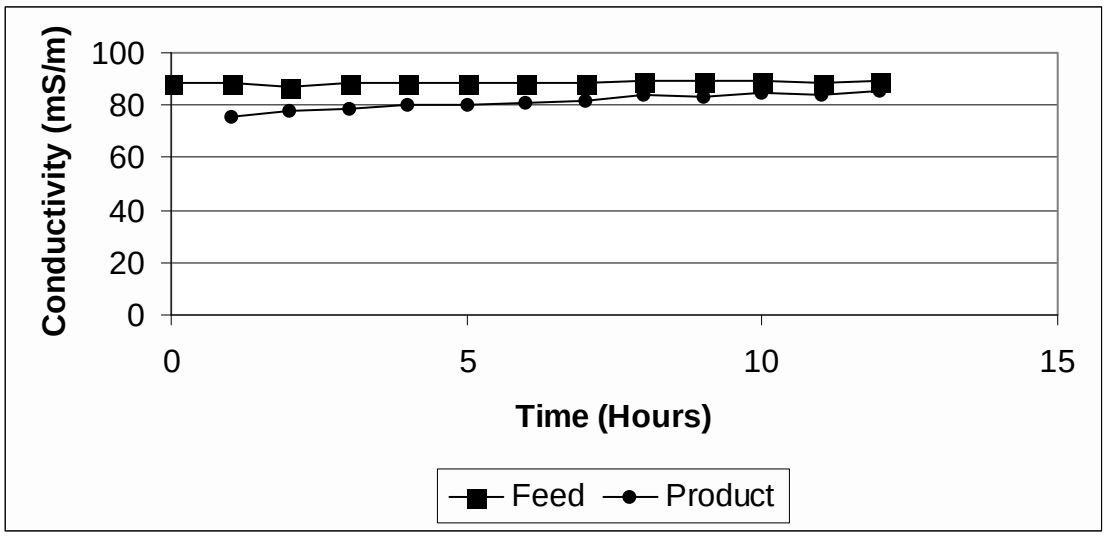


Figure 9.7 : Electrical conductivity of the feed and product as a function of time.

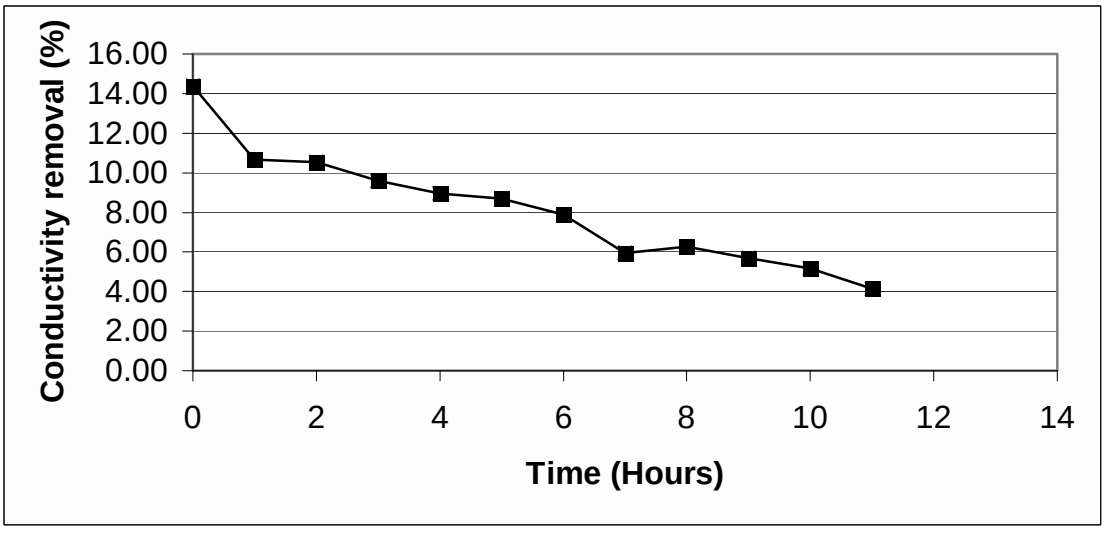


Figure 9.8 : Electrical conductivity removal as a function of time.

Table 9.4 : Chemical composition of the nanofiltration (MPT31) feed, brine and product (feed, 19/10/2000, untreated).

Constituents*	Feed	Product	Brine	Removal (%)	Feed (g)	Product (g)	Brine (g)	Product/ Feed (%)
Conductivity (mS/cm)	86,4	82	89,4	5,09				
pH								
Chemical oxygen demand (COD)	69 000	40 000	92 000	42,03	828,00	292,80	430,56	35,36
Total organic Carbon	18 484	4 804	21 437	74,01	221,81	35,17	100,33	15,85
Alkalinity as $CaCO_3$	10 244	4 937	14 683	51,81	122,93	36,14	68,72	29,40
Total dissolved solids	117 005			100,00	1 404,06	0,00	0,00	0,00
Ammonia Nitrogen as N	1 894	1 412	2 390	25,45	22,73	10,34	11,19	45,48
Chloride	29 904	27 559	30 617	7,84	358,85	201,73	143,29	56,22
Sulphate	20 028	3 376	39 080	83,14	240,34	24,71	182,89	10,28
Calcium	1 040	534	3 260	48,65	12,48	3,91	15,26	31,32
Magnesium	993	152	1 120	84,69	11,92	1,11	5,24	9,34
Sodium	23 000	16 600	28 200	27,83	276,00	121,51	131,98	44,03
Potassium	7 090	4 430	7 130	37,52	85,08	32,43	33,37	38,11
Phenolic compounds	679	592	661	12,81	8,15	4,33	3,09	53,18

* Concentration in mg/ℓ , unless stated otherwise

Feed	12
Product	7,32
Brine	4,68

9.4 Treatment of the Industrial Solid Waste Leachate with Tubular MPT36 Membranes

Leachate was treated as under 9.1, but at a feed inlet pressure of 3 700 kPa. Three runs were conducted.

The experimental conditions and results are shown in Table 9.5.

Permeate flux as a function of time and percentage water recovery for three runs are shown in Figures 9.9 and 9.10 (3rd run only), respectively. Permeate flux started at approximately 400 ℓ/m².d and was about the same when the run was terminated.

The CWF was approximately 3 500 ℓ/m².d before the first run was started (Figure 9.9). The CWF at the end of the first run was about 1 900 ℓ/m².d (Table 9.5). The CWF decreased to about 1 200 ℓ/m².d after the second run, and was about 900 ℓ/m².d after the third run. Two cleanings with Ultrasil cleaning solution increased the CWF to approximately 2 600 and 3 300 ℓ/m².d, respectively. Therefore, CWF was almost restored. However, the leachate will foul the membrane and regular chemical cleaning will be necessary to restore performance.

The electrical conductivity of the feed and permeate is shown in Figure 9.11. The conductivity removal as a function of time is shown in Figure 9.12.

Table 9.5 : Experimental conditions and results for the treatment of the leachate with MPT36 NF membranes (run 1).

Time	Pressure in	Pressure out	Temperature	Flow rate	Flux	Feed	Product	Volume recovered		Recovery	Rejection
	(kPa)	(kPa)	(deg C)	(mℓ/min)	(ℓ/m2.d)	(mS/cm)	(mS/cm)	(ℓ)	(ℓ)	(%)	(%)mS/m
0	3 700	3 500	23	64	3 531,03	CWF	-	12		-	
0	3 700	3 500	20	10	551,72	89,8	-	0	0	0.00	
0.5	3 700	3 500	32,5	12	662,07	87,6	80,7	0,26	0,26	2,17	7,88
1	3 700	3 500	38,5	11,5	634,48	87,1	81,8	0,32	0,58	4,83	6,08
2	3 700	3 500	40	10,5	579,31	87,8	83,5	0,7	1,28	10,67	4,90
3	3 700	3 500	41,5	11,5	634,48	86,5	83,3	0,7	1,98	16,50	3,70
4	3 700	3 500	40	10	551,72	86,6	83,6	0,6	2,58	21,50	3,46
5	3 700	3 500	40,5	10	551,72	86,8	83,4	0,5	3,08	25,67	3,92
6	3 700	3 500	37	11,5	634,48	88,2	85,7	0,6	3,68	30,67	2,83
7	3 700	3 500	42	11,5	634,48	87,1	84,4	0,82	4,5	37,50	3,10
8	3 700	3 500	40	11,5	634,48	87,3	85,2	0,62	5,12	42,67	2,41
9	3 700	3 500	41	11	606,90	87,7	85,7	0,7	5,82	48,50	2,28
10	3 700	3 500	40,5	11	606,90	87,7	85,9	0,52	6,34	52,83	2,05
11	3 700	3 500	41,5	10	551,72	87,7	85,8	0,55	6,89	57,42	2,17
12	3 700	3 500	43	11	606,90	87,6	86,9	0,675	7,57	63,04	0,80
12	3 700	3 500	23	34	1 875,86	CWF					
12	3 700	3 500	23	66	3 641,38	CWF (After cleaning with 0,5% P3 Ultrasil 91)					

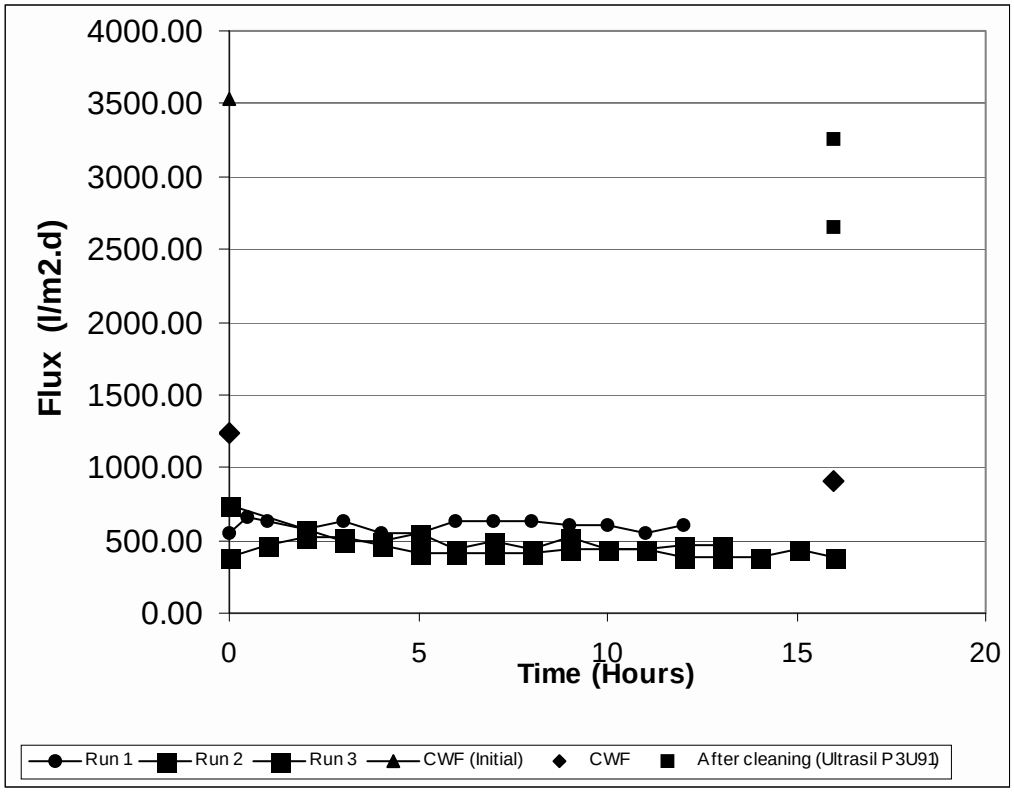


Figure 9.9 : Permeate flux as a function of time.

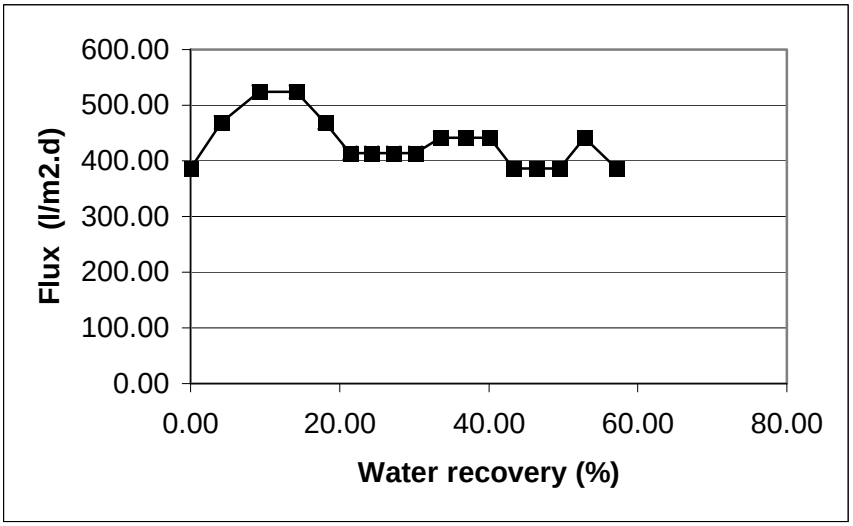


Figure 9.10 : Permeate flux as a function of percentage water recovery.

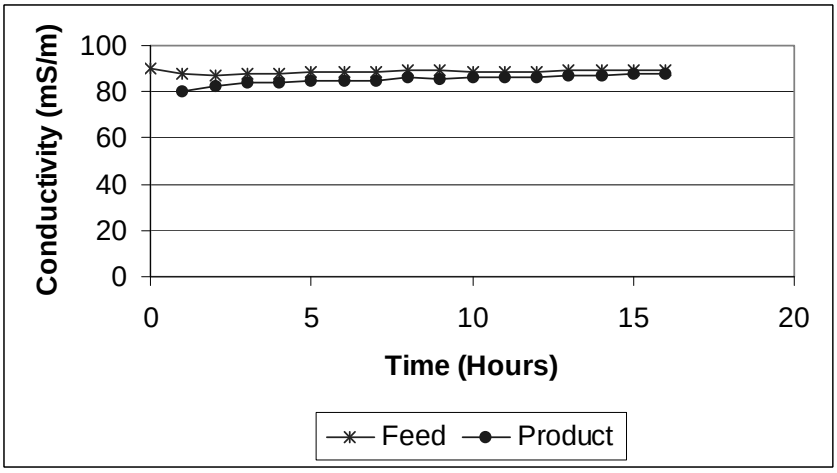


Figure 9.11 : Electrical conductivity of the feed and permeate.

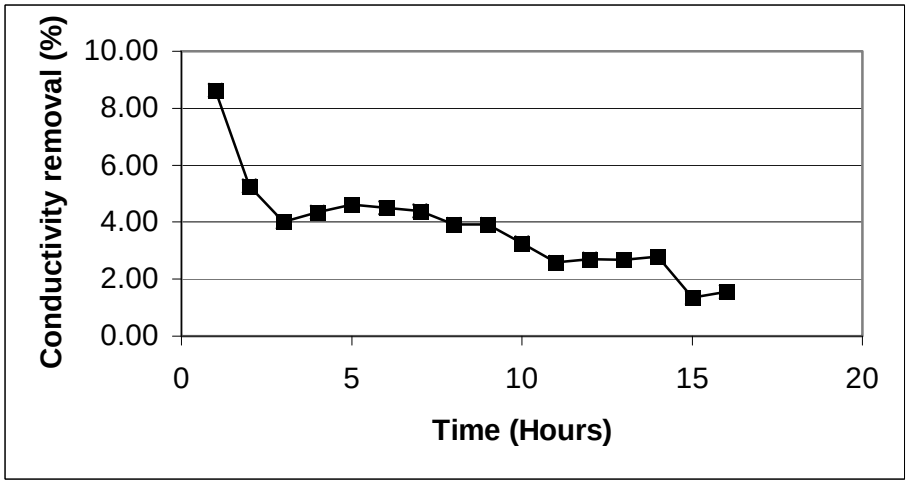


Figure 9.12 : Electrical conductivity removal as a function of time.

The electrical conductivity removal was approximately 8% in the beginning of the run, and decreased to approximately 2% at the end of the run.

The chemical composition of the feed, permeate and brine is shown in Table 9.6. Most of the organics as measured by the COD (36,6% in permeate) and TOC (18% in permeate) remained in the brine. However, more phenolics are present in the feed. It is also interesting to note that a significant quantity of organics permeate all the membranes. This shows that there are a significant quantity of low molecular mass organics present in the leachate. Significantly more chloride ions (58,1%) are present in the permeate than sulphate ions (25,1%). The same applies to the monovalent sodium (48%), potassium (42,2%) and the divalent calcium (30,9%) and magnesium (33%) ions.

Table 9.6 : Chemical composition of the nanofiltration (MPT36) feed, brine and product (feed, 19/10/2000, untreated).

Constituents*	Feed	Product	Brine	Removal (%)	Feed (g)	Product (g)	Brine (g)	Product/ Feed (%)
Conductivity (mS/cm)	89,8	83,4	89,5	7,13				
PH								
Chemical oxygen demand (COD)	69 000	40 000	92 000	42,03	828,00	302,80	407,56	36,57
Total organic Carbon	18 484	5 265	27 168	71,52	221,81	39,86	120,35	17,97
Alkalinity as $CaCO_3$	10 244	6 196	12 454	39,52	122,93	46,90	55,17	38,16
Total dissolved solids	117 005	72 780	128 575	37,80	1 404,06	550,94	569,59	39,24
Ammonia Nitrogen as N	1 894	1 521	2 021	19,69	22,73	11,51	8,95	50,66
Chloride	29 904	27 562	29 180	7,83	358,85	208,64	129,27	58,14
Sulphate	20 028	7 963	28 606	60,24	240,34	60,28	126,72	25,08
Calcium	1 040	509	1 240	51,06	12,48	3,85	5,49	30,87
Magnesium	993	520	1 180	47,63	11,92	3,94	5,23	33,03
Sodium	23 000	17 500	23 700	23,91	276,00	132,48	104,99	48,00
Potassium	7 090	4740		33,15	85,08	35,88		42,17
Phenolic compounds	679	615	571	9,43	8,15	4,66	2,53	57,14

* Concentration in mg/ℓ, unless stated otherwise.

Feed 12 ℓ
Product 7,57 ℓ
Brine 4,43 ℓ

The presence of a larger quantity of chloride ions than sulphate ions in the permeate will reduce the scale-forming tendency of the permeate for further desalination with a membrane process.

Only three nanofiltration membranes were evaluated for the removal of organics from the leachate and for the separation of monovalent ions from divalent ions. This type of membrane holds promise for the removal of organics from a leachate and separation of monovalent ions from divalent ions to reduce the scaling potential of an effluent prior to

further desalination. Therefore, more nanofiltration membranes with possibly better separation capabilities should be identified and studied for leachate treatment.

9.5 Treatment of the Industrial Solid Waste Leachate with Tubular Poly-Sulphone Membranes

Leachate was batch treated (see 3.6, Figure 3.2) with tubular ultrafiltration membranes (0,261 m²) and the COD and conductivity of the UF feed, permeate and brine were measured. The results are shown in Tables 9.7, 9.8 and 9.9.

Table 9.7 : UF treatment of the leachate (40 kD cut-off membrane).

Time	Pressure	Temperature	Flow rate	Flux	Water Recovery	
(min)	(kPa)	(°C)	(mℓ/min)	(ℓ/h,m^2)	(ℓ)	(%)
0	200	15,9	115	634 (CWF)	15	
0	200	17,8	30	166	0	0,00
30	200	29,4	25,5	141	1,06	7,07
60	200	32,1	23	127	1,86	12,40
90	200	33,5	22	121	2,56	17,07
120	200	34,5	22,5	124	3,24	21,60
150	200	32,8	22	121	3,84	25,60
180	200	31,6	22	121	4,44	29,60
240	200	31,4	18	99	5,51	36,73
300	200	31,4	18	99	6,46	43,07
360	200	32	17	94	7,36	49,07
420	200	31,2	18	99	8,1	54,00
480	200	37,4	19,5	108	9,18	61,20
480	200	15,9	48	265 (CWF)		

Table 9.8 : UF treatment of the leachate (10 kD cut-off membrane).

Time	Pressure in	Temper-ature	Flow rate	Flux	Feed	Pro-duct	Volume recovered		Water Recovery	Conduc. Rejec-tion
(min)	(kPa)	(°C)	(mℓ/min)	(ℓ/m^2/d)	(mS/m)	(mS/m)	(ℓ)	(ℓ)	(%)	(%)
0	200	23	250	1379,31 (CWF)			20	CWF	-	
0	200	24,4	86	474,48	84,3	-	0	0	0,00	
30	200	32,2	64	353,10	83,3	82,4	1,5	1,5	7,50	1,08
60	200	33,4	56	308,97	83	82	1,75	3,25	16,25	1,08
90	200	34,5	53	292,41	83	82	1,5	4,75	23,75	1,08
120	200	35,6	52	286,90	83	82	1,75	6,5	32,50	1,08
150	200	37,2	50	275,86	83	82	1,4	7,9	39,50	1,08
180	200	37,9	50	275,86	83	82	1,25	9,15	45,75	1,08
210	200	39,5	50	275,86	83	82	1,25	10,4	52,00	1,08
240	200	42,2	50	275,86	83	82	1,75	12,15	60,75	1,08
270	200	43	50	275,86	83	82	1,25	13,4	67,00	1,08
270	200	23,9	88	485,52				CWF		
270	200	23	220	1213,79 (CWF)				CWF	After cleaning with 0,5 % Ultrasil P3U91	

Membrane fouling was experienced during treatment of the leachate. The CWF has dropped from 634 ℓ/m^2.h (initially) to 265 ℓ/m^2.h at the end of the run (Table 9.7). The membranes, however, were cleaned with Ultrasil P3U91 after the second run (Table 9.8) and the CWF was almost restored (1 379,31 ℓ/m^2.d initially and 1 213,79 ℓ/m^2.d after cleaning).

Table 9.9 : COD of the UF feed, product and brine.

Feed mg/ℓ	Product mg/ℓ	Brine mg/ℓ	% Removal
64 000(1)	57 600		10
69 000(2)	55 000	74 500	20,3

(1) 40 kD cut off

(2) 10 kD cut off

Only 10% of the COD was rejected by the 40 kD membrane, while approximately 20,3% of the COD was rejected by the 10 kD membrane. Therefore, most of the organics in the leachate permeate the membranes

10. DETERMINATION OF THE FOULING POTENTIAL OF THE INDUSTRIAL SOLID WASTE LEACHATES FOR ELECTRODIALYSIS MEMBRANES

10.1 Introduction

Industrial solid waste leachate (5 litres) from Holfontein was passed through a specially designed ED membrane fouling test cell (membrane area 0,5077 cm^2) under a current density of 20 mA/cm^2 and the voltage drop across the membranes was measured as a function of time (see 3.7, Figure 3.3). An increase in voltage drop across the membranes indicates membrane fouling. Membrane resistance before and after ED was also determined. The fouling potential of different ion-exchange membranes were evaluated with the aim of selecting the most fouling resistant membrane for treatment of the leachate.

10.2 Fouling Potential of the Leachate for Selemion AMV and CMV Membranes

The voltage drop across the anionic membrane (AMV) as a function of time for two runs is shown in Figures 10.1 and 10.2. An increase in voltage drop across the membrane was experienced in the beginning of the run, whereafter it remained more or less constant. This increase in voltage drop across the membrane can be ascribed to membrane fouling.

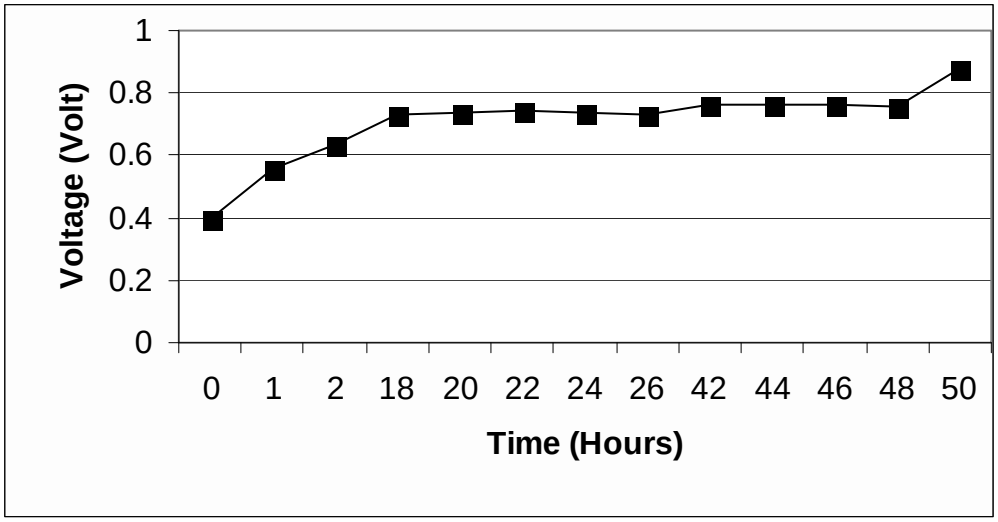


Figure 10.1 : Voltage drop across the anionic membrane (AMV) as a function of time (run 1).

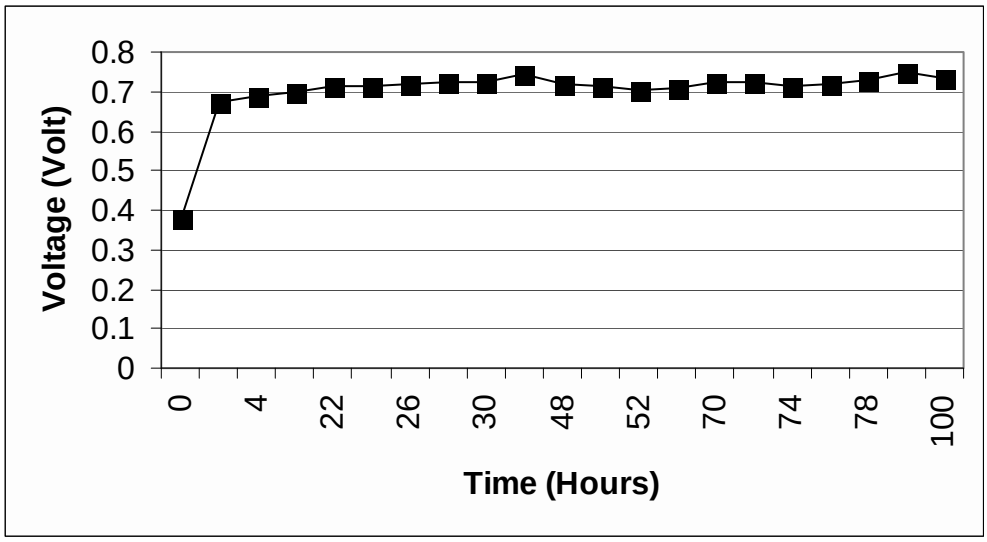


Figure 10.2 : Voltage drop across the anionic membrane (AMV) as a function of time (run 2).

The electrical resistances of the new, used and cleaned membranes were also measured. The results are shown in Table 10.1.

Table 10.1 : Electrical resistance of the new, used and cleaned membranes.

Membrane resistance ohm.cm² (1)								
Run No.	New membranes		Used membranes from cell		Used membrane after equilisation		Used membrane after cleaning (2)	
	AMV	CMV	AMV	CMV	AMV	CMV	AMV	CMV
1	3,10	6,55	9,54	6,90	7,92	6,45	2,34	6,50
2	2,34	6,50	8,69	4,57	6,40	3,50	4,67	5,38

(1) Membranes were equilibrated overnight in 0,5 N NaCl prior to resistance measurements. The used membranes from the cell were not equilibrated prior to resistance measurements.

(2) Soaked overnight in 10% NaCl at pH 10,5.

The used anionic membrane just after the fouling experiment showed a large increase in resistance as a result of membrane fouling. It was found that a black layer of material had deposited on the anionic membrane. However, mechanical removal of this layer and equilisation of the membrane in 0,5 NaCl reduced the membrane resistance significantly. Cleaning of the membrane in 10% NaCl at pH 10,5 restored the membrane resistance of the anionic membrane (run 1). However, the resistance of the membrane was not completely restored after the second run. The cationic membrane, on the other hand, showed very little signs of membrane fouling.

The effectiveness of the fouling cell for determining the fouling potential of effluents for membranes was evaluated by spiking the leachate with DBS during ED of the leachate in the

fouling cell. This was also done during the desalination of a 3 000 mg/ℓ NaCl solution in the fouling cell. The results are shown in Figures 10.3 and 10.4.

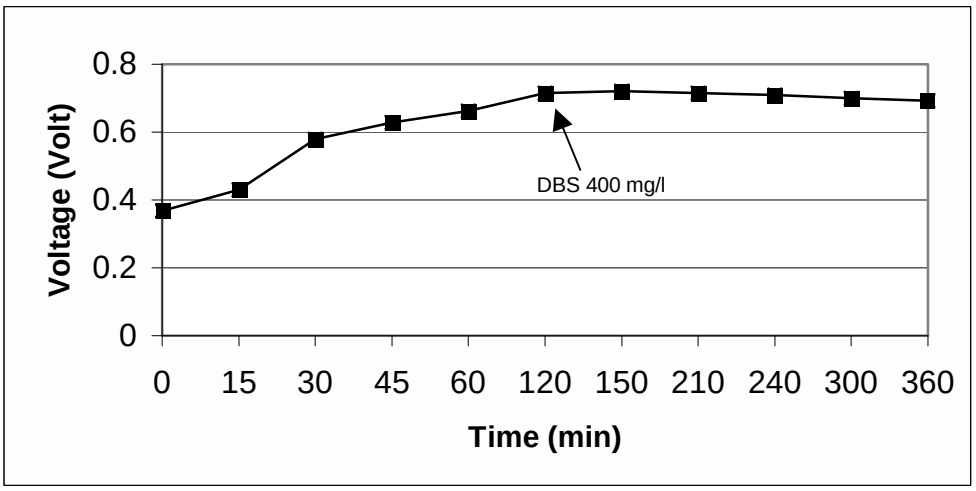


Figure 10.3 : Voltage drop across the anionic membrane (AMV) as a function of time (400 mg/ℓ DBS added).

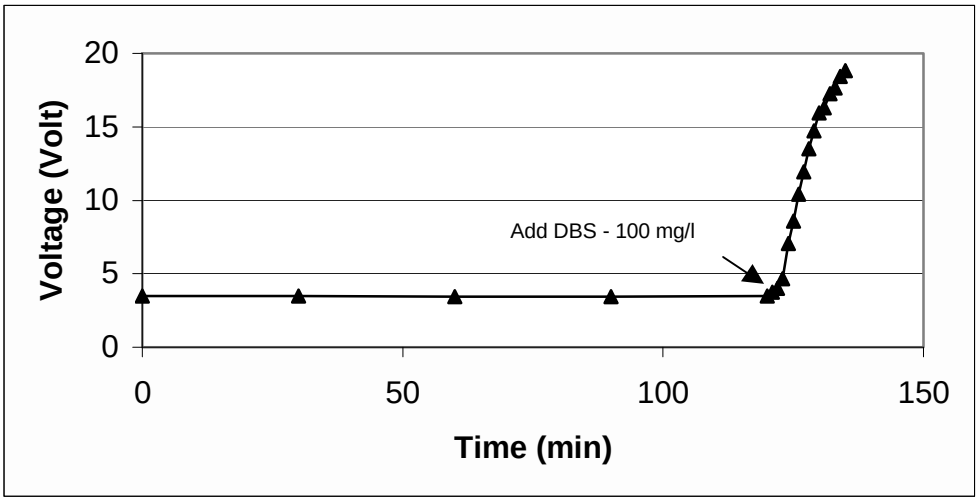


Figure 10.4 : Voltage drop across the anionic membrane (AMV) as a function of time (3 000 mg/ℓ NaCL, 100 mg/ℓ DBS added).

No increase in voltage drop was experienced across the anionic membrane after the addition of 400 mg/ℓ DBS (Figure 10.3). The reason for this may be that the negative charges of the foulants on the membrane surface are repelling the negative DBS anions. A significant increase in voltage drop, on the other hand, was experienced on the clean membrane surface (Figure 10.4). Therefore, the membrane fouling cell is effective in indicating membrane fouling caused by organic anions.

10.3 Fouling Potential of the Leachate for Tokuyama Soda ACS and CMS Membranes

The voltage drop across the anionic membrane (ACS) as a function of time is shown in Figure 10.5. A rapid increase in voltage drop across the anionic membrane has taken place, which shows that the membrane is severely fouled by the leachate.

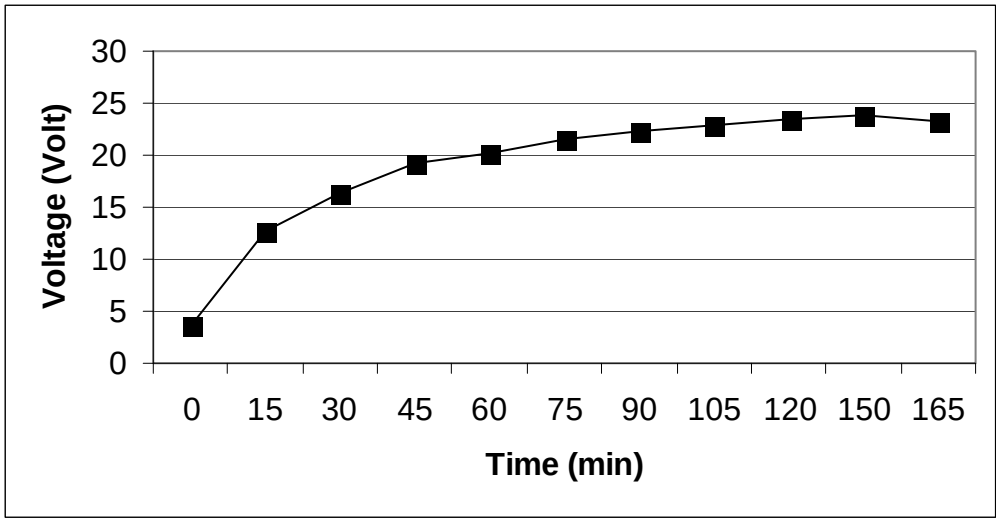


Figure 10.5 : Voltage drop across the anionic membrane (ACS) as a function of time.

The electrical resistances of the new, used and cleaned membranes are shown in Table 10.2.

Table 10.2 : Electrical resistance of the new, used and cleaned membranes.

Membrane resistance ohm.cm² (1)								
Run No.	New membranes		Used membranes from cell		Used membrane after equilisation		Used membrane after cleaning (2)	
	ACS	CMS	ACS	CMS	ACS	CMS	ACS	CMS
1	19,5	7,92	347,5	8,94	254,7	8,0	47,2	8,2

(1) Membranes were equilibrated overnight in 0,5 N NaCl prior to resistance measurements. The used membranes from the cell were not equilibrated.
(2) Soaked overnight in 10% NaCl at pH 10,5.

The membranes were previously used, therefore, the high initial resistance of the anionic membrane. It is clear from the resistance data that the anionic membrane is severely fouled by the leachate. Even soaking in a 10% NaCl solution at pH 10,5 could not restore membrane resistance. The cationic membrane, on the other hand, appeared not to be fouled by the leachate.

10.4 Fouling Potential of the Leachate for Ionics Anion AR204SZRA and CR67-HMR-412 Cation Membranes

The voltage drop across the anionic membrane (Ionics) as a function of time is shown in Figures 10.6 and 10.7. It appeared that some initial fouling took place in the beginning of run 1 (Figure 10.6), whereafter there was a decrease in voltage drop across the anionic membrane. A second run was conducted on the cleaned membranes from Run 1 and an increase in voltage drop was experienced as a function of time (Figure 10.7). Therefore, it appears that the Ionics anion membrane is fouled by the leachate.

The electrical resistances of the new, used and cleaned membranes are shown in Table 10.3.

Table 10.3 : Electrical resistance of the new, used and cleaned membranes.

Membrane resistance ohm.cm² (1)								
Run No.	New membranes		Used membranes from cell		Used membrane after equilisation		Used membrane after cleaning (2)	
	A	C	A	C	A	C	A	C
1	6,2	8,9	17,1	7,5	6,8	10,6	6,9	7,4
2	6,9	7,4	10,4	7,8	7,8	9,1	4,7	6,3

(1) Membranes were equilibrated overnight in 0,5 N NaCl. Used membranes from cell not equilibrated.
(2) Overnight soaking in 10% NaCl at pH10,5.

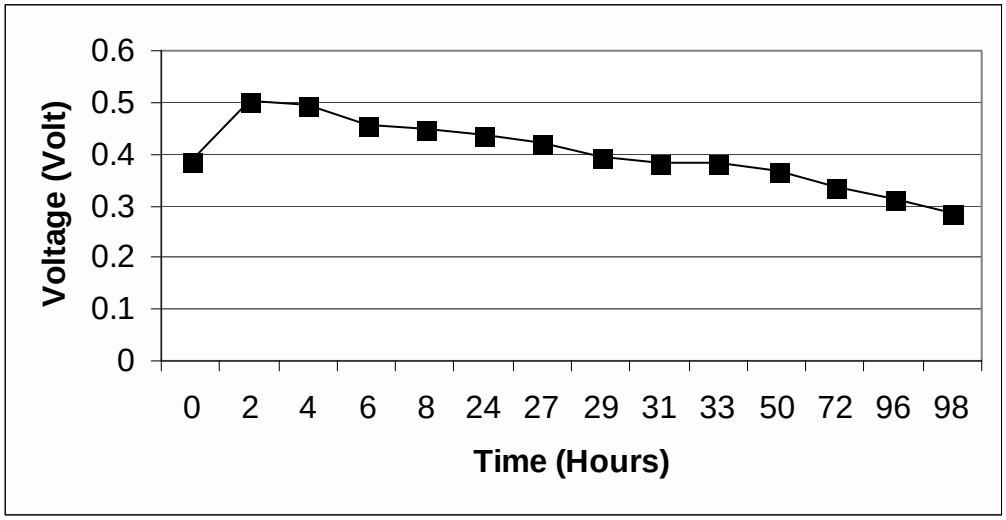


Figure 10.6 : Voltage drop across the Ionics anion membrane as a function of time (run 1).

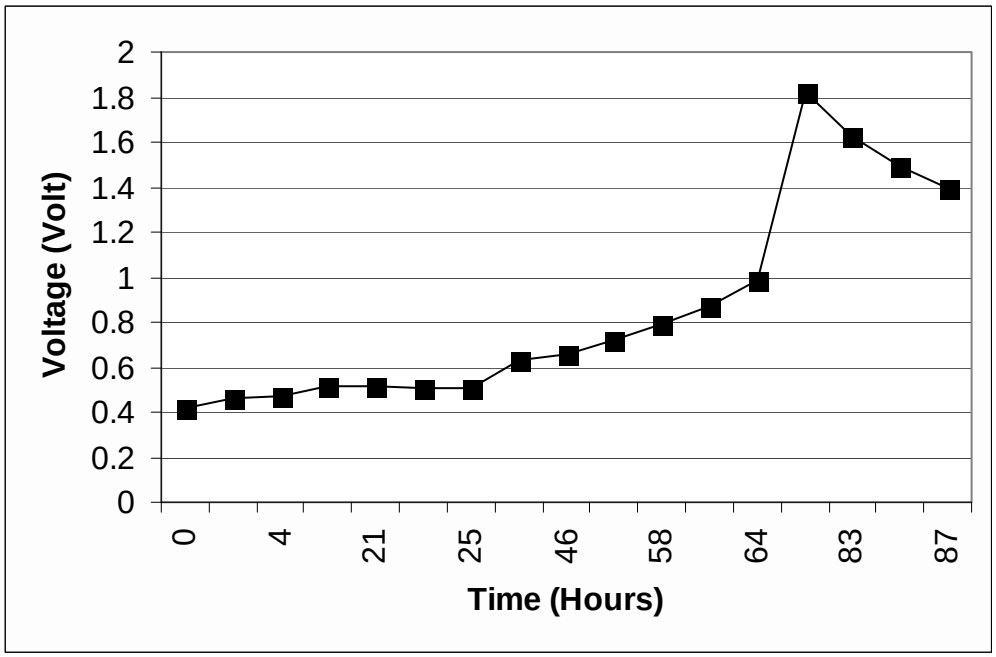


Figure 10.7 : Voltage drop across the Ionics anion membrane as a function of time (run 2).

Very little fouling, if any, was indicated by the membrane resistance measurements. The increase in potential drop across the anionic membrane (Figure 10.7) could most probably be ascribed to the deposition of a solid layer on the membrane surface that could be mechanically removed. Chemical cleaning has shown that membrane resistance could be restored.

10.5 Fouling Potential of the Leachate for Selemion ASV and CSV Membranes

The voltage drop across the anionic membrane (ASV) as a function of time is shown in Figures 10.8 and 10.9. Cleaned membranes from run 1 were used for run 2. An increase in potential drop across the anionic membrane was experienced right from the beginning in each case. Therefore, the membranes are easily fouled by the leachate. The electrical resistances of the new, used and cleaned membranes are shown in Table 10.4.

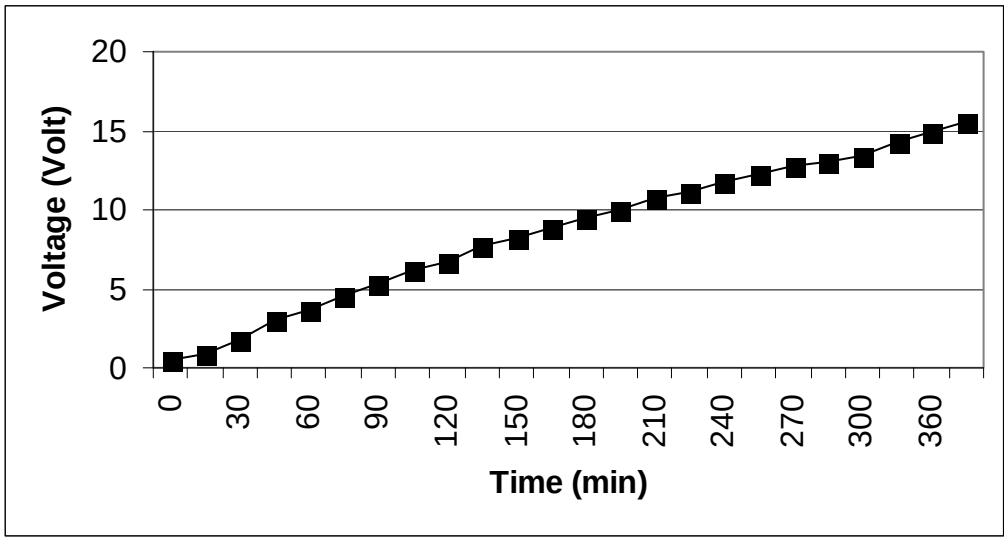


Figure 10.8 : Voltage drop across the anionic membrane as a function of time (run 1).

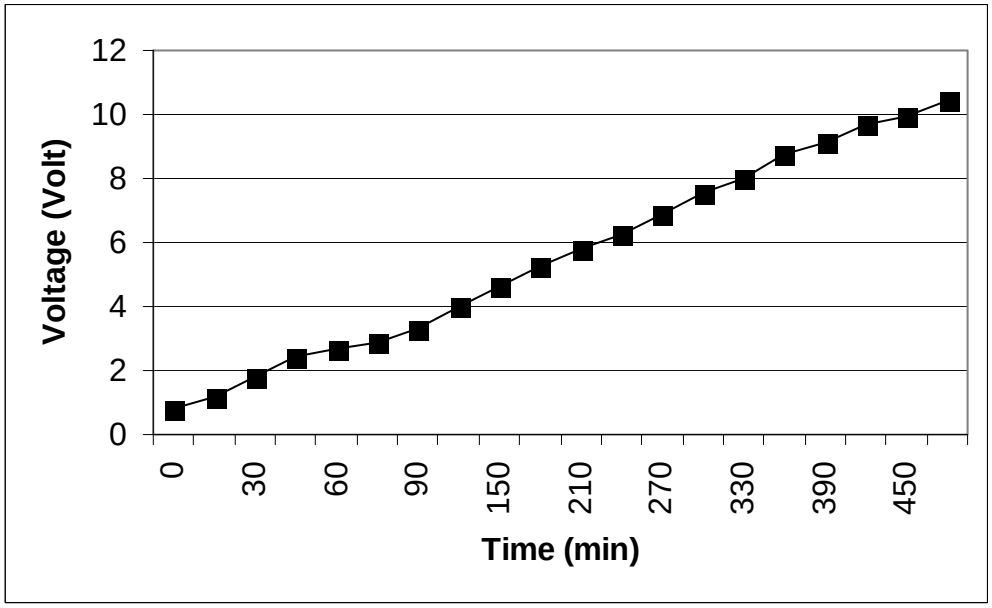


Figure 10.9 : Voltage drop across the anionic membrane as a function of time (run 2).

Table 10.4 : Electrical resistance of the new, used and cleaned membranes.

Membrane resistance ohm.cm² (1)								
Run No.	New membranes		Used membranes from cell		Used membrane after equilisation		Used membrane after cleaning (2)	
	ASV	CSV	ASV	CSV	ASV	CSV	ASV	CSV
1	4,7	8,9	97,6	7,1	15,1	8,6	5,8	7,1
2	5,8	7,1	27,7	12,1	6,5	5,1		

(1) Membranes were equilibrated overnight in 0,5 N NaCl. Used membranes from cell not equilibrated.

(2) Overnight soaking in 10% NaCl at pH 10,5.

The resistance measurements before and after cleaning showed very little membrane fouling. This, again, showed that the fouling layer on the membrane can be mechanically and chemically removed.

10.6 Fouling Potential of the Leachate for Tokuyama Soda AFN and Selemion CMV Membranes

The voltage drop across the anionic membrane (AFN) as a function of time is shown in Figure 10.10. No increase in voltage drop across the anionic membrane was experienced as a function of time.

The electrical resistances of the new, used and cleaned membranes are shown in Table 10.5.

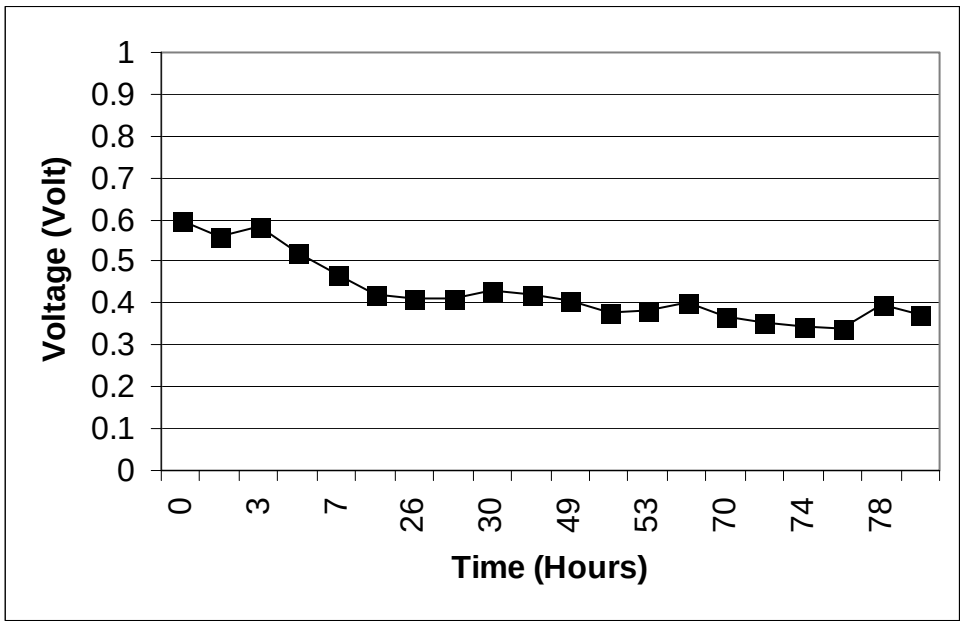


Figure 10.10 : Voltage drop across the anionic membrane as a function of time.

Table 10.5 : Electrical resistance of the new, used and cleaned membranes.

Membrane resistance ohm.cm² (1)								
Run No.	New membranes		Used membranes from cell		Used membrane after equilisation		Used membrane after cleaning (2)	
	AFN	CMV	AFN	CMV	AFN	CMV	AFN	CMV
1	5,9	6,4	7,9	5,1	4,6	5,6	2,1	4,1

(1) Membranes were equilibrated overnight in 0,5 N NaCl. Used membranes from cell not equilibrated.
(2) Overnight soaking in 10% NaCl at pH 10,5.

It appears that no or very little membrane fouling is experienced with the AFN membranes. These membranes are claimed to be more resistant to organic fouling than the other conventional ion-exchange membranes. Therefore, the AFN membrane is a candidate for the treatment of ISW leachate.

10.7 The Effect of Current Density on the Treatment of the Industrial Solid Waste Leachate with Tokuyama Soda AFN and Selemion CMV Membranes

The effect of current density on the potential drop across the anionic membrane (AFN) is shown in Figures 10.11 to 10.15. Very little increase in potential drop across the anion membrane was experienced at low current densities (20 and 40 mA/cm²) (Figures 10.11 and 10.12). However, an increasing potential drop was experienced across the anionic membrane at higher current densities (80, 100 and 120 mA/cm²) (Figures 10.13 to 10.15). This can be ascribed to membrane fouling. Most of this fouling was most probably caused by the deposition of solid material on the membrane surface. However, some fouling could also have been caused by soluble organics in the effluent which could have penetrated the membrane.

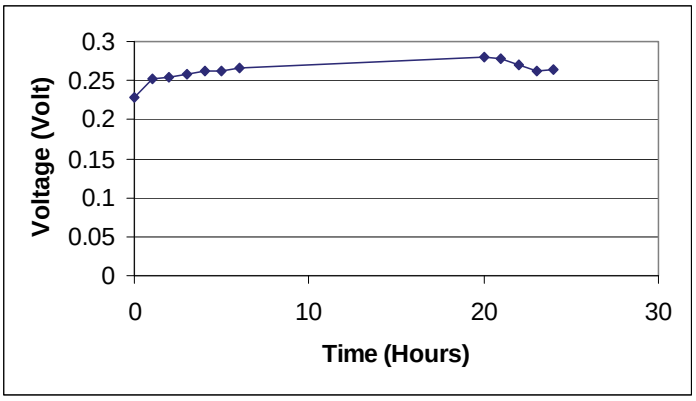


Figure 10.11 : Voltage drop across the anion membrane as a function of time (20 mA/cm^2).

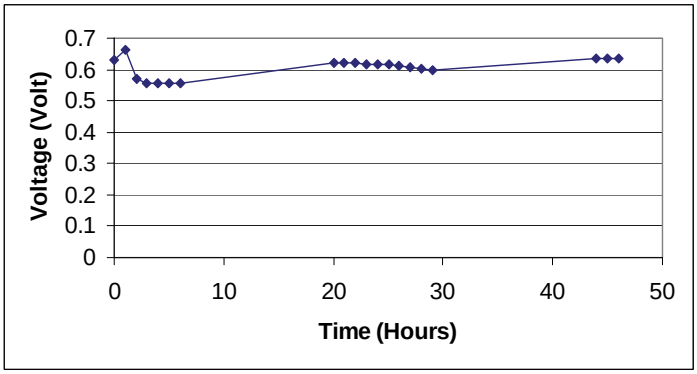


Figure 10.12 : Voltage drop across the anion membrane as a function of time (40 mA/cm^2).

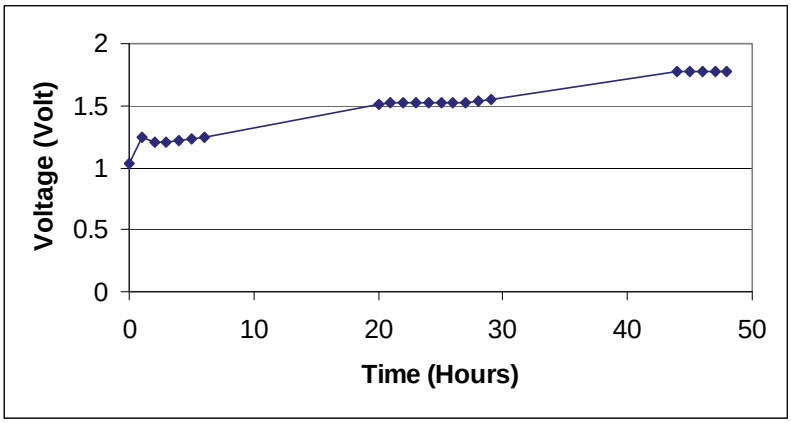


Figure 10.13 : Voltage drop across the anion membrane as a function of time (80 mA/cm^2).

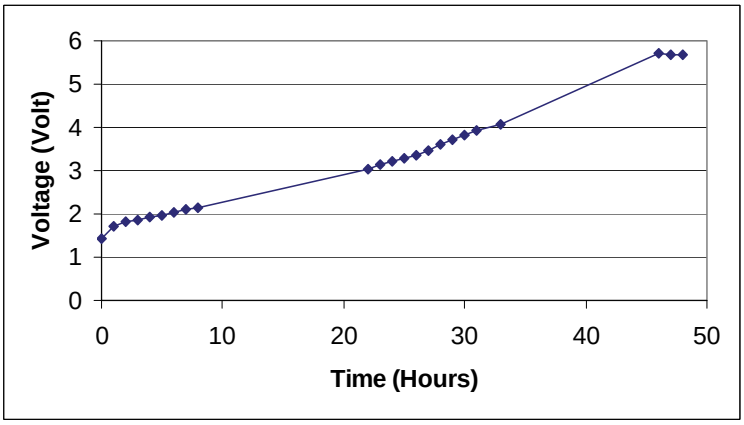


Figure 10.14 : Voltage drop across the anion membrane as a function of time (100 mA/cm^2).

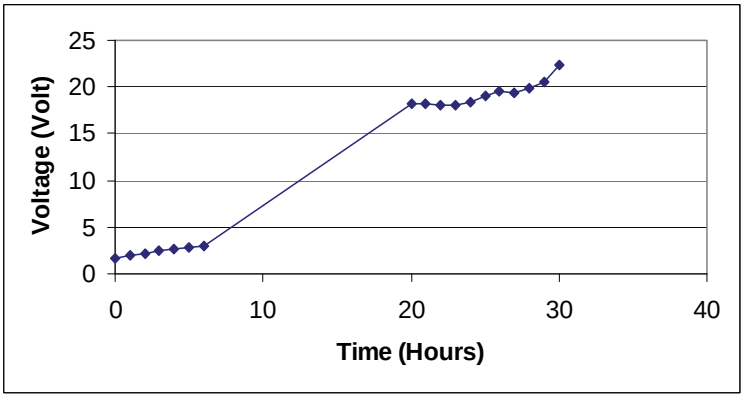


Figure 10.15 : Voltage drop across the anion membrane as a function of time (120 mA/cm^2).

Membrane resistance and the permselectivities of the membranes were also measured before and after fouling, and after membrane cleaning. The results are summarised in Tables 10.6 to 10.8.

The electrical resistance of the anionic (AFN) and cationic (CMV) membranes showed clearly that the original membrane resistance could be restored after cleaning (3% NaCl, at pH 10,5, overnight, Tables 10.6 and 10.7). It is also interesting to note that the salt cleaning solution removed material which had not been deposited on the membrane surface from the membrane. The material that was removed gave the cleaning solution a brownish colour.

The permselectivities of the membranes remained unchanged at the lower current densities (Table 10.8). However, there was a definite decrease in permselectivity at the higher current densities. This might be ascribed to the introduction of organic anions into the membrane matrix, making it less selective for anions.

Table 10.6 : Resistance of the AFN membranes after fouling at different current densities.

	Current densities (mA/cm²)				
	20	40	80	100	120
New Membrane (ohm.cm^2)	1,07	1,07	1,07	1,07	1,07
Used direct from cell (ohm.cm^2)	3,66	7,92	6,19	8,17	0,91
After Equilibration (0,5 N NaCl) (ohm.cm^2)	1,17	1,52	5,03	5,03	1,37
After Cleaning and Equilibration (ohm.cm^2)	0,96	1,22	1,02	0,66	0,86

Cleaning - 3% NaCl, pH 10,5 (overnight)

Table 10.7 : Resistance of the CMV membranes after fouling at different current densities.

	Current densities (mA/cm²)				
	20	40	80	100	120
New Membrane (ohm.cm^2)	2,13	2,13	2,13	2,13	2,13
Used direct from cell (ohm.cm^2)	2,84	2,59	2,59	8,78	1,57
After Equilibration (0,5 N NaCl) (ohm.cm^2)	2,08	2,23	3,1	3,1	2,18
After Cleaning and Equilibration (ohm.cm^2)	2,13	2,13	2,23	1,93	1,88

Cleaning - 3% NaCl, pH 10,5 (overnight)

Table 10.8 : Permselectivities (delta t) of the AFN membranes after fouling at different current densities.

	Current densities (mA/cm²)				
	20	40	80	100	120
New Membrane (Δt)	0,72	0,72	0,72	0,72	0,72
After Equilibration (0,5 N NaCl) (Δt)	0,76	0,76	0,74	0,70	0,64
After Cleaning and Equilibration (Δt)	0,72	0,74	0,74	0,68	0,62

Cleaning - 3% NaCl, pH 10,5 (overnight)

10.8 The Effect of Current Density on the Treatment of the Industrial Solid Waste Leachate with Tokuyama Soda AXE and CMX Membranes

The results are shown in Figures 10.16 to 10.23.

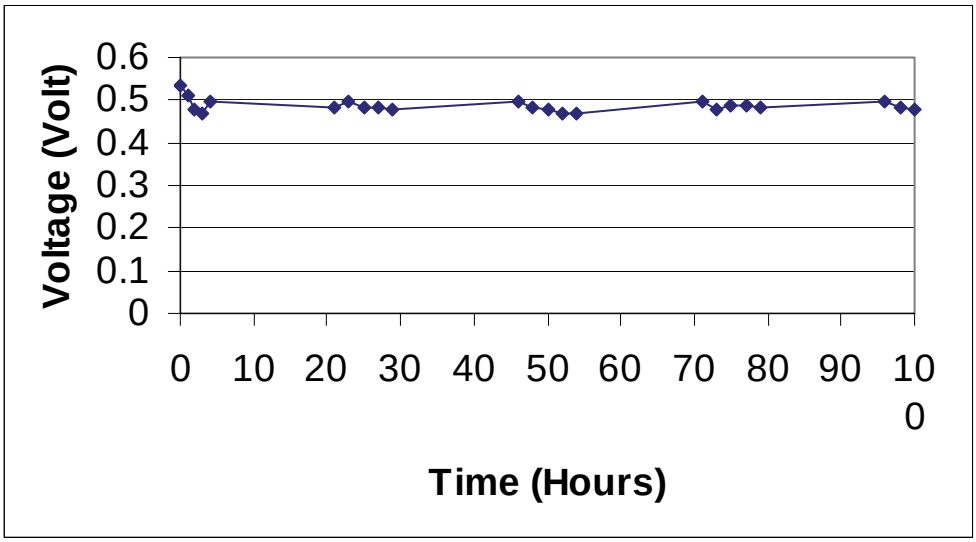


Figure 10.16 : Voltage drop across the anion membrane as a function of time (20 mA/cm^2).

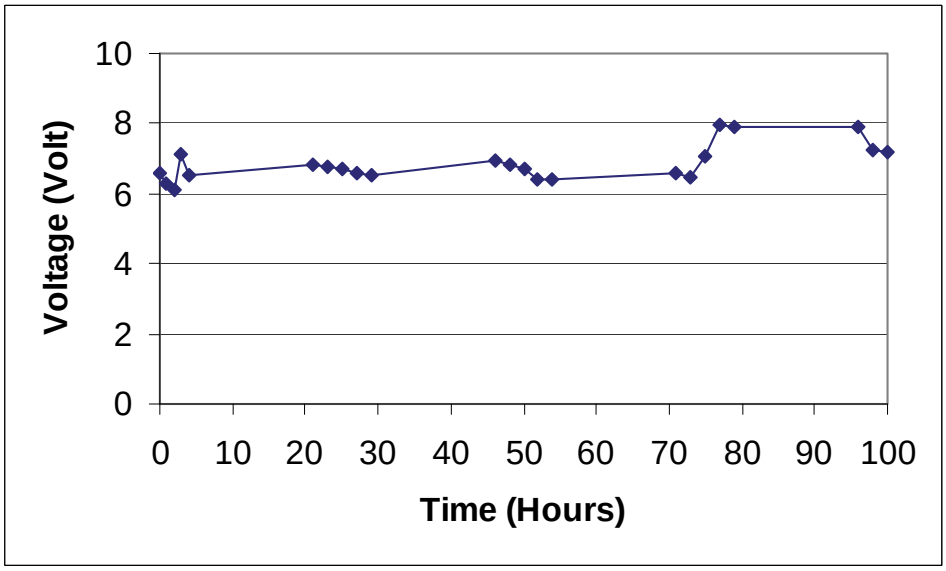


Figure 10.17 : Voltage drop across the fouling cell as a function of time (20 mA/cm^2)

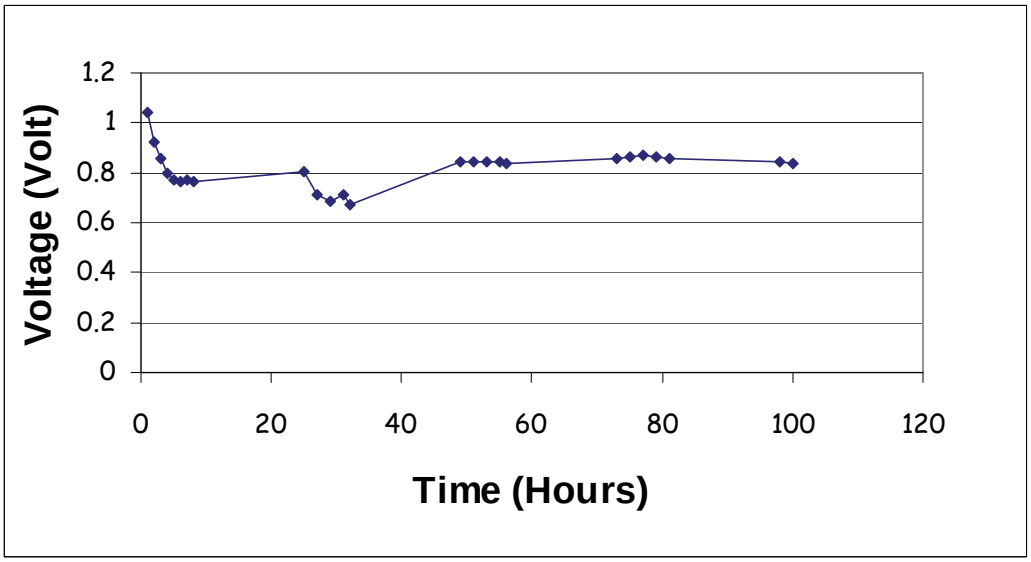


Figure 10.18: Voltage drop across the anion membrane as a function of time (40 mA/cm^2).

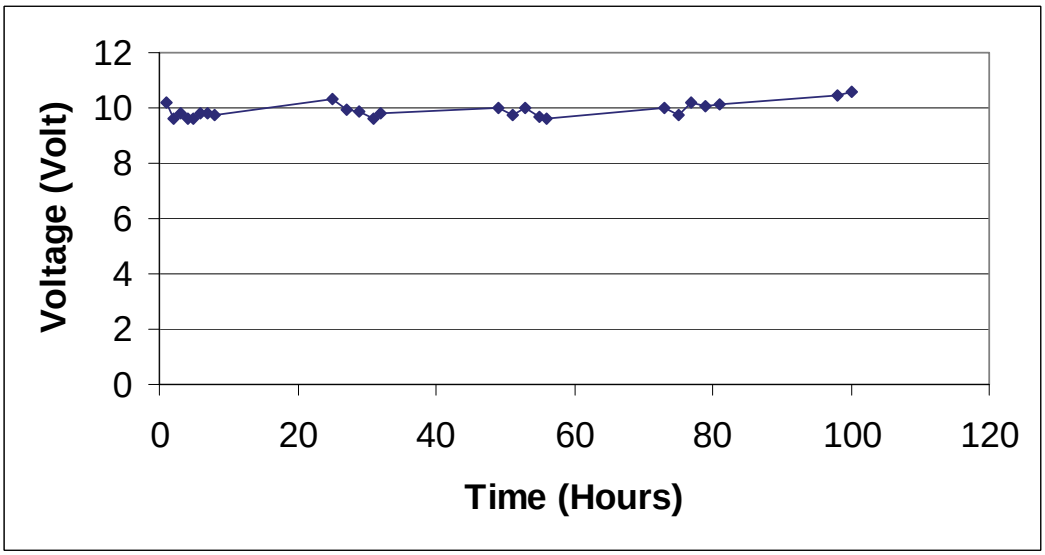


Figure 10.19: Voltage drop across the fouling cell as a function of time (40 mA/cm^2)

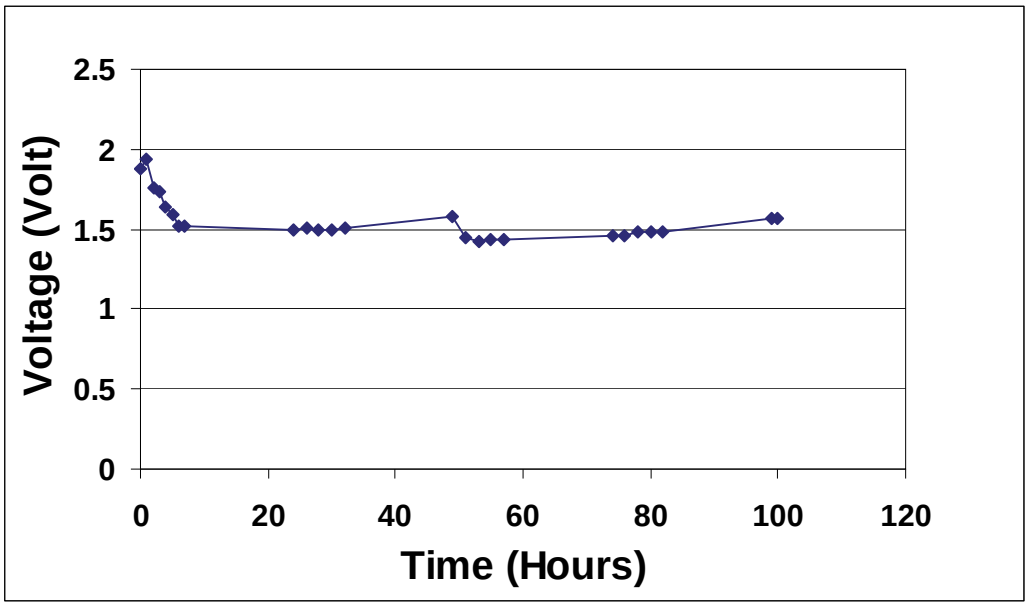


Figure 10.20: Voltage drop across the anion membrane as a function of time (80 mA/cm^2).

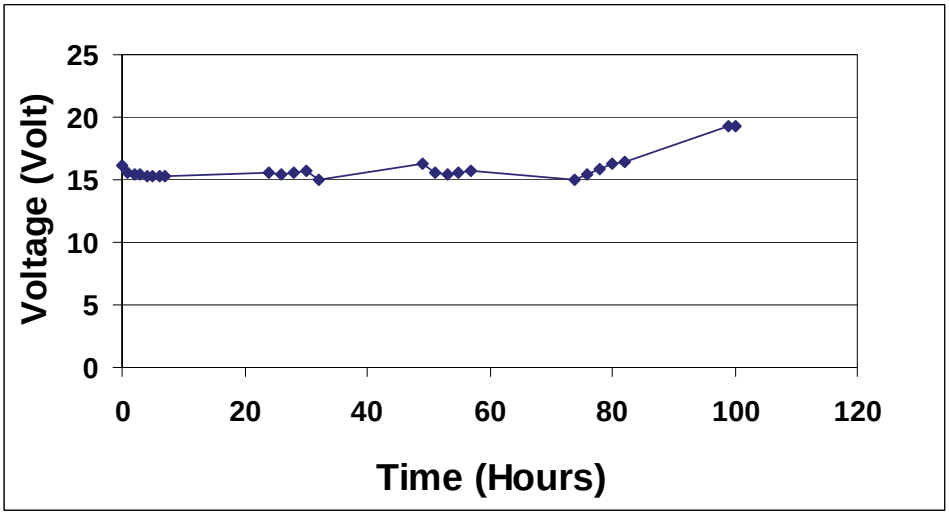


Figure 10.21: Voltage drop across the fouling cell as a function of time (80 mA/cm^2).

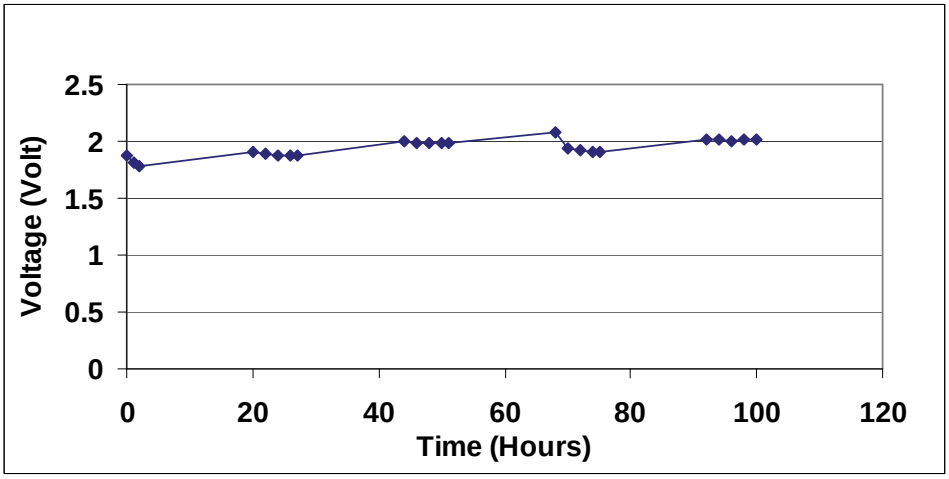


Figure 10.22: Voltage drop across the anion membrane as a function of time (100 mA/cm^2)

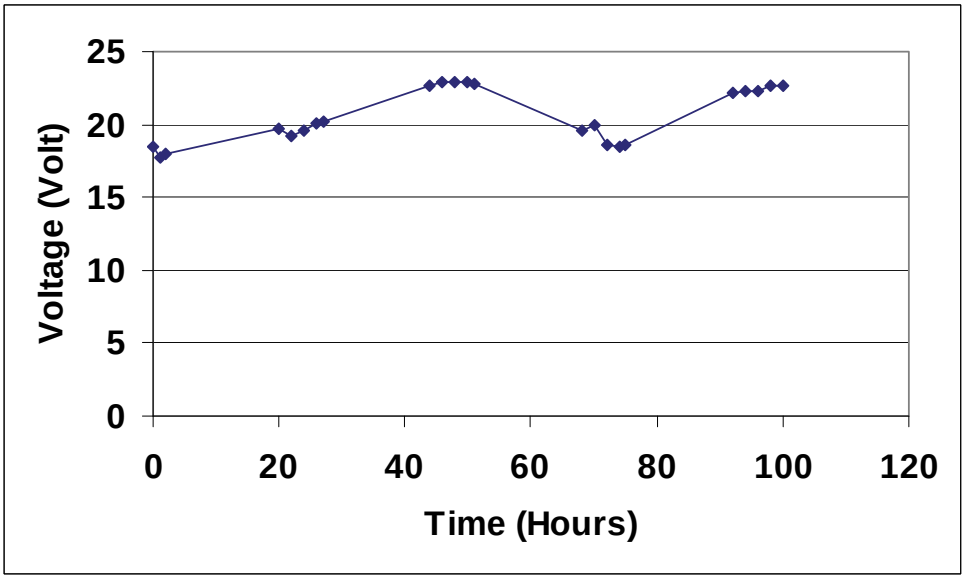


Figure 10.23: Voltage drop across the fouling cell as a function of time (100 mA/cm^2)

Membrane resistances and the permselectivities of the membranes were also measured before and after fouling, and after membrane cleaning. The results are summarized in Tables 10.9, 10.10 and 10.11.

Table 10.9 : Resistance of the AXE membrane after fouling at different current densities.

	Current density (mA/cm^2)			
	20	40	80	100
New membrane (ohm.cm^2)	1,0	1,2	2,8	—
Used, directly from cell (ohm.cm^2)	3,9	4,6	5,1	8,4
After equilibration (0,5 N NaCl) (ohm.cm^2)	3,6	2,3	1,0	0,7

Table 10.10 : Resistance of the CMX membrane after fouling at different current densities.

	Current density (mA/cm^2)			
	20	40	80	100
New membrane (ohm.cm^2)	1,6	1,4	4,7	—
Used, directly from cell (ohm.cm^2)	4,2	5,2	6,4	14,8
After equilibration (0,5 N NaCl) (ohm.cm^2)	4,2	5,4	3,7	3,7

Table 10.11 : Permselectivities (delta *t*) of the AXE membrane after fouling at different current densities.

	Current density (mA/cm^2)			
	20	40	80	100
New membrane (Δt)	0,54	0,72	0,73	—
After equilibration (0,5 N NaCl) (Δt)	0,78	0,60	0,28	0,86

No significant increase in voltage drop across the anion membrane was experienced as a function of time (Figures 10.16 to 10.23). It further appears that the resistance of the anionic membrane after equilibration in 0,5 N NaCl was higher (20 and 40 mA/cm^2) and lower (80 mA/cm^2) than the resistance of the new membrane (Table 10.9). The resistance at the higher current densities was not significantly higher than that of the new membrane. Therefore, the anionic membrane appears to be reasonably resistant towards fouling.

The resistance of the cationic membrane at higher current densities was also not significantly higher than that of the new membrane (Table 10.10). Therefore, the cationic membrane also appears to be reasonably resistant towards fouling.

The permselectivity of the anion membrane was, in one case, significantly lower than the permselectivity of the new membrane (Table 10.11). This could be ascribed to an experimental error, because the permselectivity of the membrane at the other current densities were not significantly lower than that of the new membrane. Therefore, the anion membrane also appears to be reasonably resistant to fouling.

The AFN anionic membrane was selected from all the other membranes tested for further long-term fouling tests in the fouling test cell, because this membrane appeared to be more resistant to fouling than the other membranes. This membrane is also commercially applied to de-ash industrial effluents containing high organic concentrations.

11. EVALUATION OF MEMBRANE CLEANING STRATEGIES OF THE FOULED ELECTRODIALYSIS MEMBRANES WITH COMMERCIALLY AVAILABLE MEMBRANE CLEANING AGENTS

Industrial solid waste leachate from Holfontein was pretreated with 200 g/ℓ Iscor ash, 12 g/ℓ soda ash, and 6,9 g/ℓ NaOH. The leachate was filtered through Whatman No. 1 filter paper, and the pH of the filtrate was adjusted to approximately 7 with carbon dioxide. Five litres of the pretreated leachate was circulated through the fouling cell as before (see 3.7, Figure 3.3), and the potential drop across the AFN anionic membrane was measured as a function of time (current density 100 mA/cm²). The pretreated leachate was replaced with fresh pretreated leachate after certain time intervals. The detailed results are shown in Appendix D.

The potential drop across the anionic membrane as a function of time is shown in Figure 11.1. The potential drop across the membrane was 2,6 volt in the beginning, and dropped to approximately 1,4 volt after 8 hours as a result of a temperature increase of the feed solution (Appendix D). The voltage then slowly started to increase from 8 hours to 31 hours (1,7 volt), remained at about 2,3 volt (46 to 55 hours),and then rapidly increased to 6,5 volt (120 hours). The cell was opened and the membranes were inspected. It was found that a slimy brownish layer of solid material had deposited on the feed side of the anionic membrane. The material was mechanically removed and the membrane potential had dropped to almost its initial value (1,7 volt). The membrane potential then decreased somewhat (1,3 volt, 122 hours), increased to approximately 2 volt (126 hours) and remained more or less constant until 257 hours (1,7 volt). The potential then increased to 2,74 volt (298 hours) and the polarity across the electrodes was reversed. This resulted in a drop in potential to 1,4 volt (approximate initial potential). The potential then increased after commencement of the run to 2,5 volt (300 hours), decreased somewhat and increased to 1,9 volt after 693 hours. The membranes were then soaked in 3% sodium chloride solution at pH 10,5 (overnight). Cleaning of the membranes with salt solution at high pH had the effect of reducing the potential to approximately 1 volt. The potential then increased to 1,8 volt after 719 hours. The fouling cell was not in operation for 48 hours and the potential was 0,6 volt when the run was started. Therefore, shutdown of the fouling cell appeared to have a definite effect on the membrane potential. The membrane potential then started to in crease and was 2,1 volt after 836 hours. An 8-day shutdown had the effect of reducing the potential to 1,2 volt when the run was commenced and the voltage drop was 1,4 volt when the run was terminated. Therefore, it appears that it should be possible to control membrane fouling with mechanical cleaning, polarity reversal and chemical cleaning.

Membrane resistances were measured at the end of the run. The resistances were 0,96 ohm.cm^{-2} (0,96 new) for the AFN, and 2,18 ohm.cm^{-2} (2,13 new) for the CMX membrane. Therefore, it appears that almost no change had taken place in membrane resistance over an 908 hour period.

The ED run has demonstrated that severe fouling should not be experienced with the pretreated leachate, even at high current density. However, regular membrane cleaning will be necessary to control membrane fouling.

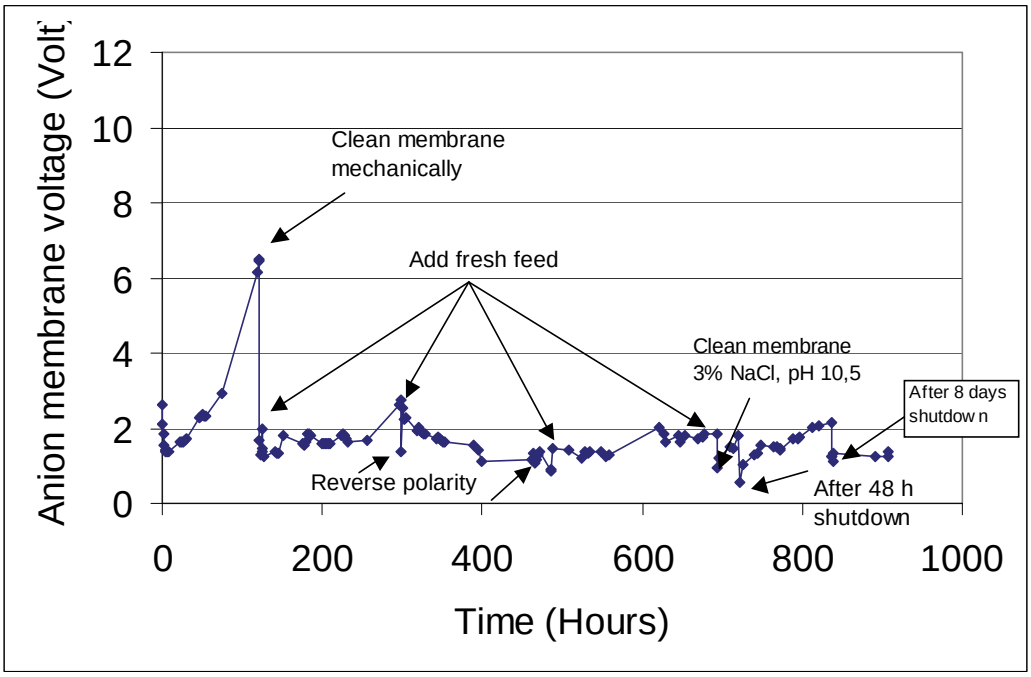


Figure 11.1 : Potential drop across the anionic membrane (AFN) as a function of time.

12. EVALUATION OF ELECTRODIALYSIS FOR THE DESALINATION / CONCENTRATION OF THE PRETREATED INDUSTRIAL SOLID WASTE LEACHATE

12.1 Introduction

Electrodialysis was conducted in the batch mode in a laboratory-scale ED unit to evaluate the desalination/concentration performance of ED for the ISWL from Holfontein (see 3.9 for experimental procedure), and to develop preliminary process design criteria for a full-scale application. A total of 10 runs were conducted on the pretreated feed. The electrical current, electrical conductivity, cell pair resistance and current density were determined as a function of time.

Electrodialysis with salt solution (approximately 3 000 mg/ℓ and sodium chloride solution simulating the concentration of the pretreated feed) was conducted before, during and after the runs to determine the current efficiency (see 3.9).

Membrane characteristics like membrane resistance, membrane permselectivity, ion-exchange capacity and gel water content were conducted before and after the runs to evaluate membrane fouling (see 3.9).

Electrodialysis runs were also conducted at low pH (approximately 4) and high pH (approximately 10) to determine the effect of pH on ED performance. The chemical composition of the raw leachate, pretreated leachate, ED product and brine was also determined. Polarisation curves were established to determine membrane polarisation characteristics. The toxicity of the raw leachate, pretreated leachate, ED product and brine was also determined (see 3.10). The preliminary economics of the process was derived from the experimental results (see 13.)

12.2 Electrodialysis Treatment of the Pretreated Leachate with Tokuyama Soda AFN and CMX Membranes

Ten ED runs were conducted on the pretreated feed. Electrodialysis runs were also conducted with salt solution before, during and after the batch runs on pretreated feed to determine membrane current efficiency. The results from a typical ED batch run on pretreated feed are shown in Table 12.1 and Figures 12.1 to 12.4.

The demineralised feed and brine concentration as a function of time during ED batch desalination is shown in Figure 12.1. The brine becomes more concentrated as desalination proceeds. Electrodialysis feed was concentrated from approximately 97 mS/cm to approximately 139 mS/cm at the end of the run (Table 12.1). The feed was desalinated from approximately 97 mS/cm to 7,6 mS/cm. Therefore, approximately 92,2% of the salinity was removed from the feed.

It should be noted that the current density was high at the constant cell pair voltage of 0,5 volt per cell pair that was applied (Table 12.1). The high current density that resulted was a result of the high salt concentration of the ED feed.

Table 12.1 : Experimental conditions and results of a typical batch ED run conducted on the pretreated leachate (run 4, pretreated sample 19/10/2000, 200 g/ℓ Iscor ash, 12 g/ℓ soda ash, 6,9 g NaOH, pH approximately 7 with CO_2).

Time	Total V	Volt (9 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	9 cp Resis	Current density
(min)	(V)	(V)	(Ampere)	(mS/cm)	° C	pH	(mS/cm)	° C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm²)
0	24,41	4,53	6,15	96,9	24,3	7,25	96,9	24,3	7,25	80,2	9,14	3,97	0,74	75,93
15	23,92	4,5	6,43	90,8	26,7		110,9	27,5				3,72	0,70	79,38
30	24,46	4,48	6,93	85,5	28,7		119,7	29,5				3,53	0,65	85,56
45	25,99	4,33	7,21	81,9	30		126,7	30,5				3,60	0,60	89,01
60	25,93	4,43	7,6	75,2	32,1		133,8	32,8				3,41	0,58	93,83
75	25,96	4,57	7,66	68,9	33,6		140,5	33,8				3,39	0,60	94,57
90	25,47	4,59	7,64	62,9	34,7		143,2	34,8				3,33	0,60	94,32
105	24,95	4,55	7,5	56,9	35,6		146,9	35,6				3,33	0,61	92,59
120	24,95	4,54	7,5	49,3	36,7		147,8	36,6				3,33	0,61	92,59
135	24,95	4,48	7,5	42,1	37,6		148,2	37,5				3,33	0,60	92,59
150	24,95	4,47	7,46	35,3	37,9		147,8	37,9				3,34	0,60	92,10
165	24,97	4,55	7,15	28,6	38,4		146	38,3				3,49	0,64	88,27
180	25,03	4,49	6,3	21,1	38,3		143,9	38,4				3,97	0,71	77,78
195	24,90	4,75	3,59	12,6	38,3		141,2	38				6,94	1,32	44,32
210	24,96	4,68	2,52	9,9	37,5		140	37,4				9,90	1,86	31,11
225	25,01	4,75	1,58	7,6	34,3	6,44	138,9	36,1	7,63	8,84	7,55	15,83	3,01	19,51
	25,05													

Feed start volume 4 ℓ Feed end volume 2,5 ℓ
Brine start volume 1 ℓ Brine end volume 2,4 ℓ

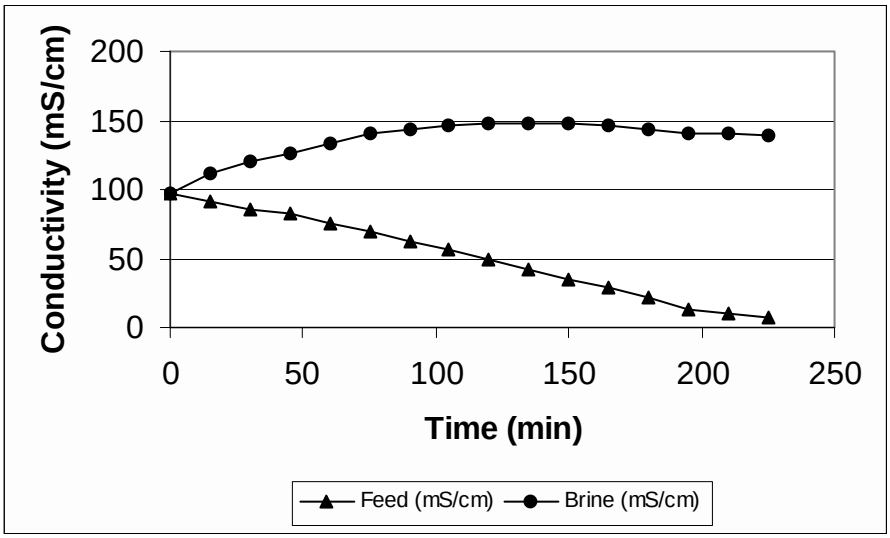


Figure 12.1 : Electrical conductivity of the ED feed (product) and brine as a function of time.

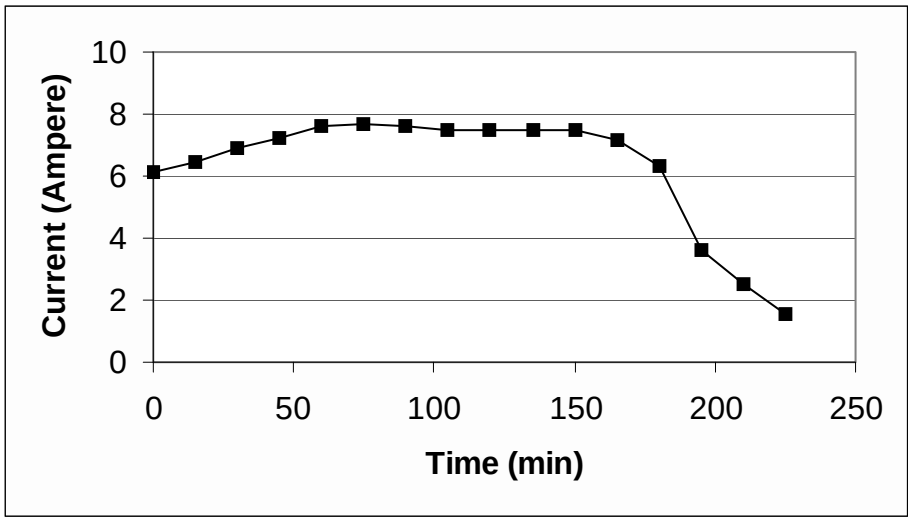


Figure 12.2 : Electrical current as a function of time.

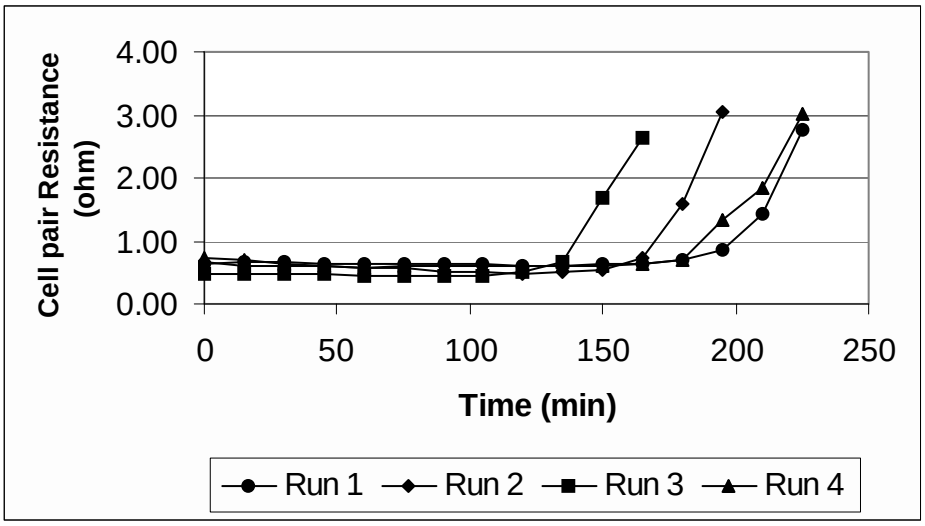


Figure 12.3 : Cell pair resistance as a function of time for different runs.

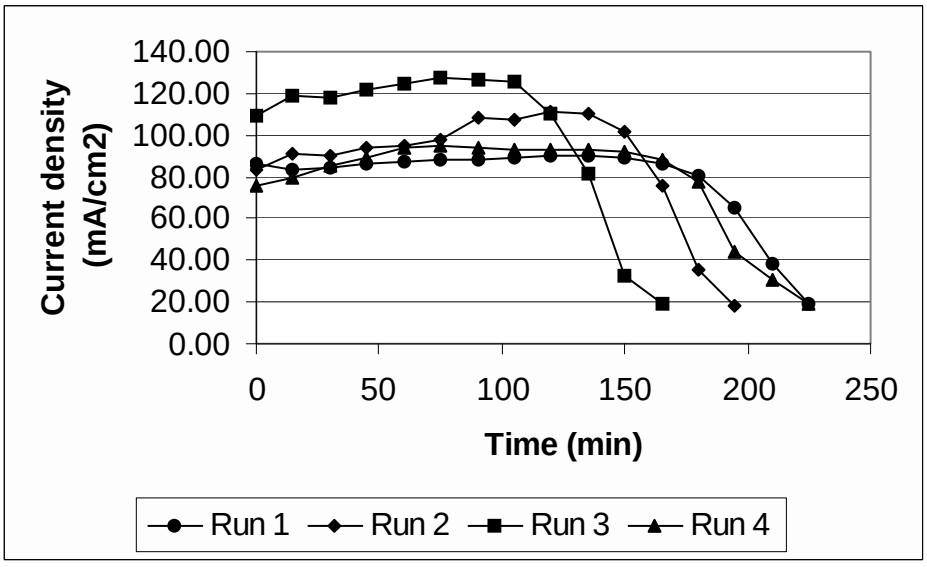


Figure 12.4 : Current density as a function of time for different runs.

The electrical current as a function of time is shown in Figure 12.2. The current first increased and then decreased as the resistance across the stack increased. The initial increase in current can be ascribed to a reduced resistance across the stack as a result of an increased concentration of salt in the brine compartments.

Cell pair resistance as a function of time for the first four runs is shown in Figure 12.3. Cell pair resistance remained more or less constant, showing that severe membrane fouling was not taking place.

The current density as a function of time for the first four runs is shown in Figure 12.4. The current density remained constant during most part of the batch desalination, and then decreased as most of the ions were removed from the diluating compartments.

The chemical composition of the raw leachate, treated leachate, ED product and ED brine is shown in Table 12.2. TDS removal was 85,5%. The TDS of the ED feed was reduced from 118 485 to 17 236 mg/ℓ. Excellent chloride (98,1%), sodium (91,2%) and potassium (97,2%) removals were obtained. Ammonia-nitrogen (74,7%), sulphate (68,7%), calcium (88,4%) and magnesium (92,5%) removals were also good.

Table 12.2 : Chemical composition of the raw feed, treated feed and ED product and brine (run 4, pretreated with 200 g/ℓ Iscor ash, 6,9 NaOH and 12 g/ℓ Na_2CO_3, pH approximately 7 with CO2).

Constituents*	Raw Feed	Treated feed	Product	Brine	Removal %
Conductivity (mS/cm)	94,2	96,9	7,6	138,9	92,16
pH	7,43	7,25	6,44	7,63	
Chemical oxygen demand (COD)	66 500	54 800	28 100	68 900	48,72
BOD_5 as O_2	16 500		5 000		
Total organic Carbon	13 765	14 409	4 112	21 611	71,46
Alkalinity as CaCO3	7 460	9 610	3 447	36 998	64,13
Total dissolved solids	116 595	118 485	17 236	215 980	85,45
Ammonia Nitrogen as N	2 049	1 558	394	2 278	74,71
Chloride	31 030	31 210	592	55 290	98,10
Sulphate	25 259	19 705	6 169	32 860	68,69
Calcium	1 040	103	12	65	88,35
Magnesium	995	663	50	1 070	92,46
Sodium	23 700	28 600	2 520	60 200	91,19
Potassium	7 090	6 420	183	12 100	97,15
Barium	0,51	0,48	0,06	0,32	87,50
Strontium	6,45	1,92	0,18	2,28	90,63
Phenolic compounds	289	17,6	13	22,2	26,14

* Concentration in mg/ℓ, unless stated otherwise

The treated effluent water recovery was determined at 62,5% (2,5 ℓ treated feed and 4 ℓ pretreated feed). The brine volume that was produced was determined at 35% (1,4 ℓ brine) of the pretreated feed (4 ℓ) (Table 12.1).

COD and TOC removals of 48,7 and 71,5% were obtained, respectively. Therefore, a significant amount of organics permeate the ED membrane. Excellent phenol removal (93,9%) was obtained with pretreatment of the leachate. The ED product, however, still contains some phenols.

The cell pair resistance as a function of time for all 10 runs that were conducted is shown in Figure 12.5. The cell pair resistance remained more or less constant for all the runs indicating that severe membrane fouling should not be a problem.

Run 8 was conducted at a feed pH of 3,9 (low pH) while run 9 was conducted at a feed pH of 9,6 (high pH). A somewhat better desalination / concentration performance was observed at the lower pH.

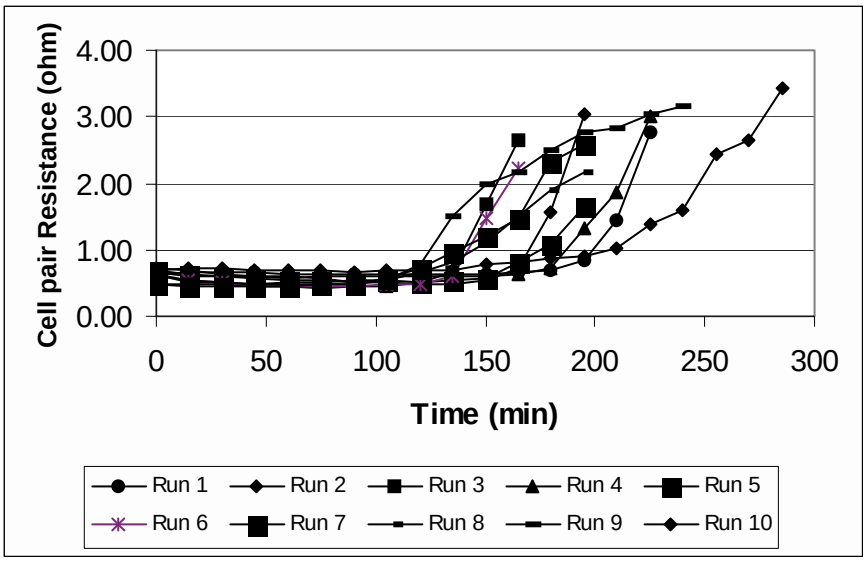


Figure 12.5 : Cell pair resistance as a function of time for 10 ED runs.

12.3 Membrane Current Efficiency Before, During and After the ED Runs

Membrane current efficiency before, during and after the ED runs are shown in Table 12.3. The current efficiency before the runs were started was determined at 83,9% (Na) and 107,6% (NaCl) on a salt solution with a concentration of 88,2 mS/cm. The current efficiency after 9 runs was 70,9% (Na) and 93,5% (NaCl). Therefore, there was some reduction in current efficiency that could be ascribed to membrane fouling.

Current efficiency measurements with an approximately 3 000 mg/ℓ salt solution show almost no reduction in current efficiency before and after 9 ED runs. Current efficiency calculated from the chemical composition after run 4 showed an efficiency of 61,7% (cations) and 63,8% (anions).

Table 12.3 : Membrane current efficiency before, between and after the batch ED runs.

Run No. [1]	CE % (3 000 mg/ℓ NaCl)	CE % (88,2 and 94 mS/m)
Before runs	**(80,6 - 89,6 mS/m)**	**83,9 (Na) 107,6 (NaCl)**
Run 1		
Run 2		
	94,4	
Run 3		
Run 4 [2]	61,7 (cations), 63,8 (anions)	
Run 5		
	92,3	
Run 6		
Run 7		
	99,1, after cleaning 103,3 [3]	
Run 8		
	97,1	
Run 9		
	101,2, after cleaning 102,9[3]	
After runs 1 to 9		70,9 (Na), 93,5 (NaCl)

(1) ED runs on pretreated sample.
(2) CE determined from chemical composition.
(3) 3% NaCl at pH 10,5.

12.4 Membrane Characteristics Before and After ED Treatment

The membrane characteristics before and after ED treatment are shown in Table 12.4. The membrane resistance and permselectivity of the anion membrane (AFN) before and after ED treatment (cleaning) were about the same. Therefore, it appears that these membrane characteristics changed very little. However, there was a more significant change in the ion-exchange capacity and gel water content characteristics of the anion membrane before and after ED treatment. This indicates a certain degree of membrane fouling.

Table 12.4 : Membrane characteristics bore and after ED treatment.

	Resistance ohm.cm^2		Permselectivity		Ion-exchange capacity meq/dry g		Gel water content wt%	
	AFN	**CMX**	**AFN**	**CMX**	**AFN**	**CMX**	**AFN**	**CMX**
New membrane	0,81	2,49	0,63	1,07	2,71	2,05	42,87	23,21
Used membrane	1,27	2,64	0,64	1,07	2,32	2,14	36,27	22,90
Membrane[1] after cleaning	1,12	-	0,62	-	2,21	-	31,49	-

(1) Cleaning with 3% NaCl at pH 10,5.

12.5 Polarisation Curves

Polarisation curves were established at different concentrations. The results are shown in Figures 12.6 to 12.9. An inflection point was only detected at the lowest concentration at approximately 27 volt (Figure 12.9). Therefore, ED should be conducted at 80% of this voltage so as not to exceed the limiting current density. The other concentrations were too high for polarisation to occur.

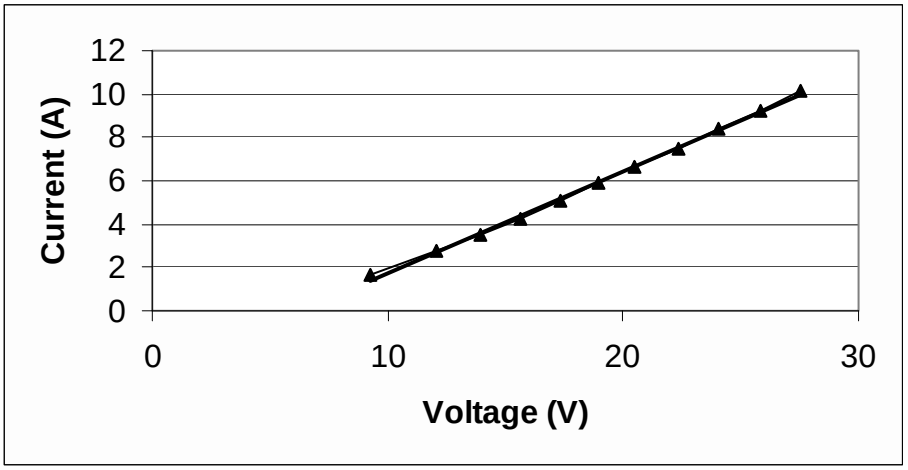


Figure 12.6 : Electrical current vs stack voltage (97,4 mS/cm).

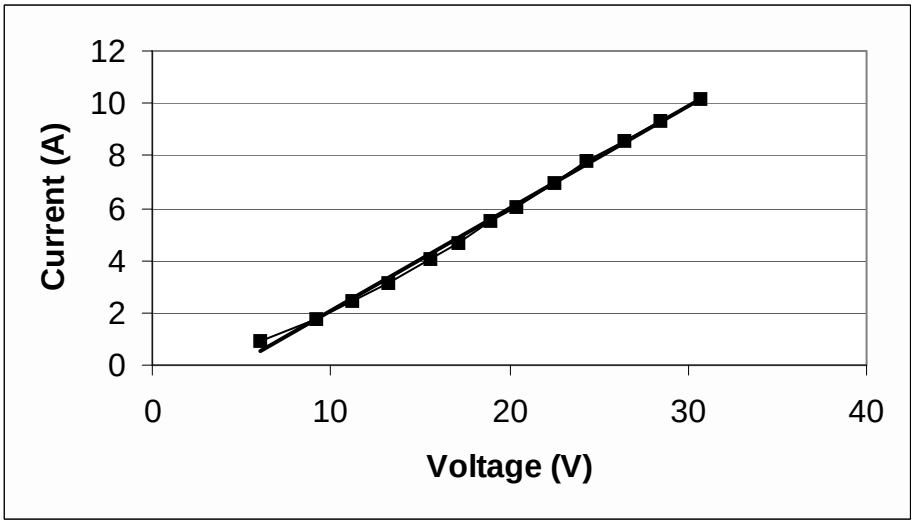


Figure 12.7 : Electrical current vs stack voltage (48,9 mS/cm).

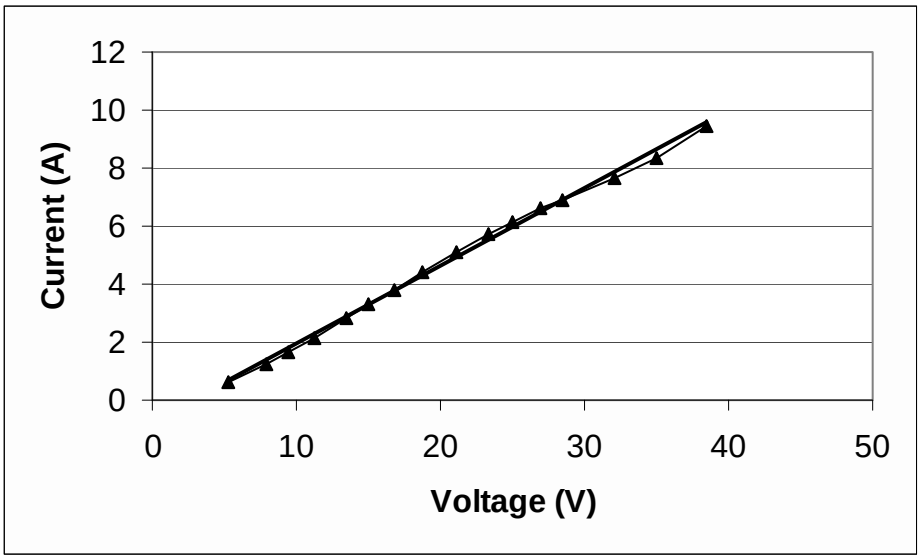


Figure 12.8 : Electrical current vs stack voltage (24,5 mS/cm).

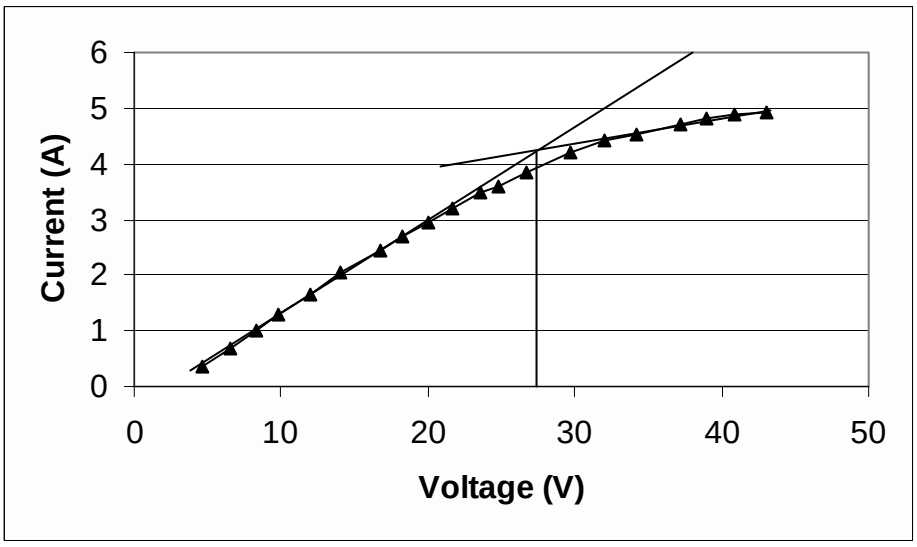


Figure 12.9 : Electrical current vs stack voltage (9,7 mS/cm).

12.6 Electrodialysis Treatment of the Pretreated Leachate with Tokuyama Soda AXE and CMX Membranes

Eight runs were conducted on the pretreated feed with the Tokuyama soda AXE and CMX membranes. The results from a typical ED batch run on pretreated feed are shown in Table 12.5 and Figures 12.10 to 12.13.

A better desalination/concentration performance was obtained with the AFN and CMX membranes than with the AXE and CMX membranes (Tables 12.1 and 12.5, Figures 12.1 and 12.10, Figures 12.2 and 12.11, Figures 12.3 and 12.12, Figures 12.4 and 12.13). Desalination was much faster with the AFN and CMX membranes than with the AXE and CMX membranes. A better quality product could be produced with the AFN and CMX membranes (7,6 mS/m vs 17,1 mS/m). A higher brine concentration could also be produced

with the AFN and CMX membranes (148 mS/m vs. 133 mS/m). Therefore, the AFN and CMX membranes were selected for treatment of leachate with ED.

Table 12.5 : Experimental conditions and results of a typical batch ED run conducted on the pretreated leachate using AXE and CMX membranes (run 5, pretreated sample 14/09/2001, 200 g Iscor ash, 12 g/ℓ soda ash, 6,9 g NaOH).

Time	Total V	Volt (9 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Cell Resist	9cp Resist	Current Density
(min)	(Vt)	(Vcp)	Ampere	(mS/cm)	(°C)	(pH)	(mS/cm)	(°C)	(pH)	(ohm)	(ohm)	(mA/cm^2)
0	20	4,5	3,67	102,4	20	7,32	91,3	20	7,43	5,45	1,23	45,31
15	20	4,5	4,2	97,6	21,8		98,5	22,3		4,76	1,07	51,85
30	20	4,5	4,32	92,8	22,8		100	23,3		4,63	1,04	53,33
45	20	4,5	4,76	86,6	25		108,1	25		4,20	0,95	58,77
60	19	4,5	4,86	82,2	25,6		113,4	26,6		3,91	0,93	60,00
75	19	4,5	4,9	76,3	27,1		118,4	27,8		3,88	0,92	60,49
90	19	4,5	4,75	71	28,1		122,7	29		4,00	0,95	58,64
105	18	4,5	4,57	65,8	29,3		125,2	29,7		3,79	0,98	56,42
120	18	4,5	4,34	63	29,4		127,4	30,3		4,15	1,04	53,58
135	18	4,5	4,16	59,4	29,7		129,4	30,7		4,33	1,08	51,36
150	17	4,5	3,86	55,9	29,7		132,9	30,5		4,40	1,17	47,65
165	16	4,5	3,57	52,9	30		133	30,6		4,48	1,26	44,07
180	16	4,5	3,25	50,5	30,2		133	30,7		4,92	1,38	40,12
195	16	4,5	3,03	48,9	29,9		133	30,7		5,28	1,49	37,41
210	16	4,5	2,83	45,7	30,4		133	31		5,65	1,59	34,94
225	16	4,5	2,71	44,3	29,3		132	30,5		5,90	1,66	33,46
240	15	4,5	2,55	42,3	29,7		133	30,5		5,88	1,76	31,48
255	15	4,5	2,52	41,5	29,6		133	30,2		5,95	1,79	31,11
270	15	4,5	2,49	39,9	29,2		132	30,1		6,02	1,81	30,74
285	15	4,5	2,49	37,6	29,4		132	30		6,02	1,81	30,74
300	16	4,5	2,57	36	28,7		132	30,1		6,23	1,75	31,73
315	16	4,5	2,63	33,4	29,2		128	29,7		6,08	1,71	32,47
330	16	4,5	2,75	31,1	29,9		128,5	30,2		5,82	1,64	33,95
345	16	4,5	2,8	28,5	29,8		127,2	30,3		5,71	1,61	34,56
360	16	4,5	2,61	25,5	29,9		126,3	30,5		6,13	1,72	32,22
375	16	4,5	2,18	23,1	30,1		125,9	30,6		7,34	2,06	26,91
390	16	4,5	2,01	21,5	29,3		125,9	29,9		7,96	2,24	24,81
405	16	4,5	1,41	19,5	29,7		125,1	30,1		11,34	3,19	17,41
420	16	4,5	1,15	18	30		123,9	30,3		13,91	3,91	14,20
435	15	4,5	0,97	17,1	28,8		124,4	29,6		15,46	4,64	11,98

Feed start volume = 4 ℓ Feed end volume = 2,75 ℓ

Brine start volume = 1 ℓ Brine end volume = 2,25 ℓ

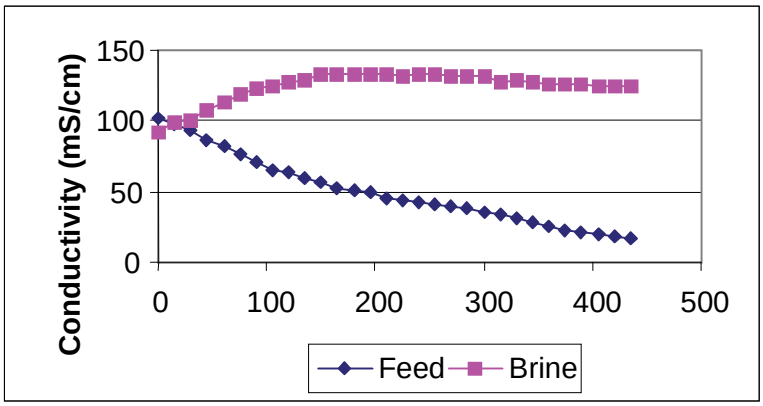


Figure 12.10 : Electrical conductivity of the ED feed (product) and brine as a function of time.

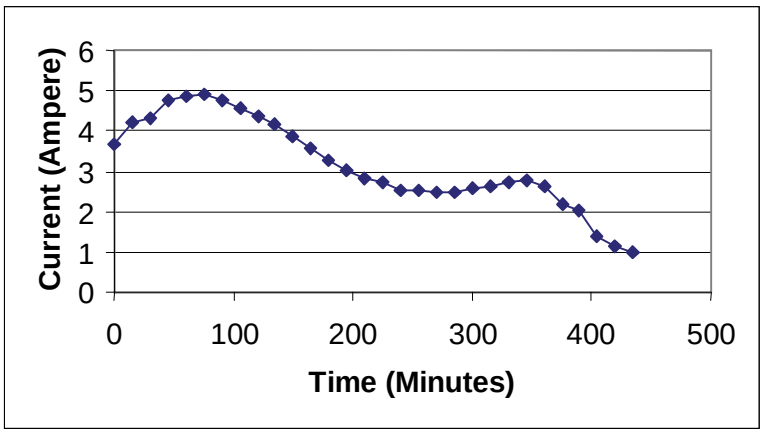


Figure 12.11 : Electrical current as a function of time.

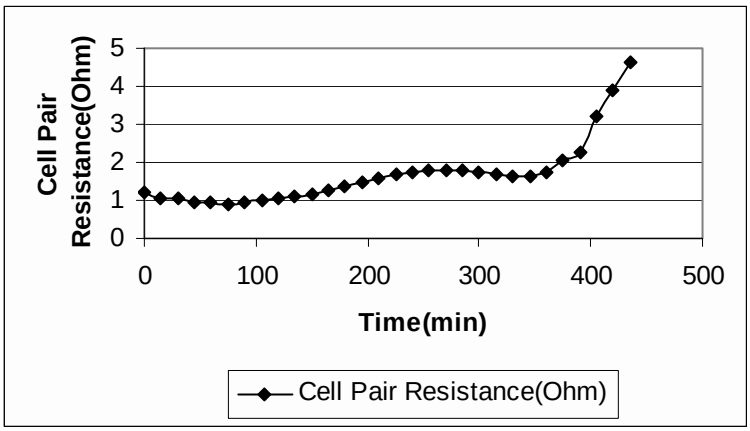


Figure 12.12 : Cell pair resistance as a function of time.

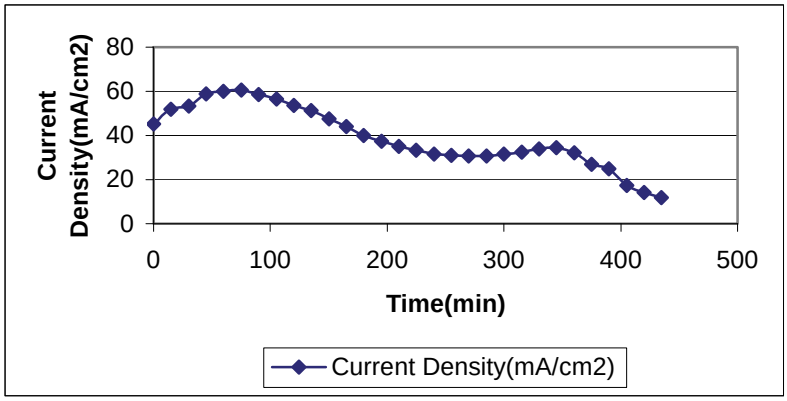


Figure 12.13 : Current density as a function of time.

12.7 Toxicity Tests

Toxicity of the strong leachate, pretreated leachate, ED product and ED brine was established by means of a 48-hour Daphnia Pulex lethality test. The LC_{50} (concentration at which 50% of the organisms die), LC_{10} (minimum effect concentration) and LC_0 (no effect concentration) values are given ion Table 12.6. Detailed results are shown in Appendix E.

Table 12.6 : LC_0, LC_{10} and LC_{50} values (%).

Sample	LC_0,	LC_{10}	LC_{50}	95% confidence limit	
				Lower	Upper
Strong leachate	0,31 (323 x)	0,92	1,72	1,50	2,03
Pretreated leachate	<0,63 (<159 x)	0,67	1,12	0,92	1,34
ED product	6,26 (16 x)	9,84	19,23	15,59	23,78
ED brine	0,125 (800 x)	0,33	0,61	0,50	0,75

( ) : dilutions

The LC_0 values in Table 12.6 indicate the dilution required to avoid acute effects on aquatic systems. For example, a 16 times dilution (100/6,26) will be required to avoid acute effects on aquatic life if the ED product is discharged into the water environment. An 800 times (100/0,125) dilution will be required in the case of the ED brine. Therefore, the ED product is significantly less toxic than the ED brine and the leachate and pretreated leachate.

13. PRELIMINARY ECONOMICS OF THE ED PROCESS

An estimation of the preliminary economics of the ED process derived from the laboratory-scale tests (Table 12.1) are as follows : -

The capital cost of an 80 kℓ/d ED plant is estimated at R5,6 million.
The capital cost of 140 kℓ/d ED plant is estimated at R8,3 million.
Operational costs are estimated at:

(a)	Electrical energy consumption (ion transport) (30,26 kWh @ R0,25/kWh)	R7,57/kℓ
(b)	Membrane replacement costs (1 year lifetime; 320 m² anionic @ R2 771/m²; 320 m² cationic @ R2 029/m²)	R52,6/kℓ
(c)	Chemical pretreatment costs ($NaCO_3$ - R20,40/kl; NaOH - R22,08/kℓ)	R42,48/kℓ
(d)	Pumping costs (3,02 kWh/kℓ @ R0,25/ℓ)	R0,76/kℓ
	TOTAL	R103,41/kℓ

NOTE: Membrane replacement costs for 2 and 3 years were determined at R26,3 and R17,5/kℓ, respectively. The cost data were derived from the batch laboratory-scale test, outlined in Table 12.1, at a high current density (approximately 90 mA/cm^2) chemical cleaning is not included in the operational costs.

14. EVALUATION OF ED ON PILOT-SCALE FOR THE TREATMENT OF THE INDUSTRIAL SOLID WASTE LEACHATE

14.1 Introduction

Pilot studies were conducted in an ED pilot plant (batch mode) on the Holfontein leachate to develop process design criteria for a full-scale application (see 3.11 for experimental procedure). Reverse osmosis was evaluated for the further desalination of the desalinated ED product. Feed-and-bleed ED tests were also conducted to develop process design criteria for a full-scale application. The preliminary economics of the process was derived from the experimental results.

14.2 Batch ED runs

Three batch ED runs were conducted on the pretreated leachate. The operational data of run 3 are shown in Table 14.1. The electrical current, electrical conductivity, cell pair resistance and current density as a function of time are shown in Figures 14.1 to 14.4.

The electrical current as a function of time is shown in Figure 14.1. The electrical current remained constant for some time before it started to decrease as more ions were removed from the feed compartments.

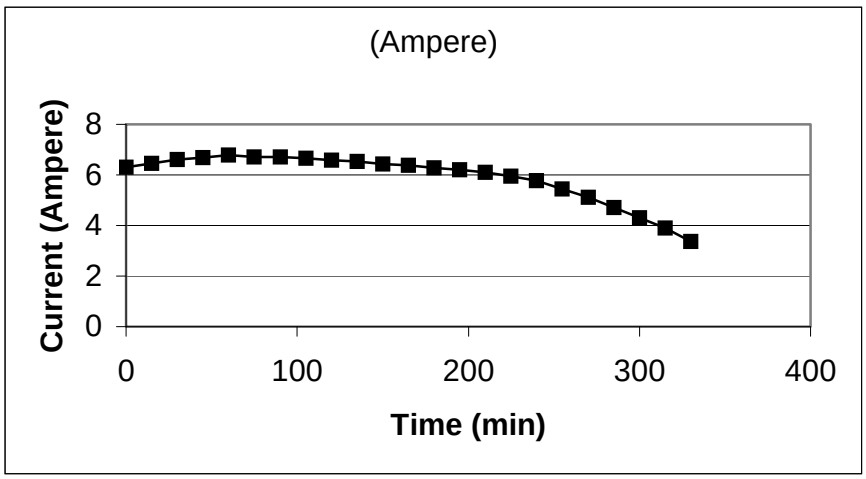


Figure 14.1 : Electrical current as a function of time.

Table 14.1 : Experimental conditions and results of a typical batch ED run conducted on the pretreated leachate (run 3, pretreated sample 14/01/2001, 200 g/ℓ Iscor ash, 12 g/ℓ soda ash, 6,9 NaOH).

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resist	Current density
(min)	(V)	(V)	(Ampere)	(mS/cm)	Deg C	pH	(mS/cm)	Deg C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm²)
0	26,06	17,03	6,31	92,7	26,1	7,54	96,3	26,5		85,4	8,28	4,13	2,70	30,93
15	26,04	16,33	6,46	87	28,9		109,3	29,3				4,03	2,53	31,67
30	26,09	16,37	6,62	82,6	30,7		117,1	31				3,94	2,47	32,45
45	26,06	16,38	6,68	78,7	32,1		123,7	32,3				3,90	2,45	32,75
60	2613	16,49	6,78	73,5	33,8		130,3	33,9				3,85	2,43	33,24
75	26,03	16,47	6,71	69,7	35		134,9	35,1				3,88	2,45	32,89
90	26	16,57	6,7	65,9	36		137,5	36,1				3,88	2,47	32,84
105	26,09	16,7	6,66	61,5	37,1		140,3	37,2				3,92	2,51	32,65
120	26,04	16,72	6,57	58	37,9		141,5	38				3,96	2,54	32,21
135	26	16,88	6,53	54,2	38,8		142,4	38,9				3,98	2,58	32,01
150	26,02	16,93	6,43	51	39,5		142,3	39,5				4,05	2,63	31,52
165	26,02	17	6,39	47,5	40,2		142,2	40,2				4,07	2,66	31,32
180	26	17,04	6,29	44,1	40,7		141,4	40,7				4,13	2,71	30,83
195	26,02	17,16	6,2	41,2	41,1		140,7	41,1				4,20	2,77	30,39
210	26,05	17,36	6,11	37,8	41,5		139,6	41,5				4,26	2,84	29,95
225	26,08	17,52	5,96	34,3	41,9		138,3	42				4,38	2,94	29,22
240	26,09	17,71	5,76	30,7	42,2		136,8	42,2				4,53	3,07	28,24
255	26,02	17,86	5,44	27,5	42,3		135,3	42,3				4,78	3,28	26,67
270	26,03	18,12	5,12	24,3	42,4		134,5	42,4				5,08	3,54	25,10
285	26,07	18,52	4,71	20,7	42,5		132,9	42,5				5,54	3,93	23,09
300	26	18,98	4,31	17,7	42,6		131,6	42,5				6,03	4,40	21,13
315	26,1	19,41	3,91	14,8	42,6		130	42,6				6,68	4,96	19,17
330	26,1	19,87	3,37	11,8	42,6		128,3	42,7				7,74	5,90	16,52
	26,05	17,37												

Feed start volume (ℓ)	40	Feed end volume	19 ℓ
Brine start volume (ℓ)	10	Brine end volume	30 ℓ

The electrical conductivity of the ED feed (demineralised feed) and brine as a function of time are shown in Figure 14.2. The ED feed was concentrated from approximately 93 mS/cm to approximately 128 mS/cm in the brine (at end of run), while the feed was desalinated to approximately 12 mS/cm (87% demineralisation) (Table 14.1). Therefore, a good percentage demineralisation of the leachate could be obtained.

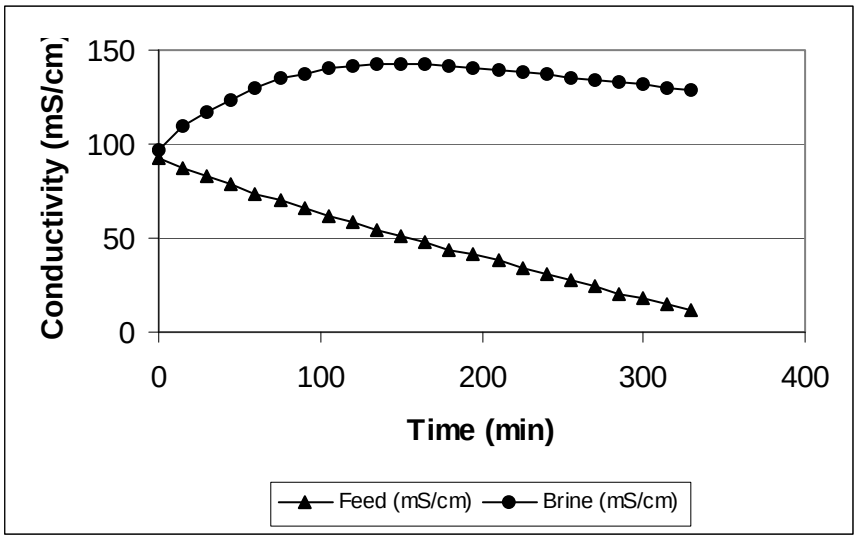


Figure 14.2 : Electrical conductivity as a function of time.

The cell pair resistance as a function of time is shown in Figure 14.3. The cell pair resistance remained more or less constant for a long time, whereafter it started to increase towards the end of the demineralisation cycle as a result of an increase in resistance in the diluate compartments.

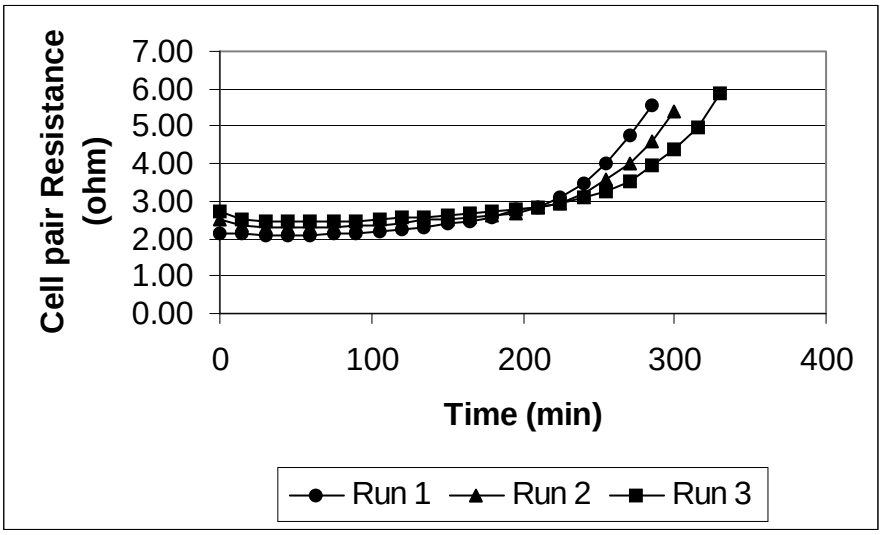


Figure 14.3 : Cell pair resistance as a function of time.

The current density as a function of time is shown in Figure 14.4. The current density also remained more or less constant during desalination and then started to decrease towards the end of the desalination cycle. (NOTE: The current density was significantly lower during the pilot tests than during the laboratory-scale ED tests (see Figure 12.4). This is due to a limitation in the ED pilot plant equipment.

The current efficiency before and after the ED runs are shown in Table 14.2.

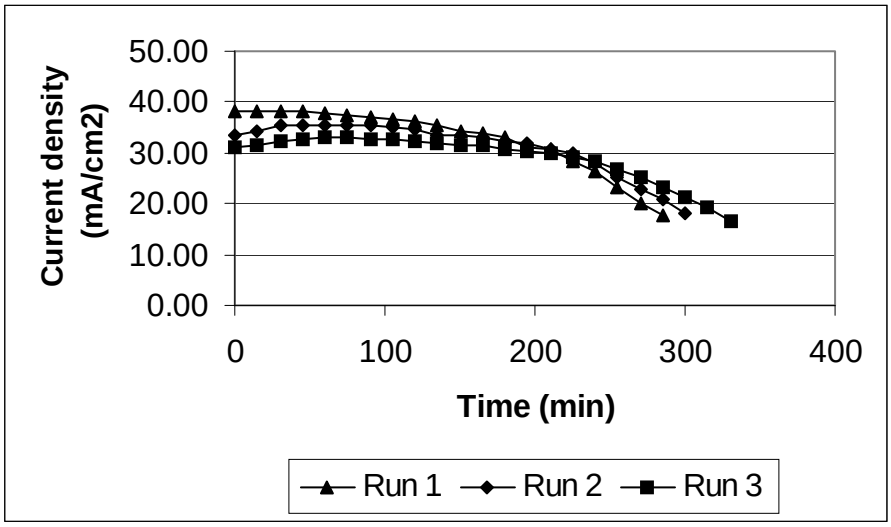


Figure 14.4 : Current density as a function of time.

Table 14.2 : Current efficiencies before and after the ED runs.

Run No.	Current efficiency %
Before (1)	91,1 (Na)
Before (2)	58,9 (Na)
Run 1	
Run 2	
Run 3	
After (3)	88,6 (Na)
After(2)	65 (Na)

(1) Feed 3 591 mg/ℓ NaCl, once through
(2) 90 mS/cm NaCl, batch desalination.
(3) Same as (1).

The current efficiency data showed that there was some reduction in current efficiency before (91,1%) and after the three batch runs (88,6%). This indicates some degree of membrane fouling. However, the reduction in current efficiency and accompanying membrane fouling appear not too serious.

14.2.1 Chemical composition of the ED feed, product and brine

The chemical composition of the ED feed, product and brine is shown in Table 14.3.

Table 14.3 : Chemical composition of the raw feed, treated feed, ED product and brine (run 3, pretreated with 200 g/ℓ Iscor ash, 6,9 caustic soda and 12 g/ℓ soda ash; pH adjusted to pH 7 with CO_2). Sample collected 14/10/2001.

Constituents*	Raw Feed	Treated feed	Product	Brine	Removal (%)
Conductivity (mS/cm)	90,7	92,7	11,8	128,3	87,27
PH	7,22	7,54	6,63	7,88	
Chemical oxygen demand (COD)	70 000	55 400	28 500	73 000	48,56
Total organic Carbon	8 687	8 290	8 967	12 542	-8,17
Alkalinity as $CaCO_3$	10 915	19 000	4 935	29 066	74,03
Total dissolved solids	100 945	104 260	11 903	169 805	88,58
Ammonia Nitrogen as N	1 983	840	105	11 314	87,50
Chloride	29 904	29 505	1 085	42 248	96,32
Sulphate	19 730	17 610	3 084	22 217	82,49
Calcium	1 010	25	13	40	48,00
Magnesium	896	29	2	52	93,10
Sodium	19 300	29 700	3 500	44 900	88,22
Potassium	4 240	4 540	393	7 120	91,34
Barium	0,46	0,1	0,05	0,19	50,00
Strontium	5,4	0,45	0,11	0,67	75,56
Phenolic compounds	516,9	38,7	41,8	35	-8,01

* Concentration in mg/ℓ , unless stated otherwise

Excellent TDS removal was obtained with batch ED treatment of the leachate. The TDS of the leachate was reduced from 104 260 to 11 903 mg/ℓ (88,6% removal). Excellent ammonia-nitrogen, chloride, sulphate, magnesium, sodium and potassium removals were also obtained.

COD removal was only 48,6%. This means that a significant quantity of the organics in the leachate can permeate the membranes. Excellent removal of phenolics were also obtained by pretreatment of the leachate (92,5% removal).

The current efficiency during the batch ED desalination (Table 14.3) was determined at 60,6% (cations) and 64,5% (anions). The electrical energy consumption was determined at 13,56 kWh/kℓ (R3,39/kℓ at R0,25/kWh) feed. Water recovery was 47,5% (19/40) and brine volume was 50% of the volume treated (20/40).

14.2.2. Chemical composition of the RO feed, product and brine

The ED product was further desalinated with RO using a tubular cellulose acetate membrane (batch mode). The results are shown in Figures 14.5 to 14.8. The RO product flux as a function of time is shown in Figure 14.5. The flux started at approximately 600 ℓ/m^2.d and decreased to approximately 200 ℓ/m^2.d at the end of the run. The initial CWF was approximately 1 000 ℓ/m^2.d and the CWF was reduced to approximately 750 ℓ/m^2.d at the

end of the run. Clean water flux, however, could be restored with chemical cleaning (1% STPP and 0,5% EDTA).

Permeate flux as a function of percentage water recovery is shown in Figure 14.6. Permeate flux decreased as a function of increasing percentage water recovery. Reverse osmosis was terminated at a water recovery of approximately 61% due to limitations (high dead volumes) in the experimental set-up.

The conductivity of the RO feed and product as a function of time is shown in Figure 14.7. The conductivity of the RO permeate started to increase towards the end of the run due to a higher feed (brine) concentration. The conductivity removal as a function of time is shown in Figure 14.8. Conductivity removal remained at approximately 95% for the entire run.

Excellent TDS removal was obtained (Table 14.4). TDS was reduced from 19 350 to 645 mg/ℓ (96,7% removal). COD was removed from 25 500 to 935 mg/ℓ (96,3% removal). Therefore, it appears that a significant quantity of the organics permeate the membranes. The RO permeate, except for COD, is of potable quality.

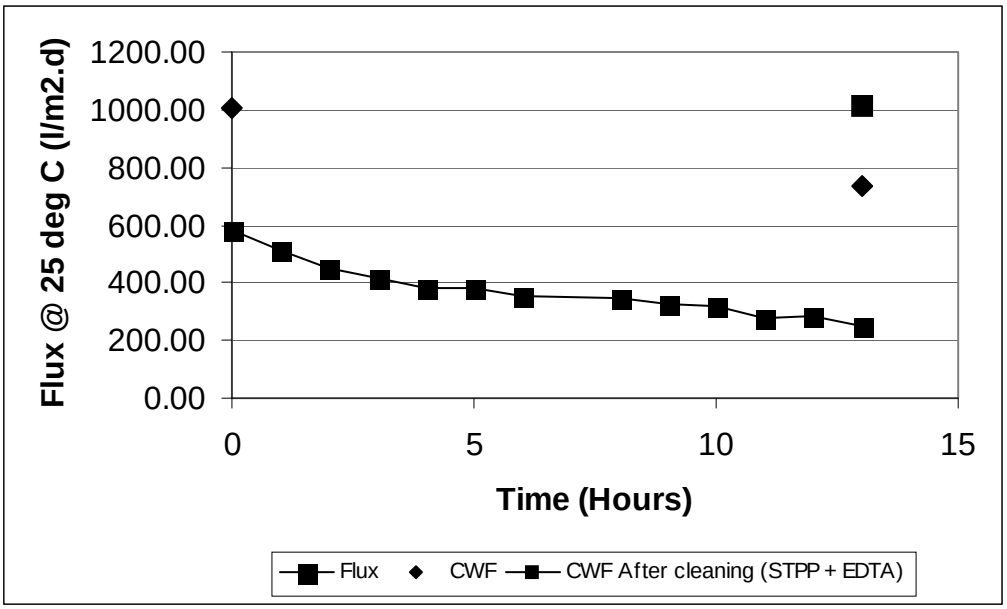


Figure 14.5 : RO product flux as a function of time.

Water recovery (Table 14.4) was only approximately 61%. However, a much higher water recovery (80%) should be possible with pretreatment of the feed for iron removal. Reverse osmosis should also be conducted with spiral membranes. The RO brine together with the ED brine (Tables 14.3 and 14.4) should be further desalinated in an evaporator to increase water recovery and to reduce effluent volume. The economics of such a possibility should be investigated.

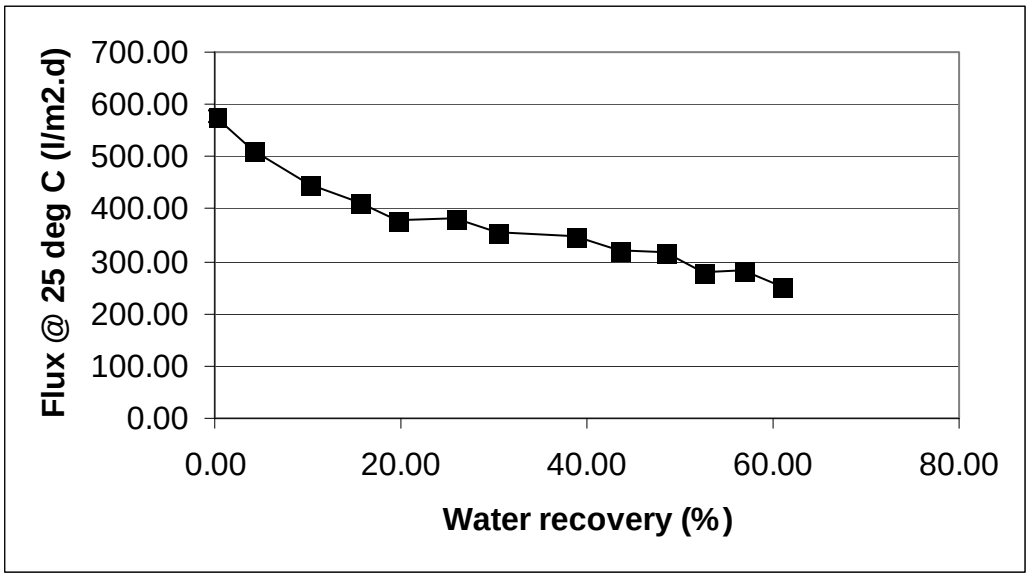


Figure 14.6 : Permeate flux as a function of water recovery.

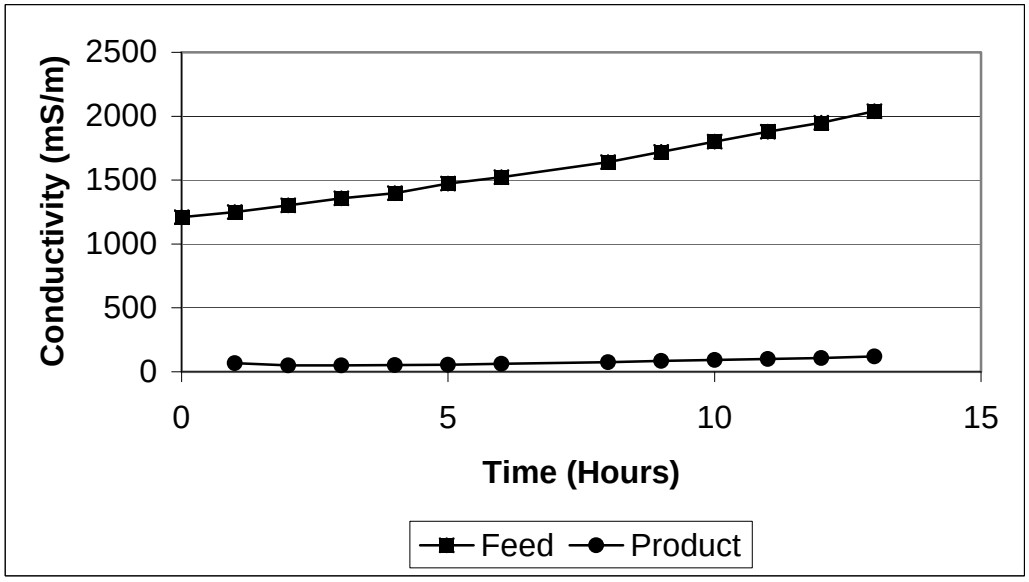


Figure 14.7 : Conductivity of RO feed and product as a function of time.

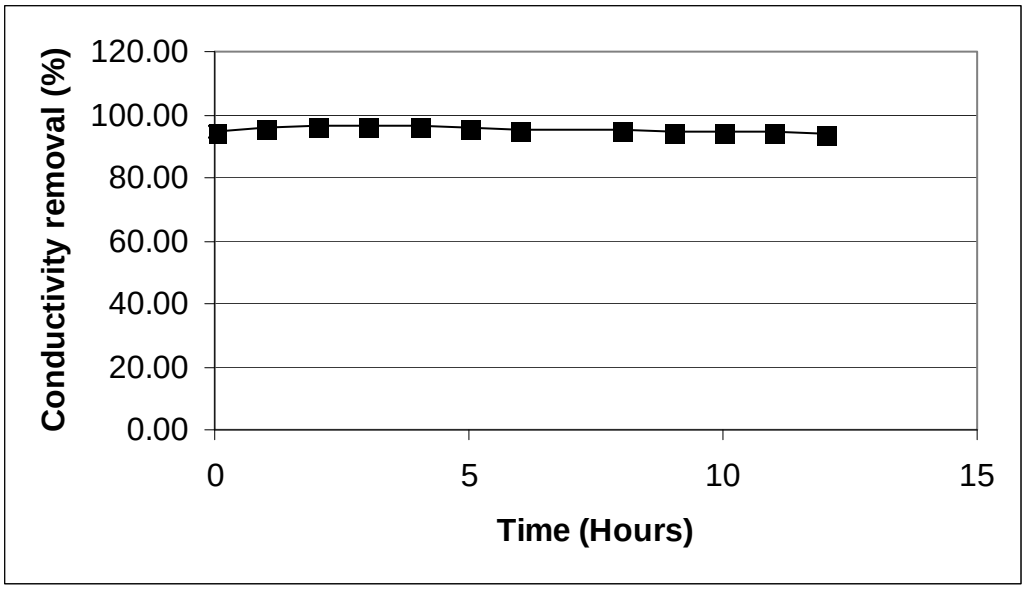


Figure 14.8 : Conductivity removal as a function of time.

Table 14.4 : Chemical composition of the RO feed, brine and product (desalination ED product as feed).

Constituents*	Feed	Product	Brine	Removal (%)
Conductivity (mS/cm)	12,1	0,75	20,4	93,80
pH	7,1	6,9	7,5	
Chemical oxygen demand (COD)	25 500	935	44 000	96,33
Total organic Carbon	4 240	162	11 733	96,18
Alkalinity as $CaCO_3$	4 918	240	8 900	95,12
Total dissolved solids	19 350	645	38 254	96,67
Ammonia Nitrogen as N	131,2	27,4	229	79,12
Chloride	1 014	59	2 254	94,18
Sulphate	2 247	10	5 240	99,55
Calcium	10	1	33	90,00
Magnesium	2	1	7	50,00
Sodium	3 590	162	7 280	95,49
Potassium	360	17	746	95,28
Barium	<0,03	<0,03	0,2	
Strontium	0,03	<0,03	0,13	
Iron	34,8	0,09	118	99,74
Silica	8,2	1,1	18,4	86,59
Phenolic compounds	60	49	110	18,33

* Concentration in mg/ℓ , unless stated otherwise

Feed	12 ℓ
Product	7,3 ℓ
Brine	4,7 ℓ

14.3 Feed-and-bleed ED tests

14.3.1 Electrodialysis at different leachate feed water concentrations

A summary of the electrodialysis results is shown in Table 14.5 and Figures 14.9 to 14.13. The detailed results are shown in Appendix F.

14.3.2 Electrical current

The electrical current was kept constant at 8 ampere for the first 3 runs (100% leachate, 80% of strength and 56% of strength). The voltage across the stack was about 33 to 29 volts, 34 to 30 volts, and 34 to 35 volts for the first 3 runs, respectively. The current for the last 2 runs was about 3 ampere (27% strength, voltage 25 to 26 volts) and 3 ampere (12% strength, voltage 45 to 44 volts) (Appendix F). The last two runs, however, were repeated in an attempt to obtain more accurate results.

14.3.3 Feed concentration

The initial feed concentration was 116 235 mg/ℓ for the first ED desalination stage (Table 14.5). The feed concentration was then diluted to 79 885, 51 600, 23 360 and 9 194 mg/ℓ for the subsequent stages of desalination. The initial concentrations for the repeat of stages 4 and 5 were 22 895 and 4 701 mg/ℓ, respectively.

Table 14.5 : Summary ED results.

Parameter	Stages					Repeat Stages	
	1	2	3	4	5	4 (Repeat)	5 (Repeat)
TDS Feed (mg/ℓ)	116 235	79 885	51 600	23 360	9 194	22 895	4 701
TDS Product (mg/ℓ)	92 000	58 485	26 310	10 360	2 435	6 765	1 990
TDS Brine (mg/ℓ)	150 845	138 835	140 145	86 125	80 470	111 360	61 590
TDS Loading (g/h.m^2)	703,95	565,85	591,05	286,26	146,11	348,74	59,25
Current efficiency $(\%)_{Cations}$	83,04	65,43	73,76	69,59	59,86	69,2	45,68
Current efficiency $(\%)_{Anions}$	71,70	70,12	67,00	66,00	58,31	66,51	31,50
Water transfer (g H_2O/g TDS)	2,71	2,99	3,22	3,7	4,35	4,25	10,72
Energy Consumption (kWh/kℓ)	4,38	4,81	5,61	1,81	2,37	4,7	0,72
Demineralisation $(\%)_{TDS}$	20,85	26,79	49,01	55,65	73,52	70,45	57,67
Brine/Product ratio	1,64	2,37	5,33	8,31	33,05	4,86	30,95
Total Voltage	30,80	32,30	35,30	26,30	44,50	50,06	30,19
Cell pair Voltage (75 cp)	20,30	22,30	25,50	19,70	36,70	38,94	22,71
Current	8,00	8,00	8,00	3,30	2,35	4,3	1,16

* Current efficiency 66% at end of runs (90 mS/cm).

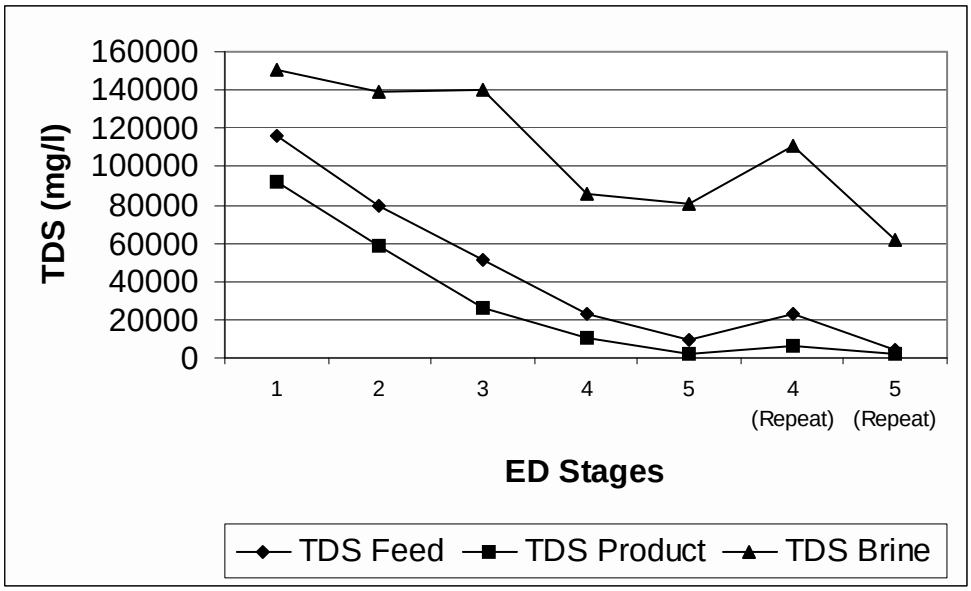


Figure 14.9 : Concentration of ED feed, demineralised feed and brine at different ED stages.

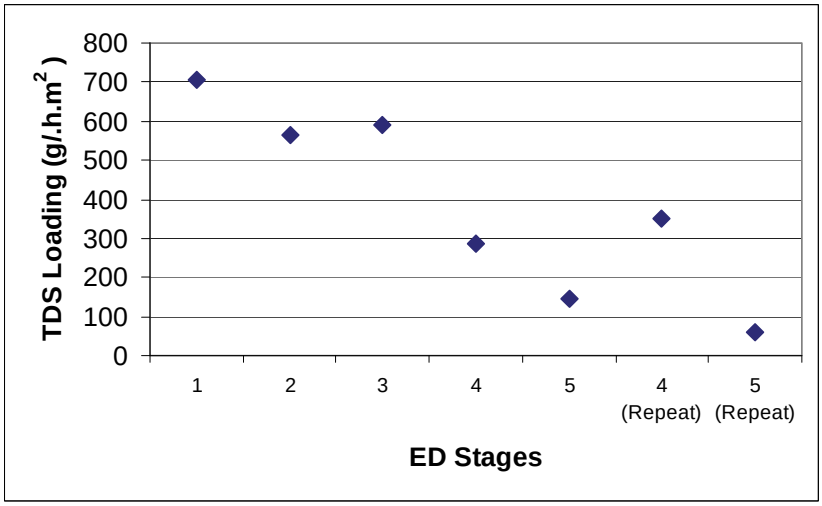


Figure 14.10 : TDS loading rate at the different ED stages.

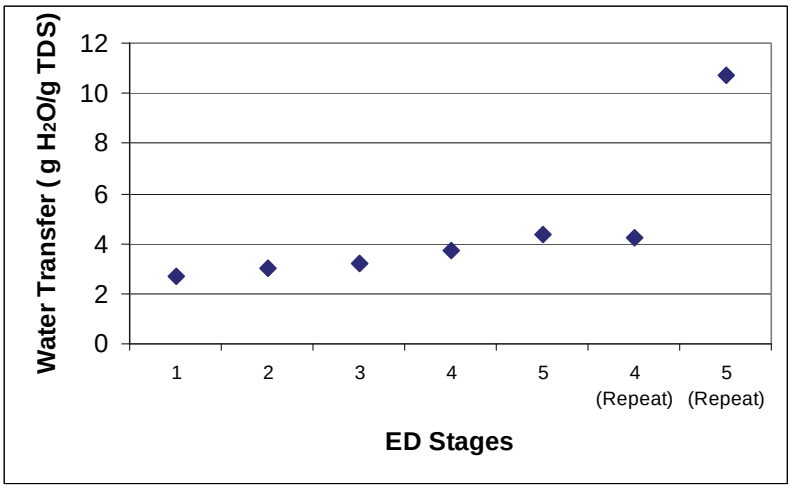


Figure 14.11 : Water transfer through the membranes at the different ED stages.

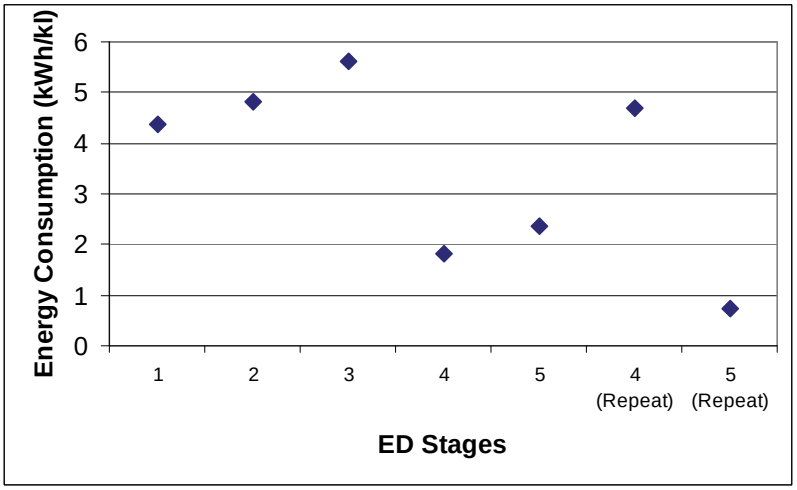


Figure 14.12 : Electrical energy consumption at the different ED stages.

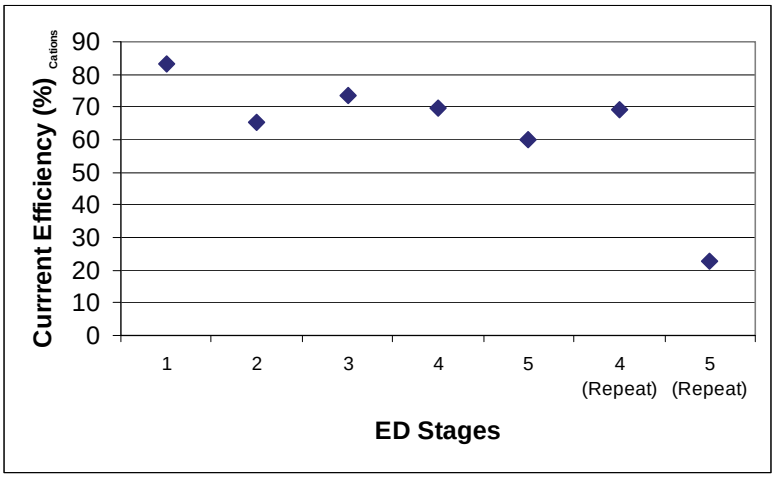


Figure 14.13 : Current efficiency at the different ED stages.

14.3.4 Demineralised feed concentration

The percentage demineralisation for the first ED stage was 20,9% (116 235 to 92 000 mg/ℓ, Table 5.14). The demineralisation percentages for the subsequent ED stages were 26,8% (2nd stage), 49,0% (3rd stage), 55,7% (4th stage) and 73,5% (5th stage). The percentage demineralisation for the repeat runs were 70,5% (4th stage) and 57,7% (5th stage). The concentrations of the ED feed, demineralised feed (product) and brine are shown in Figure 14.9. The TDS concentrations decreased as the number of stages increased.

14.3.5 Brine concentration

Brine concentration decreased with decreasing feed concentration. The brine concentration after the first stage was 150 845 mg/ℓ (1,3 times concentration). Brine concentration after the subsequent stages were 138 835 mg/ℓ (1,7 times concentration, 2nd stage), 140 145 mg/ℓ (2,7 times concentration, 3rd stage), 86 125 mg/ℓ (3,6 times concentration, 4th stage), and 80 470 mg/ℓ (8,7 times concentration, 5th stage). Brine concentration was 111 360 and 61 590 mg/ℓ for the repeat stages (4th and 5th stages), respectively.

The concentrations of the ED brine are shown in Figure 14.9.

14.3.6 TDS loading rate

The TDS loading rate is shown in Table 14.5 and Figure 14.10. The TDS loading rate decreased with decreasing feed concentration.

14.3.7 Water transfer

Water transfer through the membranes varied between 2,7 (1st stage) and 4,35 g H_2O/g TDS (5th stage) (Table 14.5 and Figure 14.11). Water transfer for the repeat 4th stage was 4,25 g H_2O/g TDS. However, water transfer for the repeat 5th stage was very high (10,72 g H_2O/g TDS).

14.3.8 Electrical energy consumption

The electrical energy consumption is shown in Table 14.5 and Figure 14.12. The electrical energy consumption was 4,38; 4,81; and 5,61 kWh/kℓ for the first three ED stages, respectively. However, the electrical energy consumption decreased significantly for the last two stages (1,81 and 2,37 kWh/kℓ) when the feed concentration was lower. The total electrical energy consumption was approximately 19 kWh/kℓ (R4,75/kℓ at R0,25/kWh). The electrical energy consumption for the repeat fourth and fifth stages were 4,7 (higher voltage) and 0,72 kWh/kℓ, respectively.

14.3.9 Current efficiency

The current efficiency for the different stages of ED desalination is shown in Table 14.5 and Figure 14.13. The initial current efficiency based on cations was high (83,0%, 1st stage) and then decreased as the concentration ratio between the demineralised feed and brine became larger. The current efficiency for the 5th stage was 59,8%. The current efficiency for the repeat fourth and fifth stages were 69,2 and 45,68%, respectively.

14.3.10 Feed and brine concentration flows through a 5-stage ED unit

The feed and brine concentration flows through a 5-stage ED unit is shown in Appendix F. The data in Appendix F (feed concentration and flow to the different ED stages and salt and water collected in the different ED stages resulting in the final brine flow and concentration) was used to determine the final ED product and brine concentration flows. The concentration of the desalinated leachate will have a TDS concentration of approximately 2,4 g/ℓ at a flow of 0,588 ℓ/min. The final brine flow exiting the ED stack will have a concentration of approximately 256,2 g/ℓ at a flow of 0,253 ℓ/min. Brine flow will comprise approximately 41,0% of the feed flow. Water recovery should be more than 90%.

14.3.11 Polarisation curve

The voltage-current characteristics of the membranes at a low feed concentrations where polarisation is likely to occur, are shown in Figures 14.14 to 14.16.

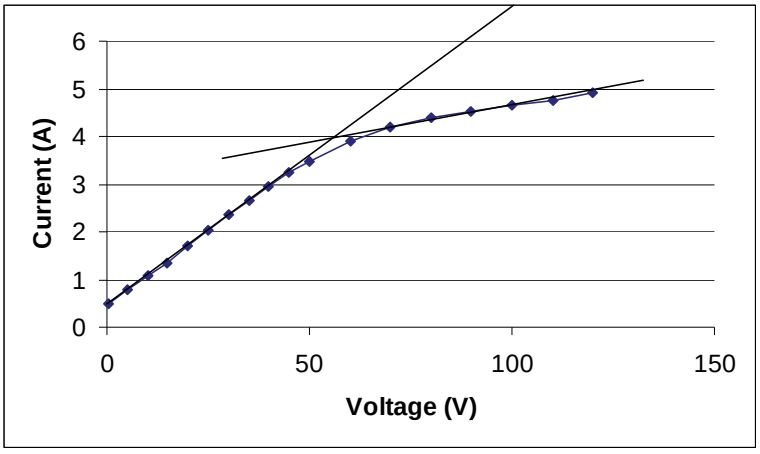


Figure 14.14 : Polarisation curve (10 mS/cm).

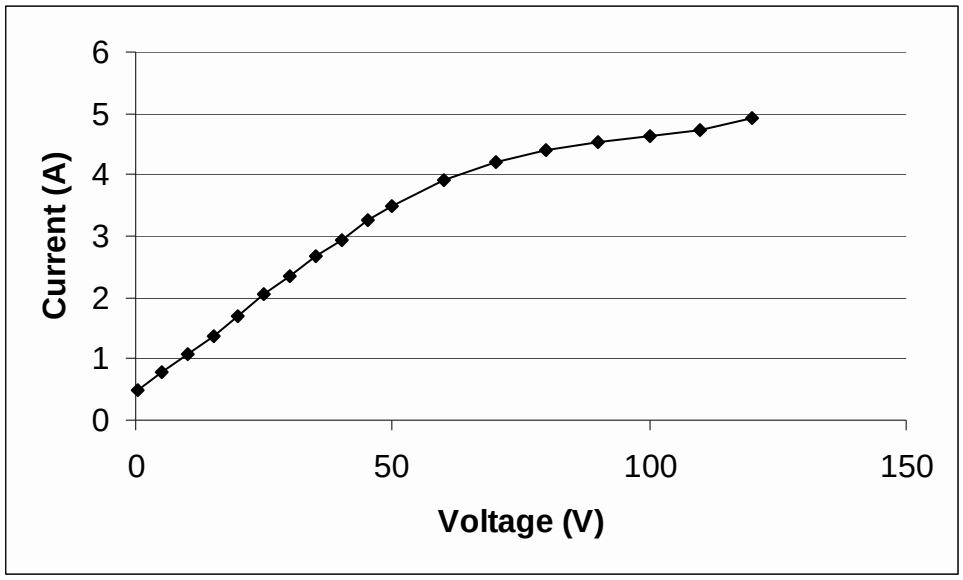


Figure 14.15 : Polarisation curve (5 mS/cm).

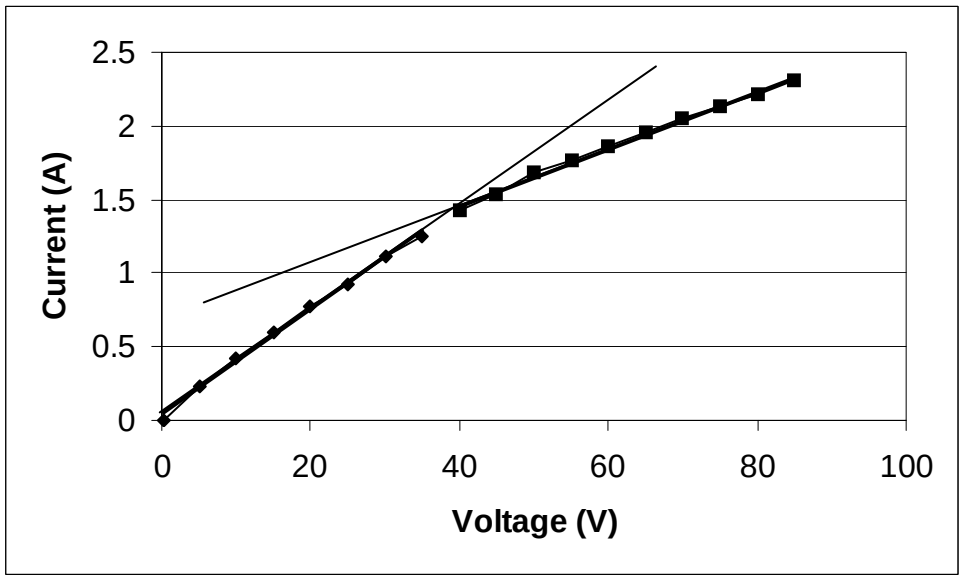


Figure 14.16 : Polarisation curve (2,5 mS/cm).

The voltage applied across an ED stack during ED treatment should not exceed 80 percent of the voltage at the inflection point. The approximate voltages at the inflection points were 53 volt (10 ms/cm, Figure 4.14), 53 volt (5 ms/cm, Figure 4.15) and 40 volt (2,5 ms/cm, Figure 4.16). Operation of an ED stack at 80% of the voltage at the inflection point (limiting voltage) will ensure that water splitting (polarisation) is eliminated during ED.

14.3.12 Estimated economics of the ED process from batch pilot test

An estimation of the economics of the ED process derived from pilot-scale tests are as follows : -

The capital cost of an 80 kℓ/d (feed) ED plant is estimated at R15,5 million. The capital cost of a 140 kℓ/d (feed) ED plant is estimated at R23,3 million.

Operational costs are estimated at:

(a)	Electrical energy consumption (ion transport) (13,56 kwh/kℓ @ R0,25/kwh)	R 3,39/kℓ
(b)	Membrane costs (1 year lifetime; 864 m^2 anion @ R2 771/m^2; 864 m^2 cation @ R2 029/m^2)	R142,03/kℓ
(c)	Chemical pretreatment costs (Na_2CO_3 - R20,40/kℓ; NaOH - R22,08/kℓ)	R42,48k/ℓ
(d)	Pumping costs (1,356 kWh @ R0,25/kwh)	R0,34/kℓ
	TOTAL	R188,24/Kℓ

Note : Membrane replacement costs for 2 and 3 years are R71,0 and R47,34/kℓ, respectively. The cost data was derived from the batch pilot-scale test outlined in Table 14.1 at a low current density (approximately 30 mA/cm^2).

The operation of the ED process at low current density results in a higher membrane area (higher costs) than at high current density (see 13). Operation at a too high current density might affect the membranes adversely. Therefore, the ED process should be operated at such a current density so as to optimise costs. This can only be established through long-term pilot studies.

The electrical energy consumption for ion transport was determined at 18,98 kwh/kℓ with the feed-and-bleed ED system (Table 14.5). This amounts to R4,75/kℓ for ion transport (R0,25/kWh). This is slightly more than the R3,39/kℓ shown for the batch system.

14.3.13 Sludge leachability

The inorganic chemical elements that can leach from the sludge (ash and chemical treatment of the Holfontein ISWL) are shown in Table 14.6 (see 3.11(b)). The organic compounds that can leach from the sludge are shown in Table 14.7

Table 14.6 : Chemical elements that can leach from the sludge

Determinant	Concentration (mg/ℓ)	
	Sample	Blank
Mg	306	0
K	305	0
Ba	0,481	0,352
Al	27,0	0,037
Cd	0	0
Cr	0,491	0,034
Co	0,115	0,006
Cu	0,009	0,008
Fe	15,4	0,043
Pb	0,301	0,060
Mn	20,5	0,015
Ni	0,212	0,013
Ag	0,010	0,014
Zn	0,867	0,221
V	0,168	0,016
Ti	7,14	0,007
Sb	1,00	0,108
As	0,263	0,002
Se	0,0224	0,0027
Hg	0	0,010
Zr	0,17	<0,10

Table 14.7 : Organic compounds that can leach from the sludge.

Constituent	Concentration (ppb)
Phenol	11,3
4-Methylphenol	13,2
Butyluted hydroxytoluene	101
Total aliphatic hydrocarbons (12 to 18)	98,8
2,6-bis(1,1-dimethylethyl-4-ethylphenol)	164

Significant quantities of magnesium, potassium, aluminium, iron, manganese and titanium can leach from the sludge. However, the concentration of these elements in a leachate is not considered to be dangerous. The concentrations of all the other elements that are leachable are less than 1 mg/ℓ .

The leachate sample was also analysed for : polynuclear aromatic hydrocarbons, chlorinated hydrocarbons, phtalate esters, polychlorobiphenyl hydrocarbons, nitrosamines, haloethers, aldehydes, ethers, ketones, anilines, pyridines, quinolines, aromatic nitro compounds, phenol compounds, including nitrophenols and chlorinated phenols and petroleum hydrocarbons.

The chromatogram obtained for the extract did not contain any traces of polynuclear aromatic hydrocarbons, chlorinated hydrocarbons, polychlorobiphenyl hydrocarbons, nitroso-amines, haloethers, aldehydes, ethers, ketones, anilines, pyridines, quinolines, aromatic nitro compounds, phenol compounds, nitro phenols, chlorinated phenols or petroleum hydrocarbon compounds (including BTEX).

There were traces of phenol and other organic compounds, typical of plastic decomposition products (see Table 14.7).

15. CHARACTERISATION OF THE MUNICIPAL SOLID WASTE LEACHATE

15.1 Introduction

The chemical, physical and biological characteristics of a number of MSW leachates were obtained from existing data. The characteristics for the Bisasar Road leachate were obtained through analysis in the laboratories of Environmentek as well as from existing data.

15.2 Characteristics of Municipal Solid Waste Leachates

The chemical composition of the Bisasar Road and Mariannhill leachates are shown in Tables 15.1 and 15.2, respectively. The chemical composition of other MSWL's is shown in Appendix G.

Table 15.1 : Typical characteristics of leachates from the Mariannhill and Bisasar Road landfill sites, and legislated discharge standards (existing data).

Parameter	Bisasar Rd Raw Leachate	Mariannhill Raw Leachate	"Special Standard" for Discharge	"General Standard" for Discharge	Discharge to Durban Metro Sewer
pH value	8,0	7,6	5,5 to 7,5	5,5 to 9,5	6,0 to 10
Alkalinity ($CaCO_3$)	6 440 (a)	5 343	nl (b)	nl	nl
Conductivity	1 291	1 285	250	250	nl
COD	2 427	1 878	30	75	nl
BOD_5	320	695	nl	nl	nl
BOD_5/COD	0,13	0,37	-	-	-
NH_4-N	1 271	442	1,0	10,0	nl
NO_2-N	0,1	<0,1	nl	nl	nl
NO_3-N	2,6	2,7	1,5	nl	nl
Chloride	1 790	1 496	nl	nl	1 000
PO_4-P	8,4	5,0	1,0	1,0	nl
SO_4	48	96	nl	nl	250
Na	897	914	50 > influent	90 > influent	nl
Mg	56	355	nl	nl	nl
K	1 022	1 232	nl	nl	nl
Ca	36	208	nl	nl	nl
Cr	0,05	0,08	0,05	0,5	20
Mn	0,12	0,86	0,1	0,4	50
Fe	2,70	18,6	0,3	nl	50
Ni	0,09	0,12	nl	nl	50
Cu	<0,01	<0.01	0,02	1,0	50
Zn	0,08	0,17	0,3	5,0	50
Cd	<0,01	<0.01	0,05	0,05	20
Pb	<0,02	0,09	0,1	0,1	20

(a) All results are in mg/ℓ except for pH value (dimensionless) and electrical conductivity(mS/m).
(b) nl - indicates either not legislated or that no record of legislation limitation could be found.
(c) The sewage discharge data relates to effluents flowing to a large works@ of >25 Mℓ/d capacity.

The electrical conductivities of the Bisasar Road leachate were determined as 1 291 (Table 15.1) and 1 650 mS/m (Table 15.2) in two cases. The TDS were determined at 7 070 mg/ℓ (Table 15.2). The conductivity of the leachate does not comply to the requirements of the general standard for the discharge to the water environment.

The COD of the Bisasar Road leachate were 2 427 mg/ℓ (Table 15.1) and 2 000 mg/ℓ (Table 15.2). The COD values also do not comply to the requirements of the general standard. The BOD/COD ratio was only 0,13 in the one case (Tale 15.1) and 0,48 in the other case (Table 15.2). The first case (low BOD/COD) shows that it will be difficult to biodegrade the leachate, while the second case (high BOD/COD) shows that the leachate should be readily biodegradable.

The ammonia-nitrogen concentrations of the Bisasar Road leachates are high. It was 1 271 mg/ℓ in the one case (Table 15.1) and 980 mg/ℓ in the other case (Table 15.2). These concentrations do not comply to the requirements of the general standard.

The sodium, chloride and potassium concentrations of the leachate are also high. The chloride concentrations (1 790 mg/ℓ, Table 15.1 and 2 625 mg/ℓ, Table 15.2) are significantly higher than the general standard for discharge. The sodium concentrations are also most probably too high for discharge to the water environment.

The heavy metals concentrations (Cr, 0,17 mg/ℓ; Pb, 0,126 mg/ℓ; Ni, 0,2 mg/ℓ) were higher in the one case (Table 15.2) than in the other case (Table 15.1) (Cr, 0,05 mg/ℓ; Pb <0,02 mg/ℓ; Ni 0,09 mg/ℓ, Table 15.1). The lead concentration does not comply to the general standards for discharge. The barium (0,495 mg/ℓ), strontium (1,09 mg/ℓ), iron (3,16 mg/ℓ) and manganese (0,382 mg/ℓ) concentrations (Table 15.2) are also high. These high concentrations can affect the performance of a membrane process adversely.

The chemical composition of the Bisasar Road leachate show that the leachate should be treated prior to discharge into the water environment. The leachate, however, will comply to the quality requirements for discharge to the Durban Metro Sewer if the chloride concentration could be reduced to less than 1 000 mg/ℓ .

The Mariannhill leachate is a weaker leachate than the Bisasar Road leachate in terms of especially ammonia-nitrogen (Table 15.1). This leachate, however, also does not comply to the quality requirements for discharge to the water environment or the sewer and should be treated prior to discharge.

The characteristics of other MWS leachates are shown in Appendix G. The salinity, COD and ammonium-nitrogen concentrations of these leachates are too high for discharge to the water environment. Some of the heavy metal concentrations are also too high. The chloride concentrations of most of the leachates are too high for discharge to sewer (>1 000 mg/ℓ Cl). Therefore, most of the MSW leachates should be treated prior to discharge to either the water environment or to the sewer system.

Table 15.2 : Chemical composition of the Bisasar Road leachate (date 15/03/2001) (Environmentek laboratories).

Constituents*	Bisasar Road raw leachate
Conductivity (mS/m)	1 650
pH	8,00
Chemical oxygen demand (COD)	2 000
Biological oxygen demand (BOD)	955
BOD/COD ratio	0,48
Total organic Carbon	112,5
Oxygen Adsorbed (OA)	0
Alkalinity as $CaCO_3$	5 552
Total suspended solids	368
Total dissolved solids	7 070
Ammonia Nitrogen as N	980
Nitrate as N	4,95
Nitrite as N	0,09
Chloride	2 625
Sulphate	149
Total Phosphate	6,91
Silicon as Si	23,6
Calcium	70,6
Magnesium	141
Sodium	1 620
Potassium	1150
Barium	0,495
Strontium	1,09
Chromium (Total)	0,17
Copper	0,008
Iron	3,16
Lead	0,126
Manganese	0,382
Nickel	0,20
Zinc	0,025
Cadmium	0,004
Phenolic compounds	9,31

* Concentration in mg/ℓ, unless stated otherwise

16. EVALUATION OF THE FOULING POTENTIAL OF THE MUNICIPAL SOLID WASTE LEACHATE FOR TUBULAR CELLULOSE ACETATE AND POLYAMIDE MEMBRANES AND EVALUATION OF MEMBRANE CLEANING STRATEGIES OF FOULED RO MEMBRANES WITH MEMBRANE CLEANING AGENTS

16.1 Introduction

Batch RO tests were conducted using tubular cellulose acetate and polyamide membranes to evaluate the fouling potential of the MSWL for the RO membranes and to evaluate membrane cleaning strategies with commercially available membrane cleaning agents. Leachate were collected at the Bisasar Road Landfill Site in Durban and transported to Pretoria for subsequent tests. Two batch runs were conducted on 12 litre feed sample for each membrane according to the procedures described in Section 3 (see 3.12)

16.2 Tubular Cellulose Acetate Membranes

The experimental conditions and results are shown in Table 16.1 (1st run) and Figures 16.1 to 16.4 and Table 16.2 (2nd run) and Figures 16.5 to 16.8.

Similar permeate fluxes (approximately 500 to 400 ℓ/m^2.d) as a function of time and percentage water recovery were obtained for the two runs (Figures 16.1 and 16.5 and 16.2 and 16.6). The CWF's after runs 1 (Figures 16.1) and 2 (Figure 16.5) were significantly lower than the initial CWF's. This indicates that the leachate will foul the membranes. It is further interesting to note that the CWF after run 1 was significantly higher than after run 2. This indicated that the membrane surface was more fouled after run 2 than after run 1. However, it appears that it should be possible to restore the CWF's after cleaning of the membranes with acid and STPP and EDTA membrane cleaning solutions (Figures 16.1 and 16.5). It is also interesting to note that the preservation of the membranes with formaldehyde solution had the effect to increase the CWF (Figure 16.5).

The conductivity of the RO feed and product as a function of time for the first two runs are shown in Figures 16.3 and 16.7, respectively. The feed water concentration was about the same in both cases (1 700 to 2 600 mS/m). A better quality RO product water, however, was produced during the 1st run (Figure 16.3) than during the second run (Figure 16.7).

Table 16.1 : Summary of experimental conditions and results using tubular cellulose acetate membranes for evaluating the fouling potential of the MSWL for the membranes (1st run) (run 1, CA membrane, Bisasar leachate, pH 6,5) H_2SO_4 - 1,815 g/ℓ).

Time	Pressure in	out	Tempera-ture	Flux	Flux normalized	Feed	Product	Volume recovered		Recovery	Rejection
	(kPa)	(kPa)	(deg C)	(mℓ/min)	(ℓ/m2.d)	(mS/m)	(mS/m)	(l)		(%)	(%)
0	4 000	3 800	20,5	54	3 314,48	CWF	-	12		-	
0	4 000	3 800	18,5	17	1 090,34	1 733	-	0	0	0,00	
1	4 000	3 800	34	13	555,86	1 690	494	0,95	0,95	7,92	70,77
2	4 000	3 800	36	12,5	500,00	1 774	135	0,7	1,65	13,75	92,39
3	4 000	3 800	30	11,5	555,17	1 856	351	0,7	2,35	19,58	81,09
4	4 000	3 800	34,5	11,5	483,79	1 920	159	0,65	3	25,00	91,72
5	4 000	3 800	35,5	11,5	467,93	1 980	153	0,63	3,63	30,25	92,27
6	4 000	3 800	33	11	485,52	2 060	167	0,58	4,21	35,08	91,89
7	4 000	3 800	34	10,5	448,97	2 160	176	0,57	4,78	39,83	91,85
8	4 000	3 800	34,5	11	462,76	2 250	184	0,54	5,32	44,33	91,82
9	4 000	3 800	45	13,5	372,41	2 320	193	0,62	5,94	49,50	91,68
10	4 000	3 800	35	10,5	434,48	2 480	213	0,57	6,51	54,25	91,41
11	4 000	3 800	35,5	10	406,90	2 600	221	0,57	7,08	59,00	91,50
11.5	4 000	3 800	35,5	9,5	386,55	2 720	226	0,25	7,33	61,08	91,69
11.5	4 000	3 800	20,5	35,5	2 178,97	CWF					
11.5	4 000	3 800	20,5	40	2 455,17	CWF After cleaning with 1% Sodiumtripolyphosphate and 0,5% EDTA (pH 10,57) NaOH					
11.5	4 000	3 800	20,5	45	2 762,07	CWF After cleaning with 1% Sodiumtripolyphosphate and 0.5% EDTA (pH 10,78) NaOH					
11.5	4 000	3 800	20,5	51	3 130,34	CWF After cleaning with 2% Sodiumtripolyphosphate and 0.5% EDTA (pH 10,86) NaOH					

The percentage conductivity removal was less after run 2 (approximately 85%, Figure 16,8) than after run 1 (approximately 93%, Figure 16.4). The conductivity of the RO product was also slightly better after run 1 (Figure 16.3) than after run 2 (Figure 16.7) as indicated in the previous paragraph. This can be ascribed to membrane fouling.

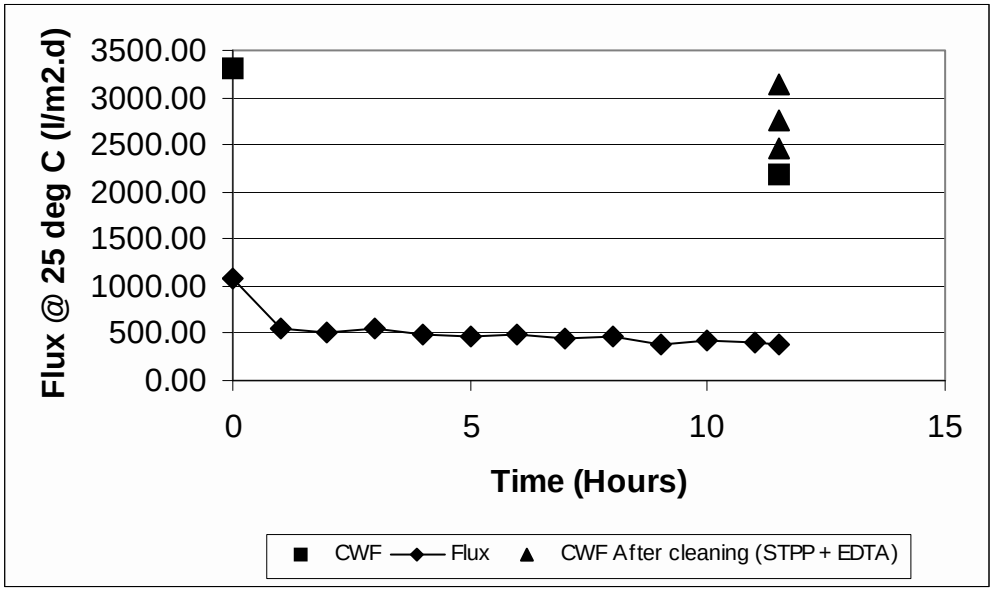


Figure 16.1 : Permeate flux as a function of time.

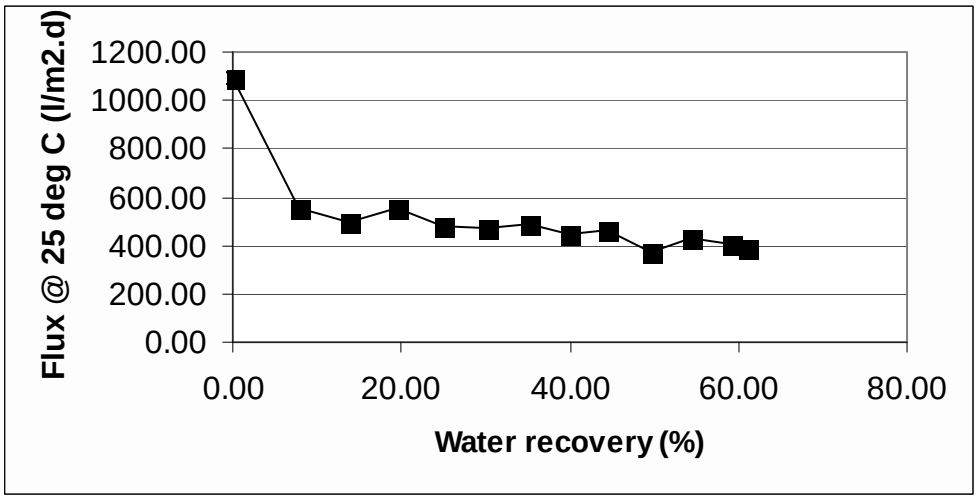


Figure 16.2: Permeate flux as a function of % water recovery.

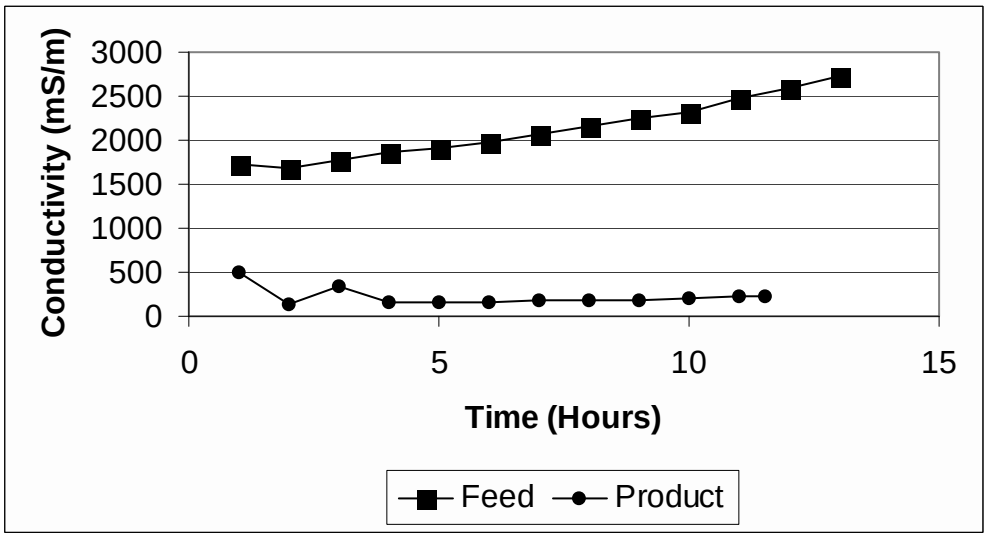


Figure 16.3 : Conductivity of the RO feed and product as a function of time.

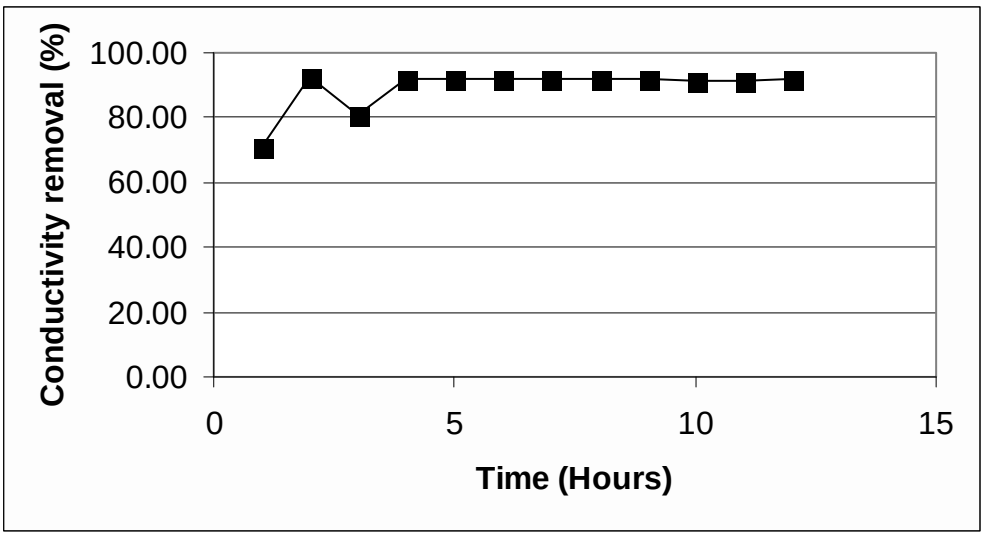


Figure 16.4 : Conductivity removal as a function of time.

Table 16.2 : Summary of experimental results using tubular cellulose acetate membranes for evaluating the fouling potential of the MSWL for the membranes (2nd run) (run 2, CA membrane, Bisasar leachate, pH 6,5) (H_2SO_4 - 1,815 g/ℓ)).

Time	Pressure in	out	Temperature	Flux	Flux normalized	Feed	Product	Volume recovered		Recovery	Rejection
	(kPa)	(kPa)	(°C)	(mℓ/min)	(ℓ/m2.d)	(mS/m)	(mS/m)	(ℓ)		(%)	(%)
0	4 000	3 800	20,5	51	3 130,34	CWF	-	12		-	
0	4 000	3 800	20	13	806,90	1 704	-	0	0	0,00	
1	4 000	3 800	32	12	546,21	1 695	303	0,62	0,62	5,17	82,12
2,5	4 000	3 800	35	12	496,55	1 779	220	1	1,62	13,50	87,63
3	4 000	3 800	35	12	496,55	1 808	219	0,33	1,95	16,25	87,89
4	4 000	3 800	36	12	480,00	1 878	226	0,68	2,63	21,92	87,97
5	4 000	3 800	36	12,5	500,00	1 948	236	0,65	3,28	27,33	87,89
6,5	4 000	3 800	35	11,5	475,86	2 000	259	0,92	4,2	35,00	87,05
7	4 000	3 800	35	11	455,17	2 060	276	0,34	4,54	37,83	86,60
8	4 000	3 800	35,5	11,5	467,93	2 130	276	0,58	5,12	42,67	87,04
9	4 000	3 800	36,5	11,5	452,07	2 230	285	0,64	5,76	48,00	87,22
10,5	4 000	3 800	33,5	10	434,48	2 380	476	0,8	6,56	54,67	80,00
11,5	4 000	3 800	33	9,5	419,31	2 490	354	0,6	7,16	59,67	85,78
12	4 000	3 800	33	10	441,38	2 530	345	0,3	7,46	62,17	86,36
12	4 000	3 800	21	18	1 092,41	CWF					
12	4 000	3 800	19,5	28,5	1 788,62	CWF after preservation in 4 g/ℓ formaldehyde for 3 days.					
12	4 000	3 800	19,5	38,5	2 416,21	CWF After cleaning with 0,3% Nitric acid (pH 1,92)					
12	4 000	3 800	19,5	46	2 886,90	CWF After cleaning with 2% Sodiumtripolyphosphate and 0,5% EDTA (pH 10,76) NaOH					
12	4 000	3 800	19,5	53	3 326,21	CWF After cleaning with 2% Sodiumtripolyphosphate and 0,5% EDTA (pH 10,84) NaOH					

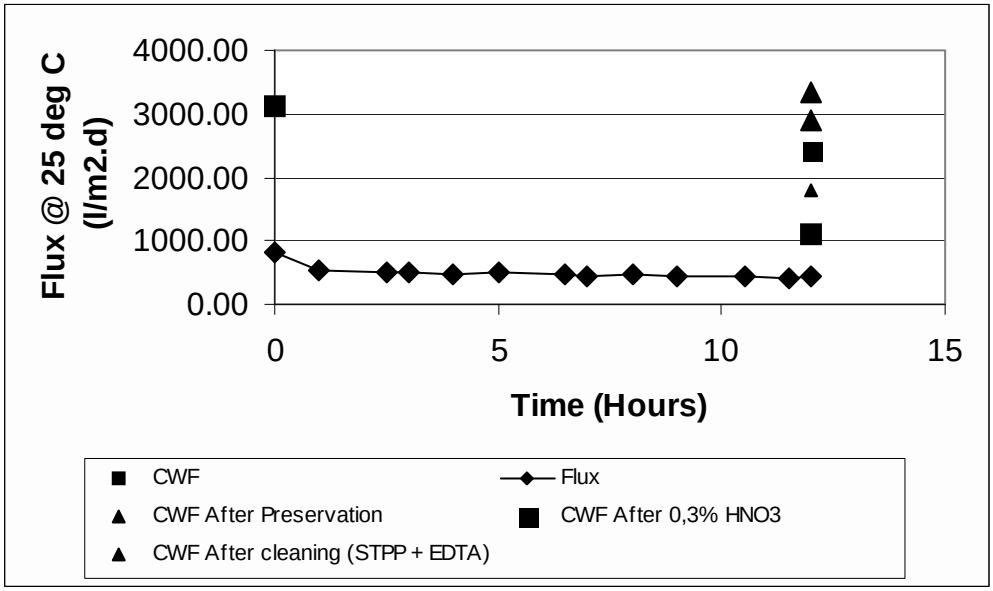


Figure 16.5 : Permeate flux as a function of time.

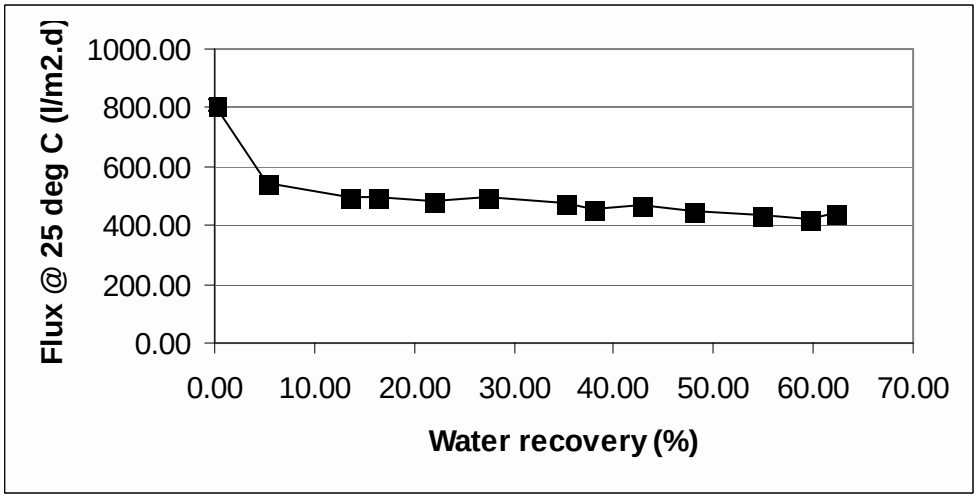


Figure 16.6 : Permeate flux as a function of % water recovery.

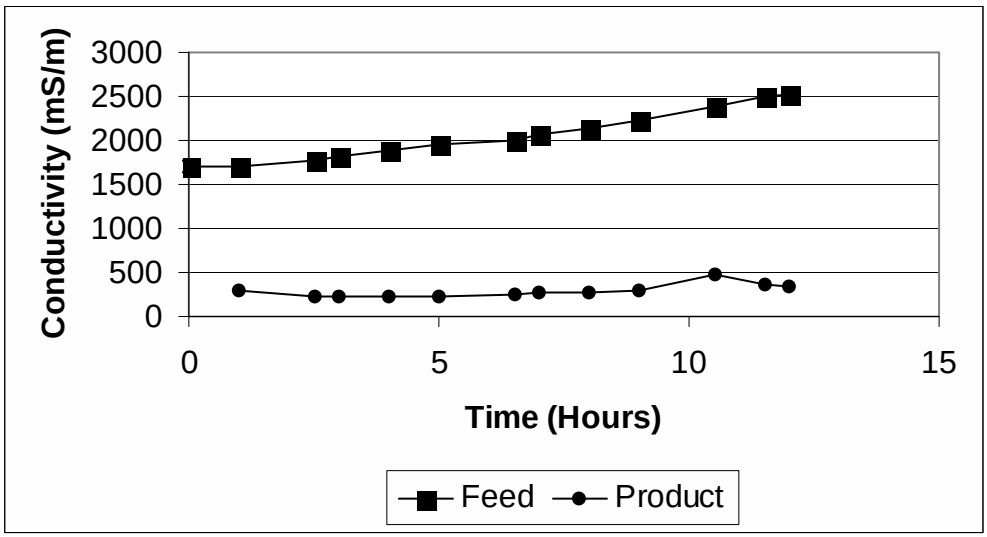


Figure 16.7 : Conductivity of the RO feed and product as a function of time.

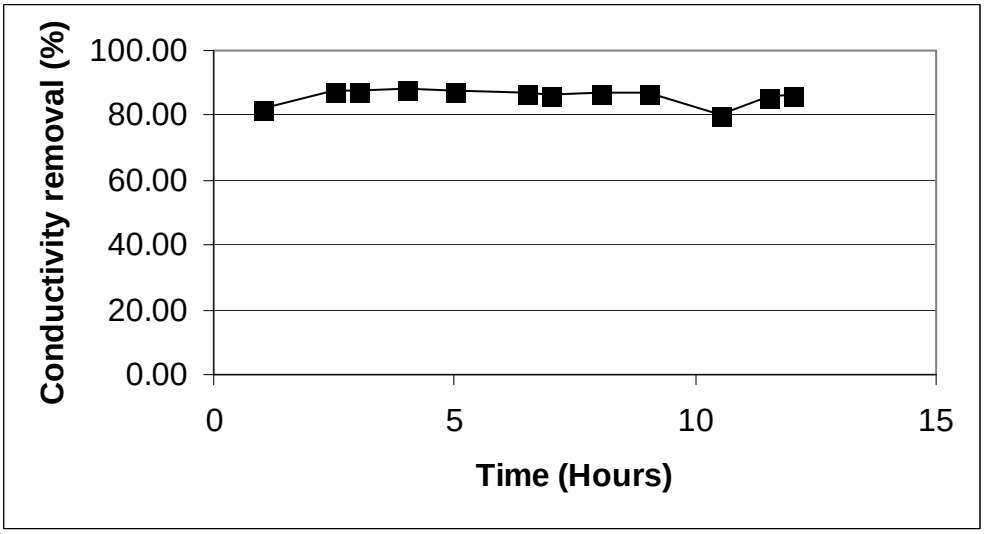


Figure 16.8 : Conductivity removal as a function of time.

16.3 Tubular Polyamide Membranes

The experimental conditions and results are shown in Table 16.3 (run 1) and Figures 16.9 to 16.12 and Table 16.4 (run 2) and Figures 16.13 to 16.16.

A somewhat higher permeate flux (800 to 600 ℓ/m^2.d) as a function of time and water recovery was obtained for the second run (Figures 16.13 and 16.14) than was obtained for the first run (600 to 500 ℓ/m^2.d) (Figures 16.9 and 16.10). The permeate fluxes obtained with the polyamide membranes were higher than the fluxes obtained with the cellulose acetate membranes. It is further interesting to note that the CWFs at the end of the two runs were approximately the same as the initial CWF's (Figure 16.9 and 16.13). Cleaning of the

membranes with SLS and EDTA solution had very little effect on the CWF because the membranes were nearly completely clean.

Conductivity removal as a function of time varied between approximately 96 and 97 percent for the two runs (Figures 16.12 and 16.16). This is higher than obtained with the cellulose acetate membranes. The conductivity of the RO permeate was also significantly better than the conductivities obtained with the cellulose acetate membranes (Figures 16.11 and 16.15). Therefore, it appears that the polyamide membranes should give a better performance for the treatment of the leachate than the cellulose acetate membranes.

Table 16.3 : Summary of experimental conditions and results using tubular polyamide membranes for evaluating the fouling potential of the MSWL for the membranes (1st run): (run 1, AFC 99 membrane, Bisasar leachate, as is).

Time	Pressure in	out	Tempera-ture	Flux	Flux normalized	Feed	Product	Volume recovered		Recovery	Rejection
	(kPa)	(kPa)	(°C)	(mℓ/min)	(ℓ/m2.d)	(mS/m)	(mS/m)	(ℓ)		(%)	(%)
0	4 000	3 800	18,5	22	1 411,03	CWF	-	12		-	
0	4 000	3 800	20	17	1 055,17	1 620	-	0	0	0,00	
1	4 000	3 800	29,5	13	636,55	1 686	49	0,91	0,91	7,58	97,09
2,5	4 000	3 800	31	12,5	586,21	1 817	57	1,4	2,31	19,25	96,86
3	4 000	3 800	30,5	11,5	547,24	1 927	59	0,74	3,05	25,42	96,94
4	4 000	3 800	29	11,5	571,03	2 020	68	0,86	3,91	32,58	96,63
5	4 000	3 800	29	11,5	571,03	2 160	62	0,88	4,79	39,92	97,13
6	4 000	3 800	29,5	11	538,62	2 310	67	0,91	5,7	47,50	97,10
7	4 000	3 800	29	10,5	521,38	2 470	72	0,7	6,4	53,33	97,09
8	4 000	3 800	32	11	500,69	2 680	80	0,93	7,33	61,08	97,01
8	4 000	3 800	18,5	19	1 218,62	CWF					
8	4 000	3 800	18,5	20	1 282,76	CWF after cleaning with 0,2% SLS, 0,1% EDTA, pH 11,73 (NaOH)					

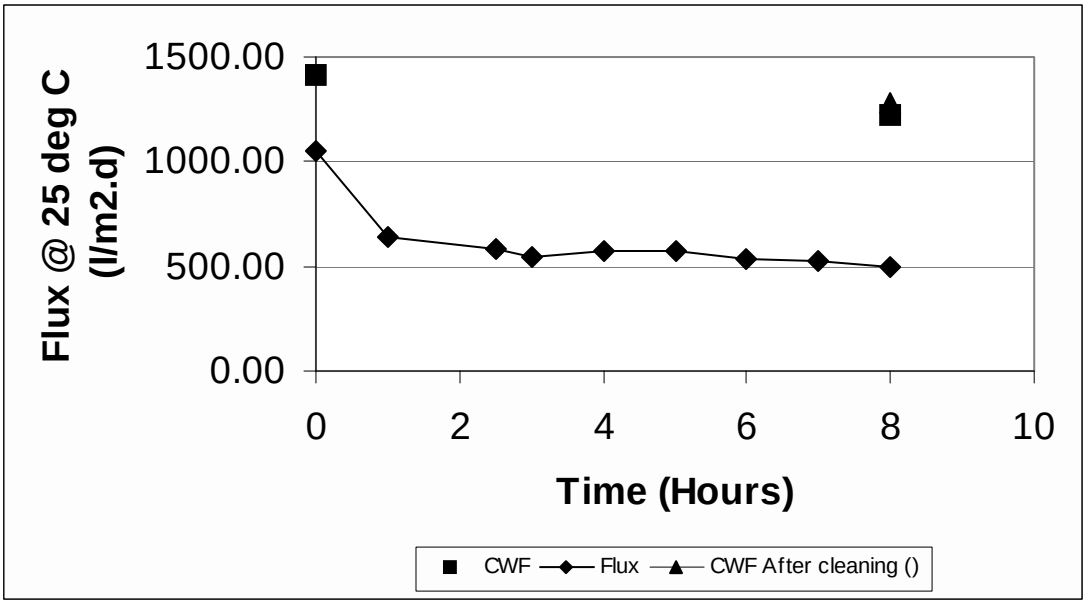


Figure 16.9 : Permeate flux as a function of time.

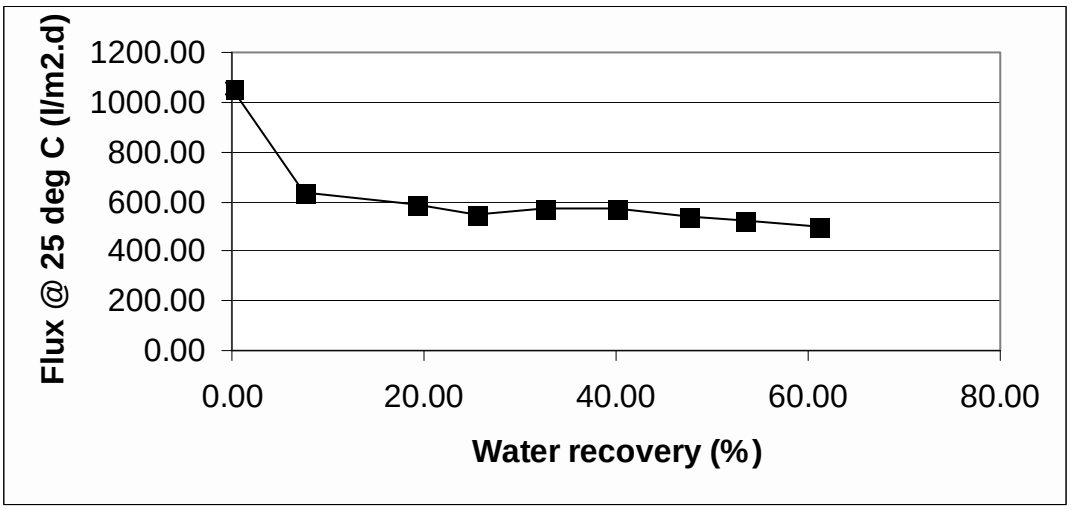


Figure 16.10 : Permeate flux as a function of % water recovery.

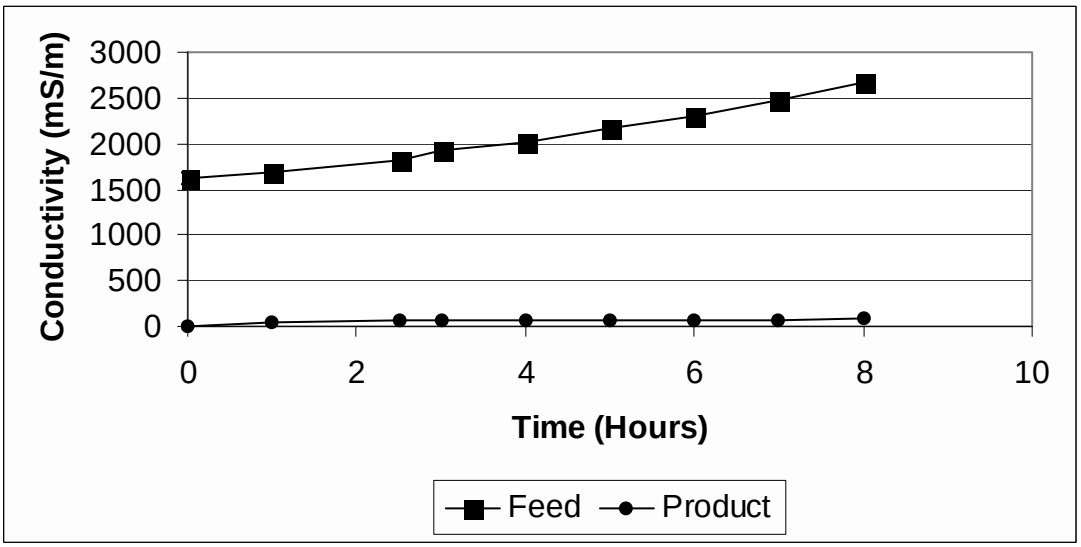


Figure 16.11 : Conductivity of the RO feed and product as a function of time.

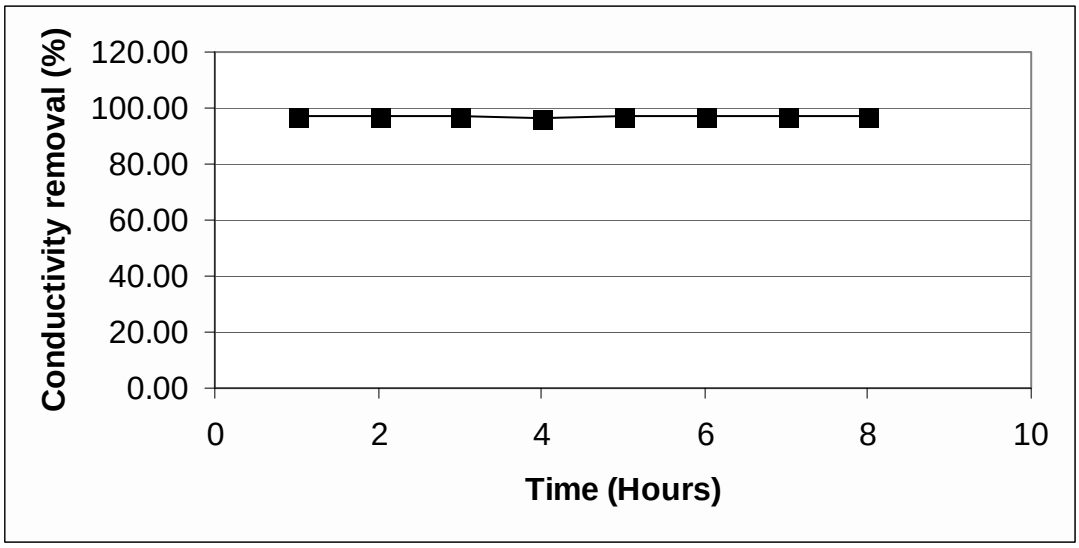


Figure 16.12 : Conductivity removal as a function of time.

Table 16.4 : Summary of experimental conditions and results using tubular polyamide membranes for evaluating the fouling potential of the MSWL for the membranes (2nd run) (run 2, AFC 99 membrane, Bisasar leachate as is) 14/05/2001).

Time	Pressure in	out	Temperature	Flux	Flux normalized	Feed	Product	Volume recovered		Recovery	Rejection
	(kPa)	(kPa)	(°C)	(mℓ/min)	(ℓ/m2.d)	(mS/m)	(mS/m)	(ℓ)		(%)	(%)
0	4 000	3 800	18,5	20	1 282,76	CWF	-	12		-	
0	4 000	3 800	18	11,5	745,52	1 587	-	0	0	0,00	
1	4 000	3 800	32,5	18	806,90	1 625	41	0,9	0,9	7,50	97,48
2	4 000	3 800	31,5	17,5	808,62	1 709	52	0,8	1,7	14,17	96,96
3	4 000	3 800	31,5	16	739,31	1 813	54	0,94	2,64	22,00	97,02
4	4 000	3 800	32,5	16	717,24	1 916	58	0,86	3,5	29,17	96,97
5	4 000	3 800	32,5	16	717,24	2 020	64	0,96	4,46	37,17	96,83
6	4 000	3 800	33	15,5	684,14	2 160	71	0,78	5,24	43,67	96,71
7	4 000	3 800	30,5	14	666,21	2 320	70	0,82	6,06	50,50	96,98
8	4 000	3 800	31	13	609,66	2 468	77	0,72	6,78	56,50	96,88
9	4 000	3 800	31	13	609,66	2 630	85	0,63	7,41	61,75	96,77
9	4 000	3 800	15	18,5	1 275,86	CWF					
9	4 000	3 800	16	18,5	1 250,34	CWF after cleaning with 0,2% SLS, 0,1% EDTA, pH 11,73 (NaOH)					

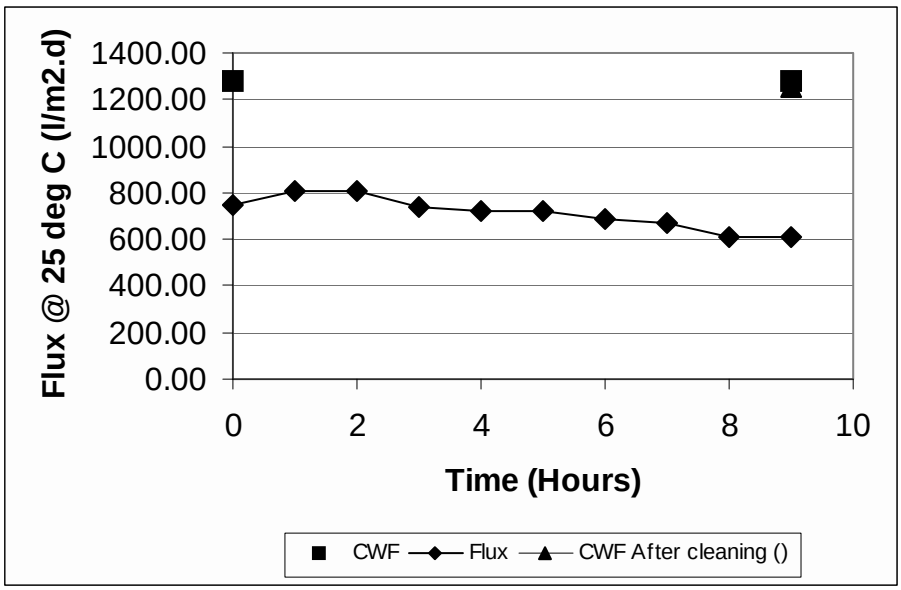


Figure 16.13 : Permeate flux as a function of time.

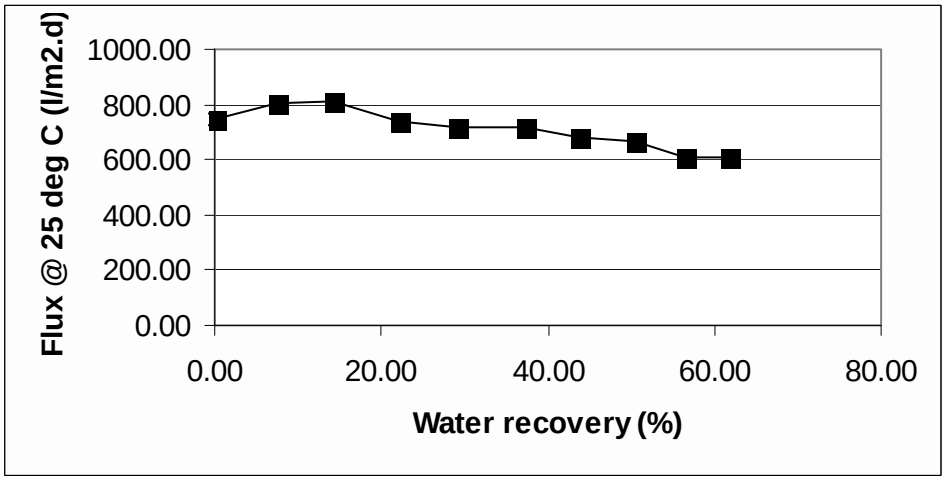


Figure 16.14 : Permeate flux as a function of % water recovery.

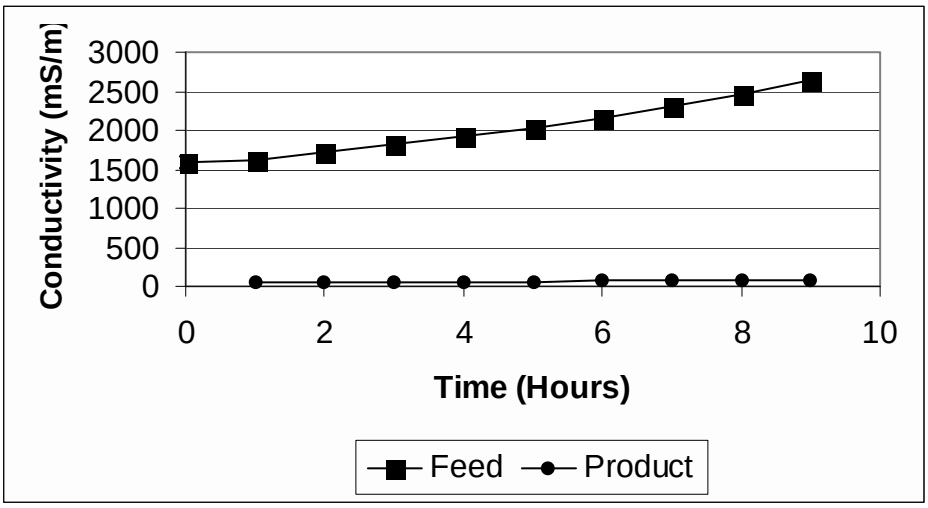


Figure 16.15 : Conductivity of the RO feed and product as a function of time.

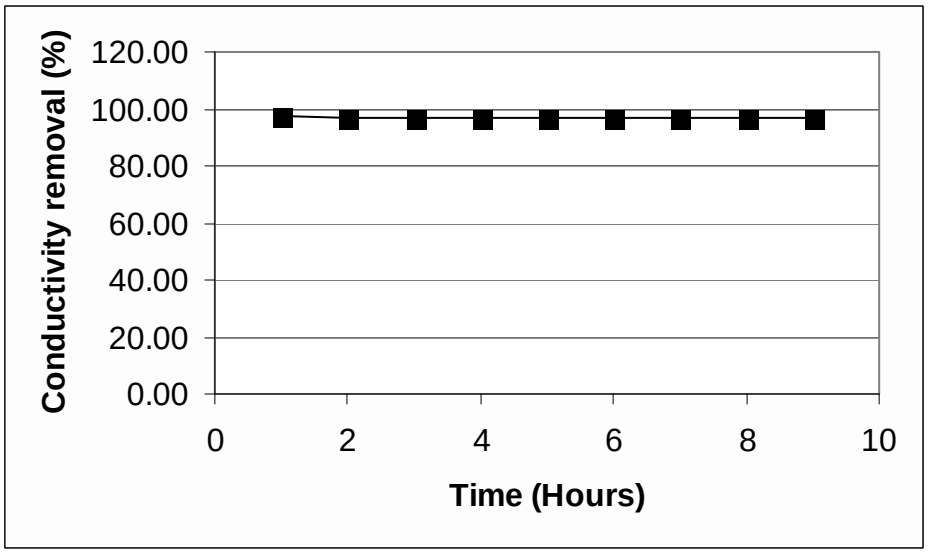


Figure 16.16 : Conductivity removal as a function of time.

17. EVALUATION OF TUBULAR REVERSE OSMOSIS FOR THE DESALINATION/ CONCENTRATION OF THE MUNICIPAL SOLID WASTE LEACHATE

17.1 Introduction

Batch reverse osmosis tests were conducted in an RO pilot plant using cellulose acetate and polyamide membranes (see 3.13) and the chemical composition of the RO feed, product and brine was determined. Permeate fluxes were determined as a function of percentage water recovery. The desalination/concentration performance of the cellulose acetate and polyamide membranes were compared. Preliminary process design criteria and the economics of the process for a full-scale application were derived from the experimental data.

17.2 Tubular Cellulose Acetate Membranes

The experimental conditions and results using cellulose acetate membranes for the batch treatment of the leachate are shown in Tables 17.1 (1^{st} run) and 17.2 (2^{nd} run) and Figures 17.1 (flux data) and Table 17.3 (chemical analysis).

Table 17.1: Summary of experimental conditions and results using cellulose acetate membranes for the desalination / concentration of the leachate (run 1, cellulose acetate membranes (1,75 m^2), pH adjusted to pH 6,5 with sulphuric acid).

Time	Flux	Flux @ 25 °C	Temperature	Feed		Product		Recovery		Pressure (kPa)		Rejection
(min)	(mℓ/30 s)	(ℓ/d.m2)	(°C)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ℓ)	(%)	(in)	(out)	(%)
0	440	887	16	28,2	(CWF)	2,7		100	0	4 000	4 000	90,4
0	270	615	9,6	17,59	6,58			0	0,0	4 000	4 000	100,00
15	270	592	11,7	17,87	6,84	1,23	3,05	10	10,0			93,1
30	290	614	13,5	18,42	7	0,75	5,12	17	17,0			95,9
45	295	595	16	20	7,3	0,83	5,46	26	26,0			95,9
60	315	600	18,7	22	7,6	1,01	6,05	35	35,0			95,4
75	320	579	21	24,7	7,81	1,25	6,58	45	45,0			94,9
90	300	543	21	27,7	7,98	1,62	7,68	53	53,0			94,2
105	285	511	21,4	31,9	8,08	1,9	8,74	62	62,0			94,0
120	260	456	22,4	38,1	8,14	2,34	8,95	68	68,0			93,9
135	205	350	23,5	45,1	8,16	3,01	9,06	75	75,0			93,3
135	415	833	16,2		(CWF)	1,46	6,69		75			91,7

Table 17.2 : Summary of experimental conditions and results using cellulose acetate membranes for the desalination concentration of the leachate (run 2, cellulose acetate membranes (1,75 m^2), pH adjusted to pH 6,5 with sulphuric acid).

Time	Flux	Flux @ 25 °C	Temperature	Feed		Product		Recovery		Pressure (kPa)		Rejection
(min)	(ml/30 s)	(ℓ/d.m2)	(°C)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ℓ)	(%)	(in)	(out)	(%)
0	415	833	16,2		(CWF)			100		4 000	4 000	
0	290	638	11,5	17,43	6,58			0	0,0	4 000	4 000	100,0
15	295	627	13,3	17,98	6,78	0,66	5,23	10	10,0			96,3
30	305	617	15,8	18,82	7	0,71	5,38	18	18,0			96,2
45	305	594	17,7	20,3	7,21	0,8	5,66	27	27,0			96,1
60	315	592	19,3	22,4	7,45	0,95	6,01	36	36,0			95,8
75	300	548	20,6	25	7,64	1,15	6,37	45	45,0			95,4
90	300	536	21,6	28,8	7,77	1,46	6,79	53	53,0			94,9
105	300	522	22,7	32,7	7,88	1,86	7,49	61	61,0			94,3
125	245	410	24,3	41,2	7,95	2,48	8,5	70	70,0			94,0
125	415	840	15,8		(CWF)	1,2	6,36					93,1

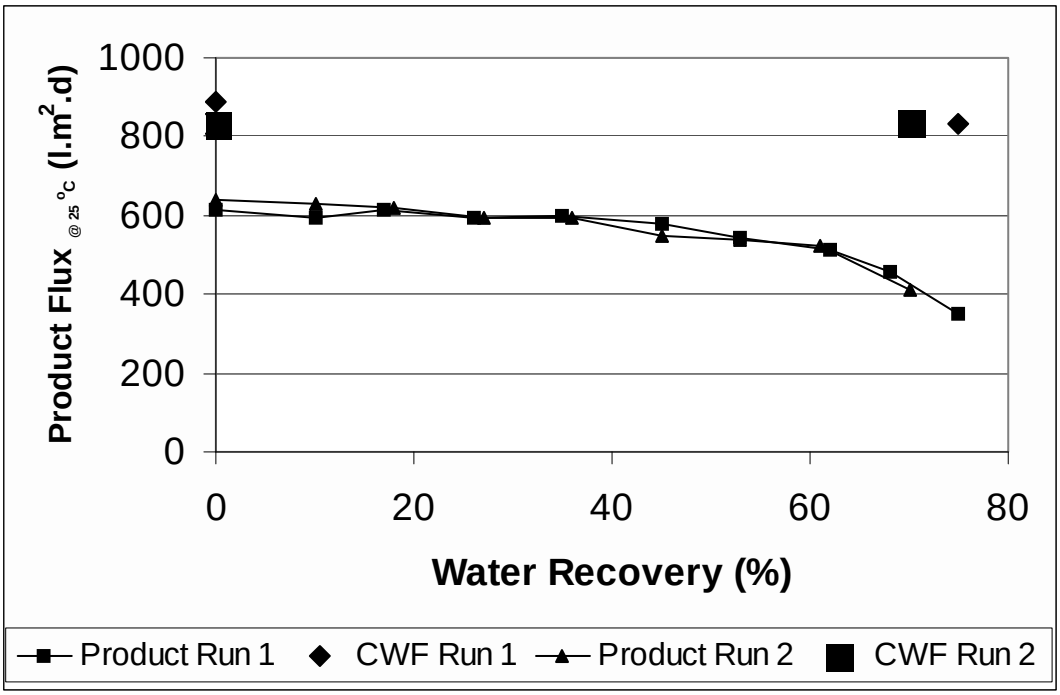


Figure 17.1 : Permeate flux as a function of % water recovery.

The initial permeate flux was approximately 600 ℓ/m^2.d and the flux decreased as a function of percentage water recovery as a result of the increased osmotic pressure of the feed at higher water recoveries (>40%) (Figure 17.1). It is interesting to note that almost identical results were obtained with the two runs that were conducted (up to approximately 70% water recovery). It is also interesting to note that the initial and CWFs at the end of the runs were almost identical. This shows that it should be possible to control membrane fouling with flow reversal and sponge ball cleaning.

The chemical composition of the RO permeate indicates that the TDS of the leachate could be reduced from 8 975 to 348 mg/ℓ (96,1% removal). Therefore, an excellent quality water could be produced with RO treatment of the leachate. Ammonia-nitrogen, however, was only reduced from 882 to 82 mg/ℓ (90,7% removal). Therefore, ammonia-nitrogen removal was not that good. The removal, however, of the other ions like chloride (92,4%), sulphate (99,5%), calcium (98,8%), magnesium (99,7%), potassium (94,9%), sodium (96,4%) and COD (97,7%) were excellent.

Table 17. 3: Chemical composition of the RO feed, product and brine (cellulose acetate membranes).

		Feed	Product	Brine	Removal	Feed	Product	Brine
Constituents					(%)	(g)	(g)	(g)
TDS		8 975	348	24695	96,1	897,500	24,360	740,850
TSS		29	6	48	79,3	2,900	0,420	1,440
Ammonia-N		882	82	1 770	90,7	88,200	5,740	53,100
Nitrate-N		0,21	0,1	0,7	52,4	0,021	0,007	0,021
Total P		9,19	0,063	13,6	99,3	0,919	0,004	0,408
Alkalinity as $CaCO_3$		2 854	165	7 758	94,2	285,400	11,550	232,740
Chloride		2 495	189	5 804	92,4	249,500	13,230	174,120
Sulphate		2 454	12,07	5 773	99,5	245,400	0,845	173,190
Fluoride		0,847	0,008	2,61	99,1	0,085	0,001	0,078
Silicon		22,8	1,86	60,9	91,8	2,280	0,130	1,827
Calcium		80	0,922	234	98,8	8,000	0,065	7,020
Magnesium		146	0,483	431	99,7	14,600	0,034	12,930
Potassium		821	41,5	2 410	94,9	82,100	2,905	72,300
Sodium		1 510	53,9	4 210	96,4	151,000	3,773	126,300
Barium		0,501	0	1,67	100,0	0,050	0,000	0,050
Strontium		1,27	0,007	3,78	99,4	0,127	0,000	0,113
Chromium (Total)		0,072	0	0,334	100,0	0,007	0,000	0,010
Iron		3,6	0	10	100,0	0,360	0,000	0,300
Lead		0,126	0,008	0,447	93,7	0,013	0,001	0,013
Manganese		0,308	0	0,843	100,0	0,031	0,000	0,025
Nickel		0,418	0,016	1,12	96,2	0,042	0,001	0,034
TOC		423	20,56	1 410	95,1	42,300	1,439	42,300
COD		2 200	51	6 150	97,7	220,000	3,570	184,500
Phenols		0,34	0,11	2,84	67,6	0,034	0,008	0,085
Conductivity (mS/m)		1759	120	4120	93,2			
pH		6,58	6,36	7,95				

Feed	100Litre
Product	70Litre
Brine	30Litre

17.3 Tubular Polyamide Membranes

The experimental conditions and results using the polyamide membranes for the batch treatment of the leachate are shown in Tables 17.4 (1st run) and 17.5 (2nd run) and Figure 17.2 (flux data) and Table 17.6 (chemical analysis).

Table 17.4 : Summary of experimental conditions and results using polyamide membranes for the desalination/concentration of the leachate (run 1, AFC99 PA membranes (0,81 m^2), no pH adjustment).

Time	Flux	Flux @ 25 °C	Temperature	Feed		Product		Recovery		Pressure (kPa)		Rejection
(min)	(mℓ/30 s)	(ℓ/d.m2)	(°C)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ℓ)	(%)	(in)	(out)	(%)
0	460	2 036	15,2	28,9	(CWF)	1,8		100	0	4 000	4 000	93,8
0	250	1 227	9,8	17,19	8.13			0	0,0	4 000	4 000	100,0
15	265	1 237	12,5	17,36	8.14	0,25	7,32	8	8,0			98,6
30	275	1 225	14,9	18,71	8.15	0,33	8,15	16	16,0			98,2
45	280	1 195	17	20,1	8.15	0,39	8,75	25	25,0			98,1
60	280	1 145	19	22,1	8.15	0,46	8,99	34	34,0			97,9
75	285	1 137	20,1	24,4	8.16	0,51	9,16	41	41,0			97,9
90	265	1 027	21,4	27,4	8.16	0,61	9,19	49	49,0			97,8
105	245	923	22,6	31,9	8.16	0,77	9,24	56	56,0			97,6
120	220	814	23,4	36,5	8.16	0,95	9,24	64	64,0			97,4
135	180	653	24,2	42	8.16	1,18	9,25	68	68,0			97,2
140	175	632	24,4	43,4	8.16	1,27	9,24	70	70,0			97,1
140	430	1 888	15,6		(CWF)	0,56	9,1		70			96,74

Table 17.5 : Summary of experimental conditions and results using polyamide membranes for the desalination / concentration of the leachate (run 2, AFC99, PA membranes (0,81 m^2), no pH adjustment).

Time	Flux	Flux @ 25 °C	Temperature	Feed		Product		Recovery		Pressure (kPa)		Rejection
(min)	(mℓ/30 s)	(ℓ/d.m2)	(°C)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ℓ)	(%)	(in)	(out)	(%)
0	430	1 888	15,6		(CWF)			100		4 000	4 000	
0	225	1 052	12,4	17,19	8,19			0	0,0	4 000	4 000	100,0
15	250	1 129	14,2	17,32	8,19	0,29	8,6	7,5	7,5			98,3
30	250	1 087	16,1	18,26	8,19	0,31	8,89	15	15,0			98,3
45	255	1 072	17,7	19,5	8,2	0,35	9,08	21	21,0			98,2
60	265	1 081	19,1	21	8,2	0,39	9,16	30	30,0			98,1
75	260	1 035	20,2	23	8,21	0,46	9,25	37,5	37,5			98,0
90	255	988	21,4	25,2	8,21	0,52	9,28	45	45,0			97,9
105	245	934	22,1	28,1	8,22	0,61	9,31	52,5	52,5			97,8
120	230	865	22,7	31,1	8,23	0,71	9,33	59	59,0			97,7
135	210	778	23,3	35	8,.23	0,85	9,34	65	65,0			97,6
140	180	653	24,2	41,4	8,22	1,13	9,33	70	70,0			97,3
140	395	1 734	15,6		(CWF)	0,53	9,22					96,92

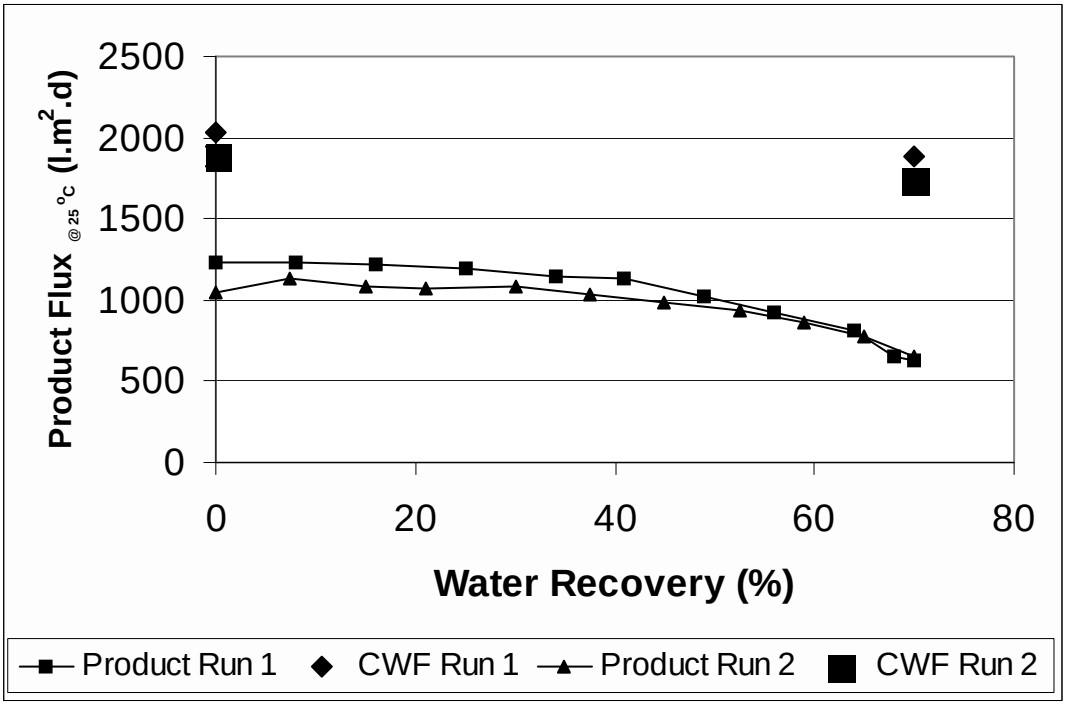


Figure 17.2 : Permeate flux as a function of percentage water recovery.

The initial permeate flux was approximately 1 200 ℓ/m^2.d (run 1, Figure 17.2) and the permeate flux decreased as a function of percentage water recovery. The permeate flux of the second run was slightly less than that of the first run in the beginning of the run but was almost the same towards the end of the run.

The initial and CWFs at the end of the runs were approximately the same. Therefore, it again appears that membrane fouling should not be a problem with the treatment of this type of effluent if flow reversal and sponge ball cleaning are applied.

The permeate flux through the polyamide membrane module was significantly higher than through the cellulose acetate membrane module (see Figures 17.2 and 17.1). Therefore, more product water can be produced with the polyamide membranes (0,81 m^2 membranes area per module) than with the cellulose acetate membranes (1,75 m^2 membrane area per module).

Table 17.6: Chemical composition of the RO feed, product and brine (polyamide membranes)..

	Feed	Product	Brine	Removal	Feed	Product
Constituents				(%)	(g)	(g)
TDS	7 070	146	21 755	97,9	707,000	0,003
TSS	368	20	304	94,6	36,800	1,400
Ammonia-N	980	81	1 759	91,7	98,000	5,670
Nitrate-N	4,95	0,05	0,06	99,0	0,495	0,004
Total P	6,91	0	10,1	100,0	0,691	0,000
Alkalinity as CaCO3	5 552	284	7 666	94,9	555,200	19,880
Chloride	2 625	33,9	7 062	98,7	262,500	2,373
Sulphate	149	0	433,6	100,0	14,900	0,000
Fluoride	0,9	0,002	3,51	99,8	0,090	0,000
Silicon	23,6	0,099	72,4	99,6	2,360	0,007
Calcium	70,6	0,205	56	99,7	7,060	0,014
Magnesium	141	0,128	503	99,9	14,100	0,009
Potassium	1 150	11,5	2 630	99,0	115,000	0,805
Sodium	1 620	12,5	5 200	99,2	162,000	0,875
Barium	0,495	0,007	1,61	98,6	0,050	0,000
Strontium	1,09	0,007	3,02	99,4	0,109	0,000
Chromium (Total)	0,17	0,001	0,43	99,4	0,017	0,000
Iron	3,16	0,009	8,47	99,7	0,316	0,001
Chromium (Total)	0,17	0,001	0,43	99,4	0,017	0,000
Iron	3,16	0,009	8,47	99,7	0,316	0,001
Lead	0,126	0	0,354	100,0	0,013	0,000
Manganese	0,382	0	0,839	100,0	0,038	0,000
Nickel	0,2	0,002	0,753	99,0	0,020	0,000
TOC	112,5	20,45	1 792	81,8	1,432	53,760
COD	2 000	46	7 200	97,7	3,220	216,000
Phenols	0,34	0,04	1,49	88,2	0,003	0,045
Conductivity (mS/m)	1 719	53	4 140	96,9		
pH	8,19	9,22	8,22			

Feed	100
Product	70
Brine	30

Higher TDS removal was obtained with the polyamide membranes (97,7%) than with the cellulose acetate membranes (96,1%) (Tables 17.6 and 17.3). Conductivity removals were 96,9% for the polyamide and 93,2% for the cellulose acetate membranes. (Note: conductivity measurements are usually more accurate than TDS measurements). A similar ammonia-nitrogen removal was obtained with the polyamide membranes (980 to 81 mg/ℓ, 91,7% removal) than with the cellulose acetate membranes (882 to 82 mg/ℓ, 90,7%). Between 98 to 100% removals of chloride, sulphate, calcium, magnesium, potassium and sodium were obtained. Better lead, nickel and phenol removals were also obtained with the polyamide membranes. Therefore, it appears that the polyamide membranes should perform better for the desalination/concentration of the leachate than the cellulose acetate membranes.

18. DETERMINATION OF THE PRELIMINARY ECONOMICS OF THE RO PROCESS FOR THE TREATMENT OF THE MUNICIPAL SOLID WASTE LEACHATE

The preliminary economics of the RO process to treat 250 kℓ/d of the MSWL was estimated from the semi-tech scale batch tests (see 17.2 and 17.3).

The capital cost to treat 250 kℓ/d of the MSWL with cellulose acetate membranes is estimated at R2,0 million. Operational costs are estimated at R11,0/kℓ.

The capital cost to treat 250 kℓ/d of the MSWL with the polyamide membranes is estimated at R8,1 million. Operational costs are estimated at R15,0/kℓ.

19. EVALUATION OF TUBULAR AND SPIRAL WRAP REVERSE OSMOSIS ON PILOT SCALE AT THE BISASAR ROAD WASTE DISPOSAL SITE FOR THE TREATMENT OF THE MUNICIPAL SOLID WASTE LEACHATE

Feed-and-bleed RO (tubular cellulose acetate and polyamide membranes) and continuous RO (spiral wrap membranes) pilot tests were conducted at Bisasar Road waste disposal site to further evaluate the performance and the fouling potential of the leachate for the membranes and to determine the economics of the process. Permeate flux was measured as a function of time and the initial permeate flux was compared with the permeate flux after approximately 500 hours of operation to evaluate the fouling potential of the leachate for the membranes. The chemical composition of the RO feed, product and brine was also determined. Process design criteria for a full-scale application as well as the economics of the process can be derived from the experimental data. The tests were performed according to the methods outlined in Section 3 (see 3.14).

19.1 Tubular Cellulose Acetate Membranes

The detailed experimental conditions and results are shown in Appendix H.

Permeate flux as a function of time is shown in Figure 19.1. The initial permeate flux was approximately 550 ℓ/m^2.d and initially declined as a result of membrane compaction and membrane fouling. The permeate flux decreased to approximately 200 ℓ/m^2.d after approximately 230 hours of operation and then increased to approximately 300 ℓ/m^2.d (approximately 350 hours of operation) and remained at about 300 ℓ/m^2.d until the end of the run (approximately 500 hours of operation). The increase in the permeate flux after approximately 350 hours of operation can be ascribed to a decrease in the feed concentration.

The initial CWF was about 700 ℓ/m^2.d. (Figure 19.1). The CWF remained at about 700 ℓ/m^2.d for the first 50 hours of operation and then declined to approximately 600 ℓ/m^2.d. Sponge ball cleaning followed by a STPP and EDTA cleaning after approximately 51 hours of operation increased the CWF from 610 to 639 ℓ/m^2.d. Clean water flux remained at approximately 600 ℓ/m^2.d until approximately 100 hours of operation and declined to approximately 550 ℓ/m^2.d after approximately 220 hours of operation. Cleaning of the membranes with nitric acid and STPP and EDTA did not increase the CWF. The run was proceeded and the CWF remained at approximately 500 ℓ/m^2.d until approximately 320 hours of operation when the membranes were cleaned with nitric acid and STPP and EDTA. Clean water flux increased from 479 to 518 ℓ/m^2.d after cleaning. Clean water flux remained at approximately 500 ℓ/m^2.d and was measured at 540 ℓ/m^2.d after approximately 400 hours of operation. Two cleanings with citric acid only increased the CWF to 548 ℓ/m^2.d. Cleaning with Ultrasil P3 50 had no effect on the CWF. A further cleaning with citric acid increased the CWF to 562 ℓ/m^2.d. Two further cleanings with EDTA and SLS increased the CWF to 586 ℓ/m^2.d. A further cleaning after 494 hours of operation with EDTA and SLS did not improve the CWF (522 ℓ/m^2.d). Sponge ball cleaning increased the CWF from 487 to 495 ℓ/m^2.d after 500 hours of operation. Cleaning of the membranes with phosphoric acid increased with CWF from 495 to 522 ℓ/m^2.d after 500 hours of operation. The run was then

terminated. It appears that it should be possible to control membrane fouling with regular chemical cleanings.

The pressure drop across the membranes as a function of time is shown in Figure 19.2 The pressure drop across the membranes was constant for more than 200 hours of operation before it started to increase. The increase in pressure drop, however, could be ascribed to a faulty pressure gauge.

The conductivity of the RO feed, permeate and brine is shown in Figure 19.3. The initial RO feed conductivity was 2 000 mS/m and decreased to about 1 500 mS/m towards the end of the run. The RO brine conductivity increased from 2 000 mS/m in the beginning of the run to approximately 4 500 mS/m after about 50 hours of operation. The brine conductivity reached a high of 5 000 mS/m after about 200 hours of operation and then decreased to approximately 3 000 mS/m towards the end of the run.

The permeate conductivity was approximately 50 mS/m in the beginning of the run and was approximately 500 mS/m after 220 hours of operation. It then increased to 1 000 mS/m, decreased and remained at 500 mS/m until the end of the run.

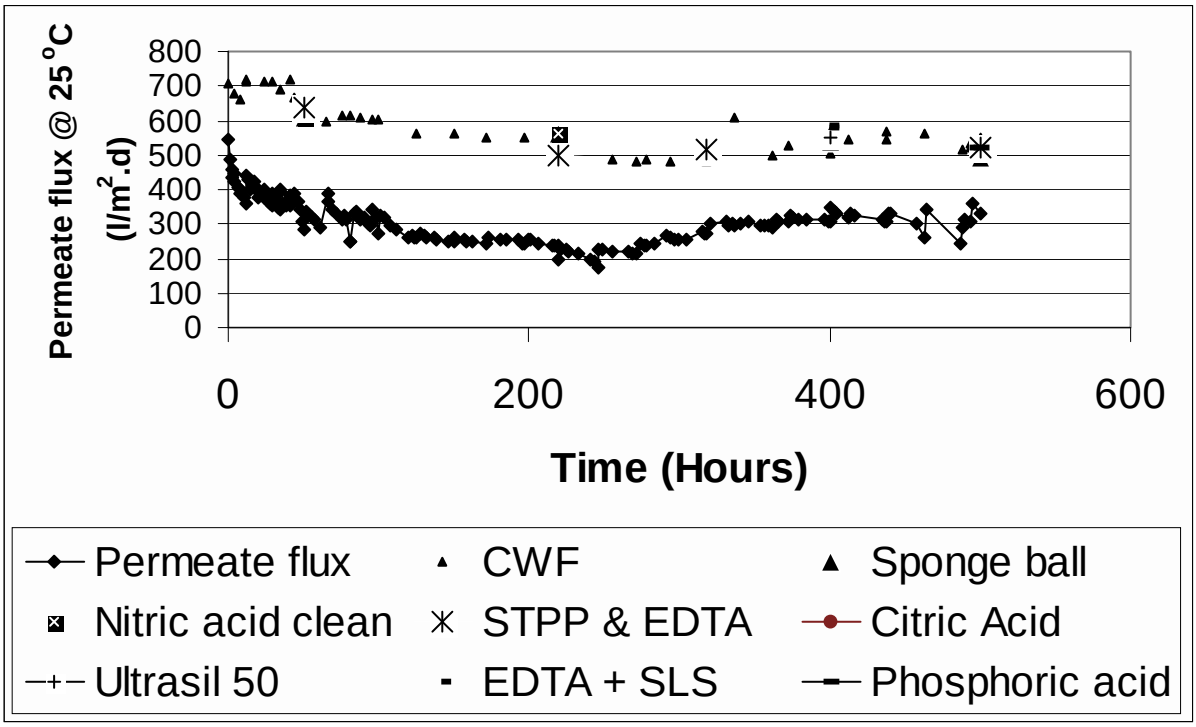


Figure 19.1 : Permeate flux as a function of time.

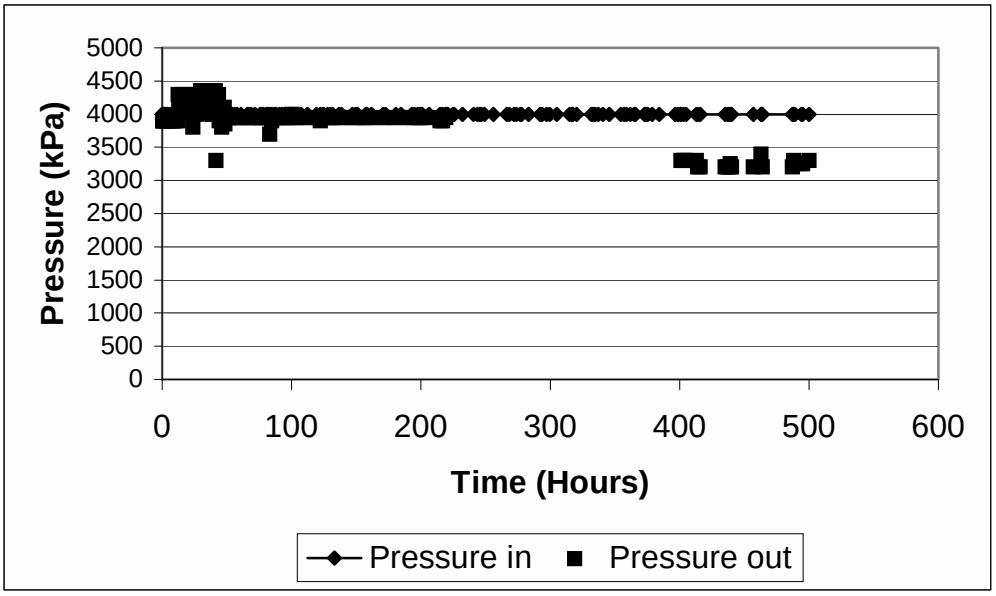


Figure 19.2 : Pressure drop across the membranes as a function of time.

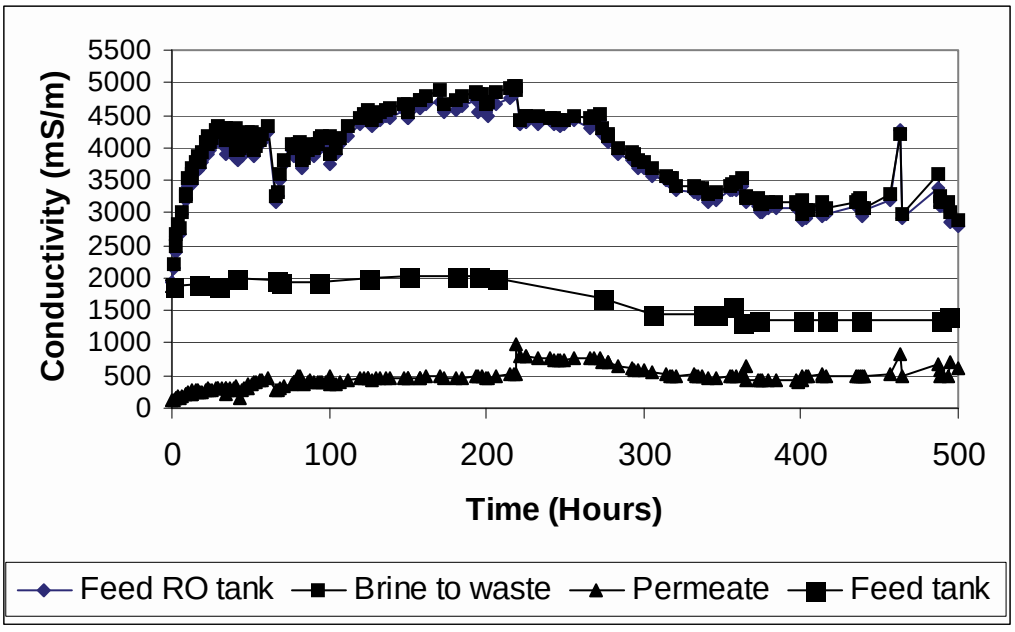


Figure 19.3 : Conductivity of RO feed, permeate and brine as a function of time.

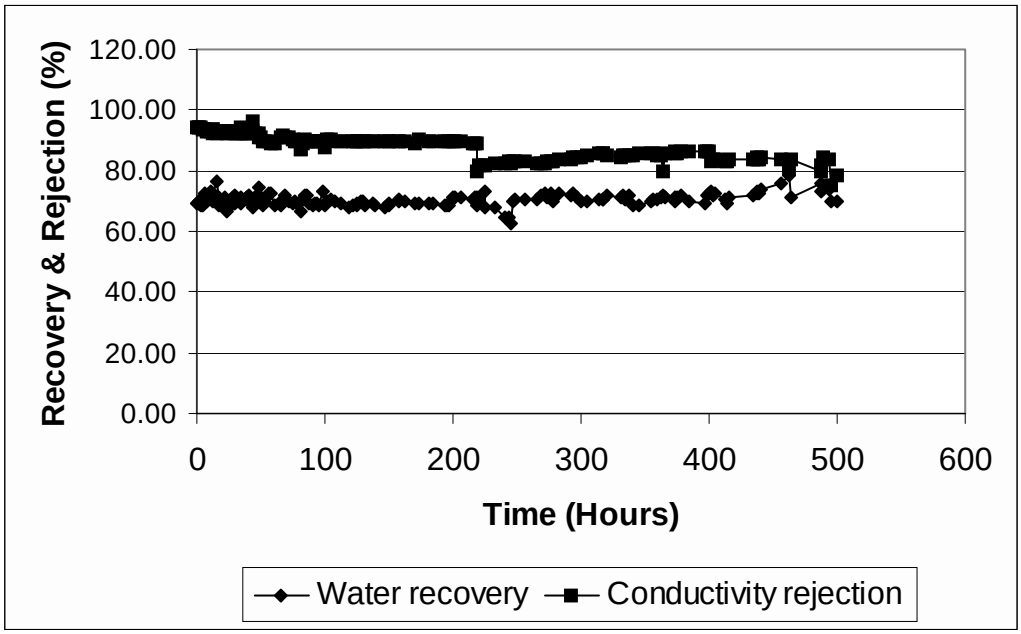


Figure 19.4 : Water recovery and conductivity rejection as a function of time.

The water recovery and conductivity rejection as a function of time is shown in Figure 19.4. The water recovery was kept at approximately 70% for the entire run. The initial conductivity rejection was approximately 93% and declined to 90% and remained at 90% for approximately 200 hours of operation. The conductivity rejection then declined to 80%, remained at 80% for some time, increased somewhat and was approximately 80% at the end of the run. The decrease in conductivity rejection could be ascribed to membrane fouling. This, however, could be expected to occur in the last RO module in the treatment train.

The pH of the RO feed, product and brine as a function of time is shown in Figure 19.5. The pH of the RO feed in the feed tank was adjusted to a pH of approximately 6,5 with hydrochloric acid. The pH of the RO brine (feed in RO unit) was approximately 7,5 while the RO permeate had a pH of approximately 6,5.

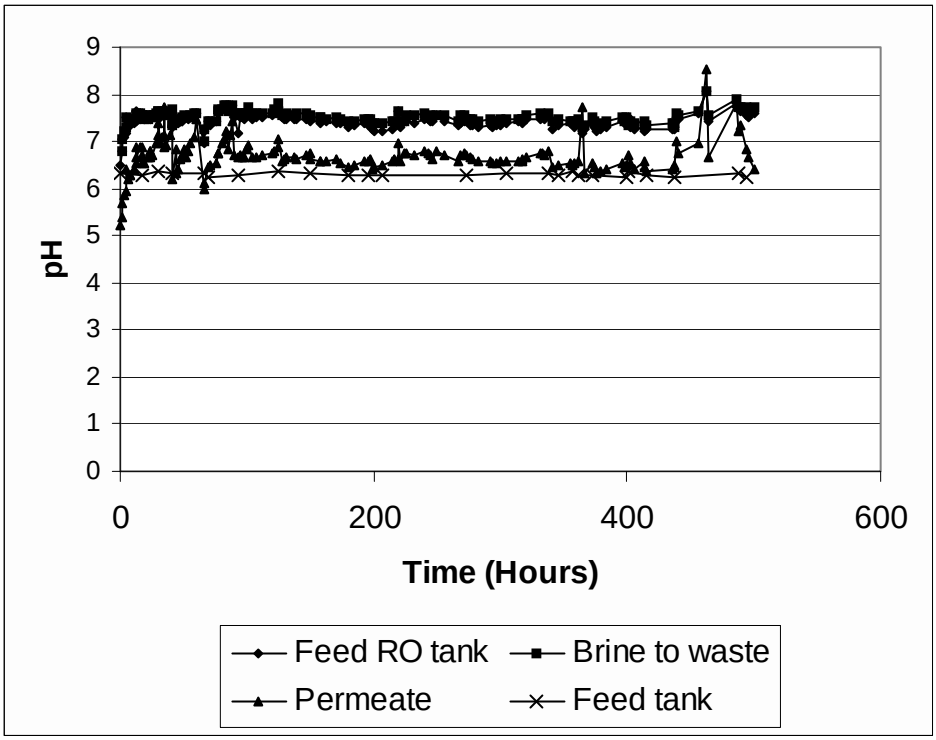


Figure 19.5 : pH of the RO feed, product and brine as a function of time.

The initial and final (after 500 hours of operation) permeate fluxes as a function of percentage water recovery are shown in Figure 19.6 (experimental conditions and results are shown in Table 19.1). The permeate flux after approximately 500 hours of operation was somewhat lower than the initial permeate flux (Figure 19.6). This indicates membrane fouling. However, membrane fouling could be expected to occur in the last RO stage and the reduction in permeate flux was not that big. The CWF before and after the run on the fouled membrane surface was approximately the same. The CWF at the end of the second run, however, was significantly lower than the initial CWF on the new membrane.

Table 19.1 : Experimental conditions and results of the batch treatment of the leachate after 500 hours of operation (70% water recovery, cellulose acetate membranes (1,75 m^2), pH adjusted to pH 6,5 (hydrochloric acid)).

Time	Flux	Flux @ 25 °C	Temperature	Feed		Product		Recovery		Pressure (kPa)		Rejection
(min)	(mℓ/min)	(ℓ/d.m2)	(°C)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ℓ)	(%)	(in)	(out)	(%)
0	590	522	22		(CWF)			100	0	4 000	3 800	
0	540	478	22	14,05	6,4			0	0,0	4 000	3 800	100,0
15	560	478	23,5	14,98	6,7	3,26	5,92	8	8,0			78,2
30	560	455	25,5	15,91	6,88	3,48	5,97	16	16,0			78,1
45	570	449	26,75	17,01	7,13	3,82	6,21	26	26,0			77,5
60	580	441	28	18,63	7,33	4,29	6,43	35	35,0			77,0
75	580	433	28,75	20,2	7,54	4,86	6,83	44	44,0			75,9
90	560	412	29,25	22,6	7,66	5,56	7,37	52	52,0			75,4
105	550	399	29,75	25,3	7,74	6,46	7,78	60	60,0			74,5
120	530	379	30,25	29	7,8	7,61	8,01	65	65,0			73,8
135	500	350	31	33,1	7,83	9,14	8,11	70	70,0			72,4
135	620	548	22		(CWF)			70	70			100,0
Composite sample used membrane			14,05	33,1	7,83	5,03	6,88					64,2
Composite sample new membrane			17,59	41,2		1,2						93,2

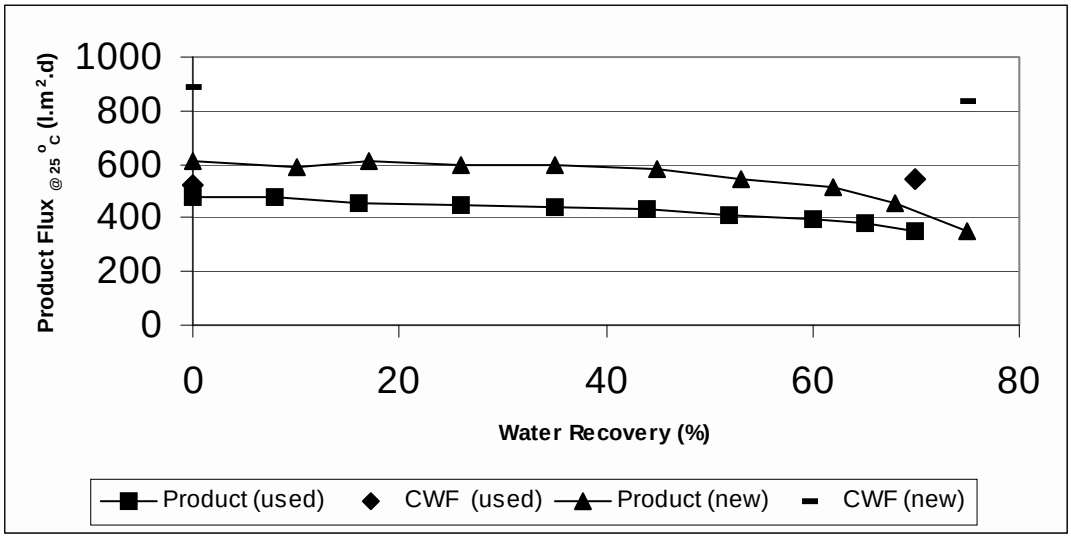


Figure 19.6 : Initial and final (after 500 hours of operation) permeate fluxes as a function of percentage water recovery.

The chemical composition of the RO feed, product and brine is shown in Tables19.2 to 19.8. The analysis shown in Tables 19.2 to 19.7 were obtained after the RO plant (feed-and-bleed) was in operation for approximately 50 to 400 hours. The conductivity removal after 50 hours of operation, for example, was 91,4% and had decreased to 86,4% after 400 hours of operation (TDS from 94,8 to 92,5%). The reduction in salinity removal can be ascribed to membrane fouling.

Table 19.2 : Chemical composition of RO feed, permeate and brine (after 50 h of operation).

Constituent*	Raw Feed	pH Adjusted	RO Feed tank	Permeate	% Removal
Alkalinity	5 242	2 850	6 615	450	93,2
Ammonia (free)	1 100	1 150	2 300	132	94,3
Barium	<0,10	<0,10	1,0	<0,10	
Calcium	48	46	120	<5,0	<95,8
Chloride	3 350	5 050	10 950	3 450	68,5
Chrome	0,16	0,18	0,26	0,22	
COD	800	1 320	2 350	24	99,0
Conductivity (mS/m)	1 827	1 991	4 240	365	91,4
Iron	3,8	3,6	7,8	5,2	
Lead	<0,05	<0,05	<0,05	<0,05	
Magnesium	147	104	306	2,8	99,1
Manganese	<0,10	<0,10	0,30	<0,10	
Nickel	0,22	<0,10	0,40	0,12	
Nitrate	0,10	0,10	0,10		
PH	7,8	6,7	7,6	7,2	
Phenols	8,5	8,7	22	2,2	
Potassium	996	48	2 074	103	
Silicon	18	18	42	20	
Sodium	1 873	1 570	3 700	222	94,0
Strontium	0,72	0,67	1,7	<0,10	
Sulphate	21	<10	<10	<10	
Suspended solids	104	181	411	2,0	
TDS	8 740	10 227	24 929	1 293	94,8
Total phosphate	5,0	5,5	9,5	2,1	

* Concentration in mg/ℓ, unless otherwise stated.

Table 19.3 : Chemical composition of the RO feed, permeate and brine (after 107 hours of operation).

Constituent*	RO Feed Tank	Permeate	Brine	% Removal
Alkalinity ($CaCO_3$)	6 650	480	6 700	92,8
Ammonia (Free) (N)	1 339	295	1 732	78,0
Barium	1,5	0,05	1,4	
Calcium	119	2,5	<5	97,9
Chloride	10 927	869	11 164	92,0
Chrome	1,24	0,32	0,44	
COD	3 675	257	3 370	93,0
Conductivity (mS/m)	4 050	409	4 150	89,9
Iron	8	1,2	0,28	
Lead	0,025	0,025	0,05	
Magnesium	276	6,9	3,8	97,5
Manganese	0,05	0,05	0,01	
Nickel	0,32	0,6	0,4	
Nitrate (N)	54	7,2	34	
PH	7,5	7,5	7,6	
Phenols	23	2,6	21	
Potassium	1 952	145	165	
Silicon	66	12	16	
Sodium	3 168	205	272	93,5
Strontium	2	0,1	0,1	
Sulphide	26	5	32	80,8
Suspended Solids	251	64	281	
TDS	24 312	1 660	25 732	93,2
Total Phosphate (P)	14	0,28	14	

* Concentration in mg/ℓ unless otherwise stated

Table 19.4 : Chemical composition of the RO feed, permeate and brine (after 175 hours of operation).

Constituent*	RO Feed Tank	Permeate	Brine	% Removal
Alkalinity ($CaCO_3$)	5 300	340	5 450	93,6
Ammonia (Free) (N)	1 987	306	2 719	84,6
Barium	1,8	0,05	2	
Calcium	68	2,5	67	96,3
Chloride	12 840	1 042	12 948	91,9
Chrome	0,24	0,05	0,05	
COD	3 380	110	3 440	96,7
Conductivity (mS/m)	4 600	456	4 690	90,1
Iron	40	0,18	4	
Lead	0,025	0,025	0,05	
Magnesium	172	2,5	169	98,5
Manganese	0,05	0,05	<0,1	
Nickel	0,32	0,05	0,46	
Nitrate (N)	43	5,5	47	
PH	7,4	6,5	7,4	
Phenols	27	2,3	29	
Potassium	1 145	138	1 130	
Silicon	45	11	46	
Sodium	1 976	178	1 919	91,0
Strontium	1,2	0,05	1,2	
Sulphide	33	5	44	84,8
Suspended Solids	340	31	251	
TDS	29 694	1 934	30 450	93,5
Total Phosphate (P)	17	0,26	18	

* Concentration in mg/ℓ unless otherwise stated.

Table 19. 5 : Chemical composition of the RO feed, permeate and brine (after 250 hours of operation).

Constituent*	RO Feed Tank	Permeate	Brine	% Removal
Alkalinity ($CaCO_3$)	3 766	401	3 766	89,4
Ammonia (Free) (N)	1 869	377	1 869	79,8
Barium	4,2	<0,1	4,2	
Calcium	120	<5	120	>95,8
Chloride	11 866	1 762	11 866	85,2
Chrome	0,28	0,24	0,28	
COD	3 760	52	3 760	98,6
Conductivity (mS/m)	4 280	735	4 280	82,8
Iron	6,4	6,4	6,4	
Lead	0,05	0,05	0,05	
Magnesium	660	7	660	98,9
Manganese	0,5	0,6	0,5	
Nickel	0,44	0,52	0,44	
Nitrate (N)	0,10	2,9	0,10	
pH	7,2	6,5	7,2	
Phenols	30	4,1	30	
Potassium	4 102	221	4 102	
Silicon	107	23	107	
Sodium	6 240	266	6 240	95,7
Strontium	4,4	0,22	4,4	
Sulphide	21,0	<10	21,0	752,3
Suspended Solids	279	21	279	
TDS	28 882	3 392	28 882	88,3
Total Phosphate (P)	18	0,13	18	

* Concentration in mg/ℓ unless otherwise stated.

Table 19.6 : Chemical composition of the RO feed, permeate and brine (after 330 hours of operation).

Constituent*	RO Feed Tank	Permeate	Brine	% Removal
Alkalinity ($CaCO_3$)	3 500	281	3 710	92,0
Ammonia (Free) (N)	1 272	269	1 558	78,9
Barium	3,8	<0,1	4,1	
Calcium	186	50	202	73,1
Chloride	12 770	1 166	13 340	909
Chrome	0,2	0,2	<0,1	
COD	3 340	86	2 140	97,4
Conductivity (mS/m)	3 230	496	3 390	84,6
Iron	13	12	8,9	
Lead	0,05	0,5	0,05	
Magnesium	290	8	305	97,2
Manganese	0,8	0,8	0,9	
Nickel	0,32	0,28	<0,1	
Nitrate (N)	3,1	2,5	0,1	
pH	7,2	6,7	7,4	
Phenols	30	4,6	26	
Potassium	130	156	142	
Silicon	102	25	105	
Sodium	2 393	237	2 537	90,1
Strontium	4,2	0,14	4,6	
Sulphide	188	<10	202	>94,6
Suspended Solids	249	10	245	
TDS	23 364	2 154	23 656	90,8
Total Phosphate (P)	13	0,08	14	

* Concentration in mg/ℓ unless otherwise stated.

Table 19.7 : Chemical composition of the RO feed, permeate and brine (after 400 hours of operation).

Constituent*	RO Feed Tank	Permeate	Brine	% Removal
Alkalinity ($CaCO_3$)	3 659	240	3 800	93,4
Ammonia (Free) (N)	1 256	246	1 308	80,4
Barium	0,10	0,32	0,15	
Calcium	230	125	1 260	45,7
Chloride	10 019	1 235	9 548	87,7
Chrome	0,10	0,10	0,10	
COD	2 680	69	2 530	97,4
Conductivity (mS/m)	3 100	424	3 150	86,3
Iron	8,6	4,8	5,2	
Lead	0,05	0,05	0,05	
Magnesium	325	173	176	46,8
Manganese	0,10	0,10	0,10	
Nickel	0,10	0,10	0,10	
Nitrate (N)	21	6,5	17	
pH	7,5	6,9	7,6	
Phenols	20	0,50	20	
Potassium	1 400	787	800	
Silicon	44	26	28	
Sodium	2 782	152	1 687	94,5
Strontium	2,4	1,2	1,3	
Sulphide	389	<10	410	>97,4
Suspended Solids	244	8,0	210	
TDS	19 210	1 444	19 668	92,5
Total Phosphate (P)	13	0,5	13	

* Concentration in mg/ℓ unless otherwise stated.

Table 19.8 : Chemical composition of the RO feed, product and brine (batch) after 500 hours of operation (batch).

Constituents*		Feed	Product	Brine	Removal (%)
Total Dissolved solids		6 106	1 398	17 030	77,1
Total suspended solids		26	5	122	80,8
Ammonia nitrogen		477	195	1 519	59,1
Nitrate nitrogen		<0,20	<0,20	<0,20	#VALUE!
Total Phosphate		12	1,5	15	87,5
Alkalinity		2 138	418	6 476	80,4
Chloride		2 787	1 273	7 839	54,3
Sulphate		67	8	195	88,1
Fluoride		0,92	0,2	3,0	78,3
Silicon		23	7	61	69,6
Calcium		60	5	249	91,7
Magnesium		120	7	416	94,2
Potassium		691	231	1 730	66,6
Sodium		1 230	334	3 370	72,8
Barium		0,47	0,1	1,4	78,7
Strontium		0,92	0,1	2,9	89,1
Iron		2,9	0,03	8,7	99,0
Lead		0,36	0,24	0,58	33,3
Manganese		0,28	0,03	0,81	89,3
Nickel		0,17	0,03	0,51	82,4
Total Organic Carbon		351	137	1 397	61,0
Chemical Oxygen Demand		1 500	350	5 350	76,7
Phenols		3	0,04	5,5	98,7
Conductivity (mS/m)		1 405	503	3310	64,2
pH		6,4	6,88	7,83	

Feed	100	litre
Product	70	litre
Brine	30	litre

* Concentration in mg/ℓ unless otherwise stated.

The TDS and conductivity removals were only 77,1 and 64,2% (after 500 hours of operation). The salinity removal has decreased significantly from the first batch run (96,1% TDS and 93,2% conductivity removal on a clean surface, see Table 17.3). This also showed that membrane fouling took place.

19.2 Tubular Polyamide Membranes

The detailed experimental conditions and results are shown in Appendix I.

Permeate flux as a function of time is shown in Figure 19.7. The initial permeate flux was about 1 200 ℓ/m^2.d and decreased as a result of membrane compaction and membrane fouling. Permeate flux decreased to about 500 ℓ/m^2.d after 100 hours of operation and remained at 500 ℓ/m^2.d until about 200 hours of operation and decreased to about 300 ℓ/m^2.d after 330 hours of operation. Permeate flux reached a low of about 100 ℓ/m^2.d after 400 hours of operation and remained at about 200 to 250 ℓ/m^2.d until the end of the run.

The initial CWF was about 1 900 ℓ/m^2.d and the CWF declined as a function of time as a result of membrane fouling. Clean water flux was 1 398 ℓ/m^2.d after 23 hours of operation and was measured as 1 577 ℓ/m^2.d after preservation. Therefore, preservation with SMBS increased the CWF. Clean water flux declined further and was determined at 1 101 ℓ/m^2.d after 100 hours of operation. Preservation with SMBS and cleaning of the membranes with Ultrasil 10 increased the CWF to 1 378 and 1 404 ℓ/m^2.d, respectively. Clean water flux was measured as 1 209 ℓ/m^2.d after 118 hours of operation and decreased to 1 191 ℓ/m^2.d after 140 hours of operation. Preservation increased the CWF to 1 351 ℓ/m^2.d (140 h operation) and the CWF was 1 102 ℓ/m^2.d after 164 hours of operation. Clean water flux then decreased to 960 ℓ/m^2.d after 186 hours of operation. Preservation of the membranes increased the CWF to 1 369 ℓ/m^2.h and the CWF declined to 1 018 (109 h) and 676 ℓ/m^2.d (after 256 h). Cleaning of the membranes with Ultrasil 10 increased the CWF to 1 056 ℓ/m^2.d. Therefore, the CWF could be significantly increased with cleaning. Clean water flux declined further and was measured as 773 (279 h) and 537 ℓ/m^2.h (330 h). Cleaning with Ultrasil 10 improved the CWF to 1 013 ℓ/m^2.h. The CWF again declined and was determined at 480 ℓ/m^2.d after 405 hours of operation. Cleaning with Ultrasil 10 and preservation of the membranes increased the CWF to 1 244 ℓ/m^2.d. The CWF declined further and was 451 ℓ/m^2.d after 501 hours of operation. Cleaning of the membranes with acid and Ultrasil 10 increased the CWF to 1 249 ℓ/m^2.d. Therefore, it appears that it should be possible to control membrane fouling with regular chemical cleaning.

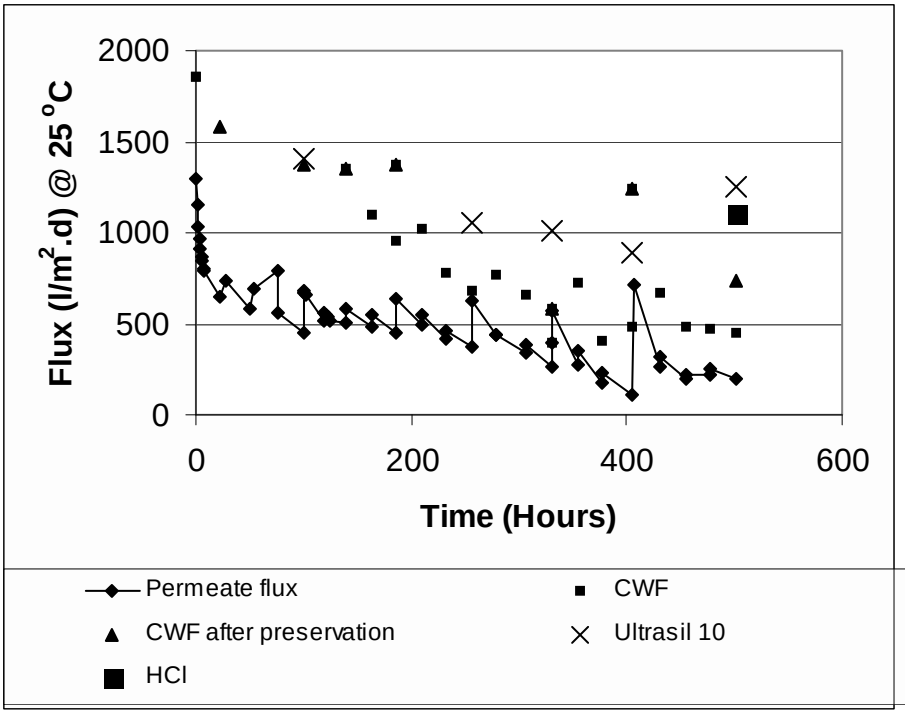


Figure 19.7 : Permeate flux as a function of time.

The pressure drop across the membranes as a function of time is shown in Figure 19.8. The pressure drop across the membranes remained almost the same for the entire run. This showed that membrane fouling should not be a serious problem.

The conductivity of the RO feed, permeate and brine is shown in Figure 19.9. The conductivity of the RO feed varied between 12 and 14 mS/cm (1 200 and 1 400 mS/m) over the test period. The RO brine conductivity reached a high of about 36 mS/cm (3 600 mS/m) and declined and was about 30 mS/cm (3 000 mS/m) at the end of the run. Permeate conductivity varied between 1 (100 mS/m) and 2 mS/cm (200 mS/m). Therefore, a significantly better quality water could be produced with the polyamide membranes than with the cellulose acetate membranes.

Water recovery and conductivity rejection are shown in Figure 19.10. Water recovery was kept at approximately 70% for the entire run. The conductivity rejection varied between 95 and 97% over the test period. Therefore, a significantly better conductivity rejection was obtained with the polyamide membranes than with the cellulose acetate membranes.

The pH of the RO feed, permeate and brine as a function of time is shown in Fig. 19.11. The pH of the RO feed was adjusted to pH 7 with hydrochloric acid (feed tank). The pH of the brine (RO feed) varied between 7,9 and 8,2 over the test period. The pH of the RO permeate, was significantly higher and varied between 8,8 and 9,2 over the test period.

The initial and final (after 500 hours of operation) permeate fluxes as a function of time are shown in Fig. 19.12. The experimental conditions and results are shown in Table 19.9. The permeate flux after about 500 hours of operation was somewhat lower than the initial permeate flux. This can be expected due to membrane fouling. The reduction in permeate flux, however, was not that big. It is interesting to note that the CWF at the end of the run (after 500 hours of operation) was about the same as before the run and that the acid cleaning of the membranes had the effect to increase the CWF. Therefore, acid cleaning of the membranes will be required from time to time together with cleaning with Ultrasil 10.

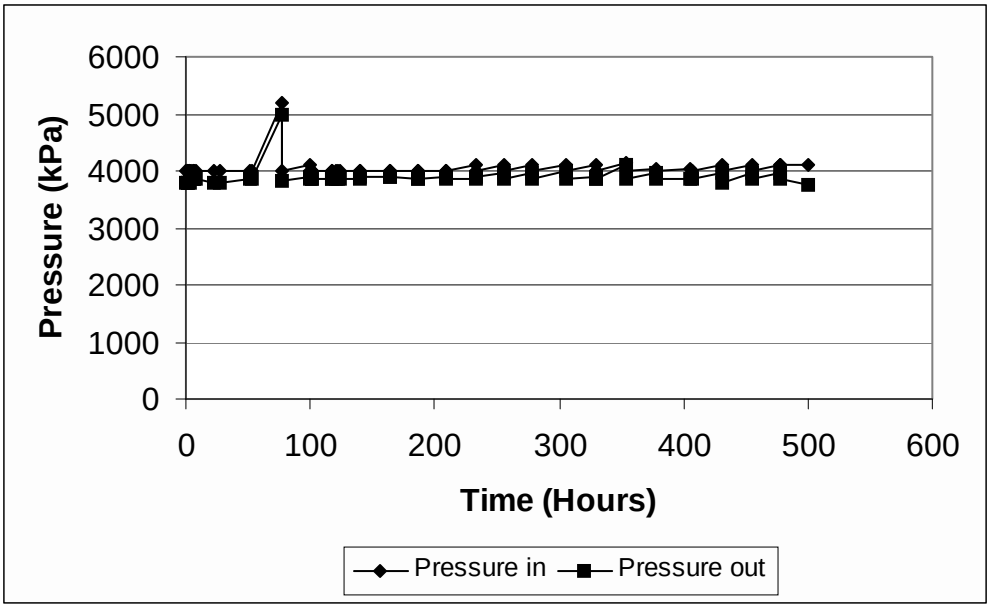


Figure 19.8 : Pressure drop across the membrane as a function of time.

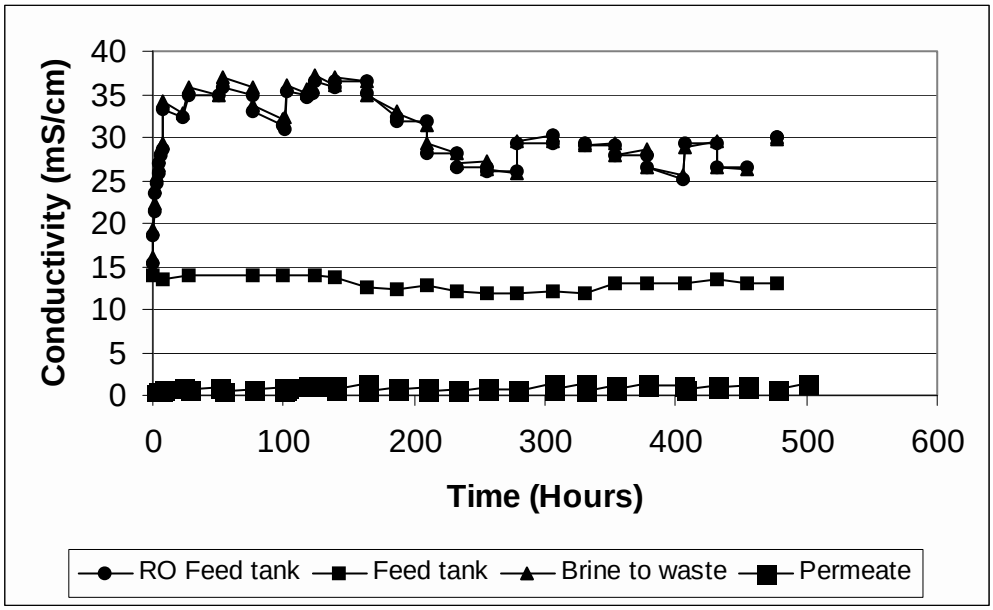


Figure 19.9 : Conductivity of RO feed, permeate and brine as a function of time.

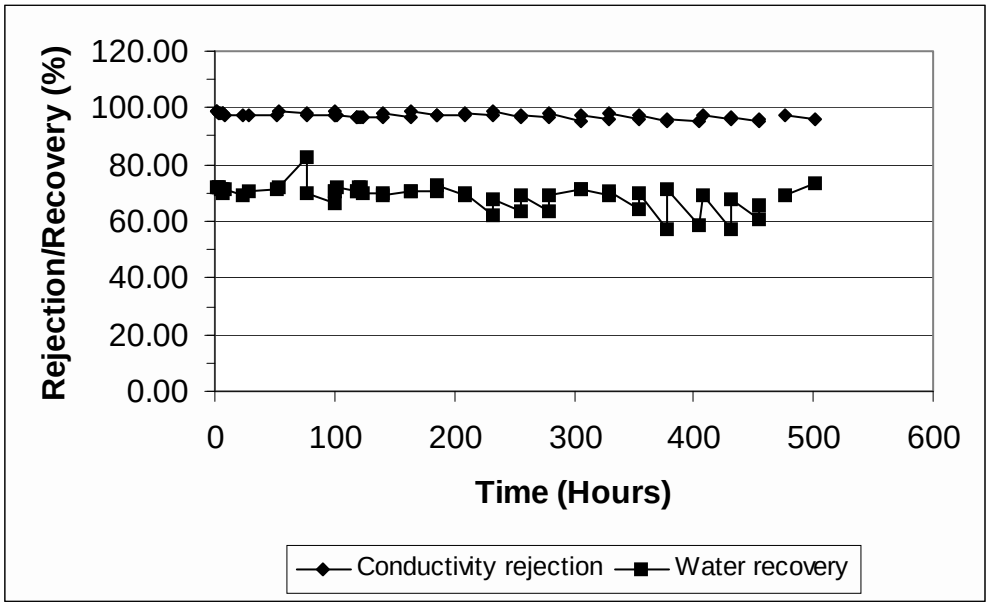


Figure 19.10 : Conductivity rejection and water recovery as a function of time.

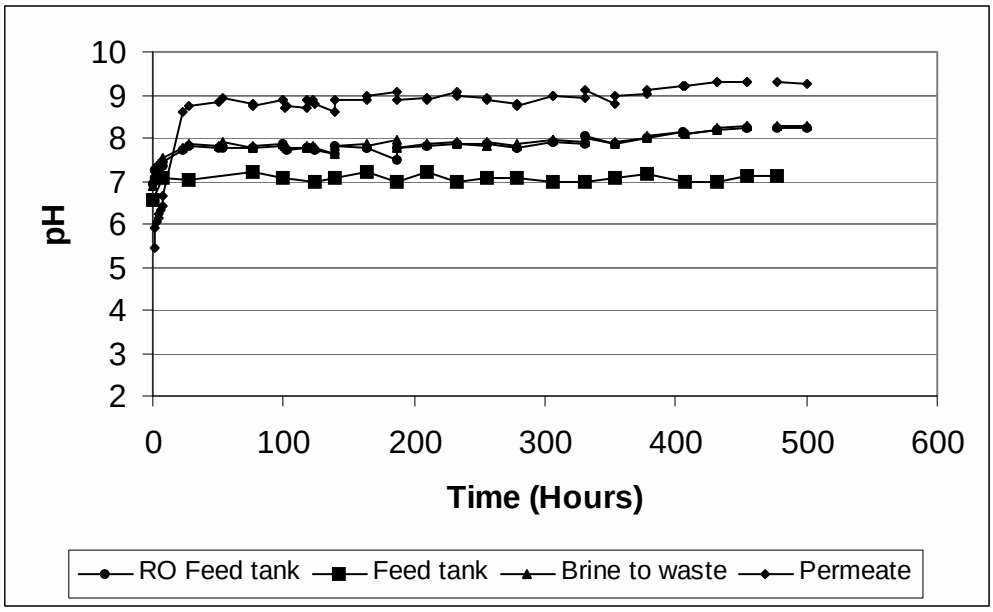


Figure 19.11 : pH of the RO feed, permeate and brine as a function of time.

Table 19.9 : Batch treatment of the leachate after 500 hours of operation. (70% water recovery, PA PCI membranes (0,81 m^2), pH adjusted to pH 7,1 (hydrochloric acid)).

Time	Flux	Flux @ 25 °C	Temperature	Feed		Product		Recovery		Pressure (kPa)		Rejection
(min)	(mℓ/min)	(ℓ/d.m2)	(°C)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ℓ)	(%)	(in)	(out)	(%)
0	745	1 275	26,5		(CWF)			100	0	4 000	3 850	
0	570	1 001	25,5	13,1	7,03			0	0.0	4 000	3 850	
15	600	1 000	27,5	14,7	7,09	0,35	5,42	8	8,0	4 000	3 850	97,6
30	615	957	30	15,4	7,24	0,46	5,92	17,5	17,5			97,0
45	610	908	31,5	17,5	7,46	0,58	6,62	27	27,0			96,7
60	620	882	33	19,8	7,55	0,76	7,63	36	36,0			96,2
75	580	793	34,25	22,8	7,61	0,89	8,26	45	45,0			96,1
90	550	727	35,25	25,6	7,64	1,05	8,43	54	54,0			95,9
105	510	652	36,25	30,4	7,63	1,28	8,52	60	60,0			95,8
120	410	506	37,25	35,8	7,62	1,58	8,56	67	67,0			95,6
130	310	375	37,75	38,9	7,64	1,8	8,59	70	70,0			95,4
135	690	1 165	27		(CWF)				0			
Clean with HCl at pH 1,8												
135	800	1 360	26,75						70			
Composite sample used membrane			13,1	38,7	7,68	0,86	8,21					93,4
Composite sample new membrane			17,19	43,4		0,56						96,7

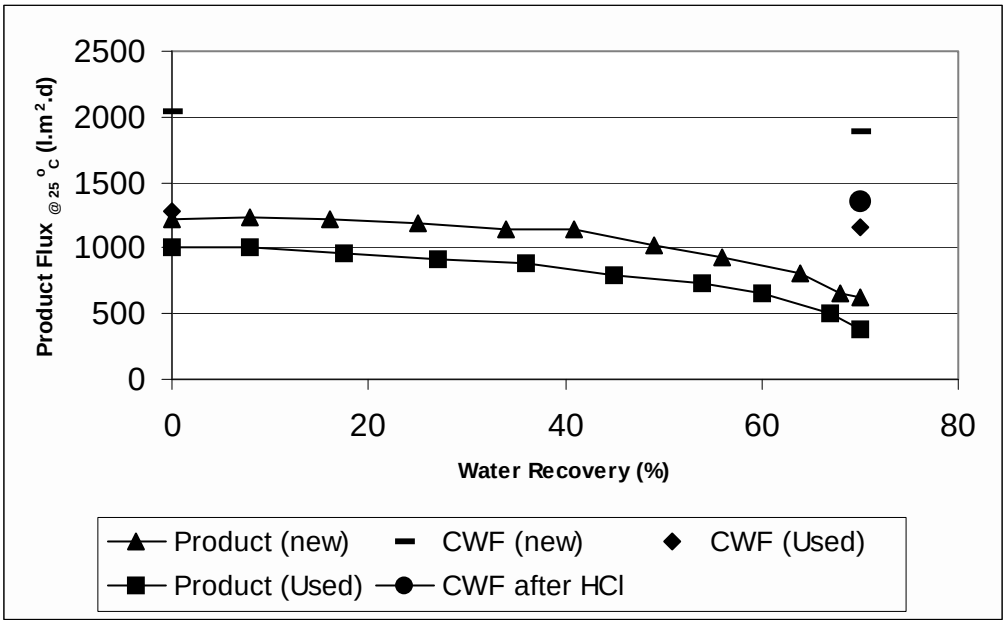


Figure 19.12 : Initial and final (after 500 hours of operation) permeate fluxes as a function of percentage water recovery.

The chemical composition of the RO feed, product and brine is shown in Tables 19.10, 19.11, 19.12 and 19.13.

Samples for chemical analysis were taken after 120 hours, 211 hours and 407 hours of operation during the feed-and-bleed RO tests. The results (Table 19.10 to Table 19.12) showed that the conductivity removal remained at 96,4 to 96,5% over the test period. The percentage removals of individual ions and organics were also good. COD removal varied between 96 and 98,2%. Chloride removal varied between 97,7 and 99,6%. Sodium removal varied between 96,2 and 98,1%. Ammonia-nitrogen removals varied between 87 and 91,2%. Therefore, excellent removals of inorganics and organics could be obtained.

The chemical composition of the leachate after 500 hours of operation (Figure 19.13, batch run) showed that the quality of the permeate that was produced after membrane fouling was still excellent. Conductivity removal was 93,4% (TDS removal 96,6%). The TDS and conductivity removals on a fresh membrane surface were 97,9% and 96,9%, respectively (see Table 17.6). Therefore, an excellent quality water could be produced with RO treatment of the leachate using polyamide membranes.

Table 19.10 : Chemical composition of RO feed, product and brine (after 120 hours of operation).

Constituents*	Raw Feed	pH Adjusted	RO Feed tank	Permeate	Removal (%)
Alkalinity	3 450	4 412	7 621	758	90,1
Ammonia (free)	1 284	1 560	4 179	366	91,2
Barium	0,3	0,28	0,8	<0,10	
Calcium	66	69	60	<5,0	
Chloride	3 904	6 362	19 995	82	99,6
Chrome	<0,10	<0,10	<0,10	<0,10	
COD	650	505	2 545	55	97,8
Conductivity (mS/m)	1 168	1 411	3 500	124	96,5
Iron	1,2	2,4	5,4	<0,10	
Lead	<0.05	<0,05	<0,05	<0,05	
Magnesium	100	110	298	3,6	
Manganese	0,28	0,22	0,44	<0,10	
Nickel	<0,10	<0,10	<0,10	<0,10	
Nitrate	0,13	0,2	0,6	0,16	
pH	7,8	7,5	7,9	9,2	
Phenols	3,2	4,7	20	0,1	
Potassium	608	604	2 326	50	
Silicon	16	17	50	0,1	
Sodium	966	1 116	3 190	96	97,0
Strontium	0,72	0,8	2	<0,10	
Sulphate	80	119	294	15	
Suspended solids	51	81	212	2	
TDS	5 931	6 589	19 791	244	98,8
Total phosphate	4,8	5,1	4,8	0,15	

*Concentration in mg/ℓ unless otherwise stated.

Table 19.11 : Chemical composition of RO feed, product and brine (after 211 hours of operation.

Constituents*	Raw Feed	RO Feed tank	Permeate	Removal (%)
Alkalinity	3 624	7 015	650	90,7
Ammonia (free)	626	1 714	157	90,8
Barium	0,32	0,77	<0,10	
Calcium	72	40	<5,0	
Chloride	1 756	6 560	71	98,9
Chrome	<0,10	<0,10	<0,10	
COD	820	2 055	37	98,2
Conductivity (mS/m)	1 142	2 830	103	96,4
Iron	1,4	3,9	<0,10	
Lead	<0.05	0,13	<0,05	
Magnesium	110	318	4,7	
Manganese	0,24	0,49	<0,10	
Nickel	<0,10	0,15	<0,10	
Nitrate	4,9	25	0,12	
pH	8	8	9,4	
Phenols	3,9	12	0,1	
Potassium	660	1 705	751	
Silicon	20	49	2,3	
Sodium	1 260	3 027	114	96,2
Strontium	0,8	1,8	<0,10	
Sulphate	106	120	<10	
Suspended solids	77	210	7	
TDS	5 320	14 146	262	98,1
Total phosphate	3,6	2,8	3,8	

*Concentration in mg/ℓ unless otherwise stated

Table 19.12 : Chemical composition of RO feed, product and brine (after 407 hours of operation).

Constituent*	Raw Feed	RO Feed tank	Permeate	Removal (%)
Alkalinity	4 600	5 600	500	91,1
Ammonia (free)	693	998	130	87,0
Barium	0,34	0,6	<0,10	
Calcium	68	18	<5,0	
Chloride	1 962	5 866	132	97,7
Chrome	<0,10	0,11	<0,10	
COD	762	2 280	92	96,0
Conductivity (mS/m)	1 304	2 550	89	96,5
Iron	2,4	4,2	<0,10	
Lead	<0,05	0,13	<0,05	
Magnesium	113	331	3,1	
Manganese	0,22	0,5	<0,10	
Nickel	<0,10	0,21	<0,10	
Nitrate	0,7	61	0,56	
pH	8,2	8,5	9,7	
Phenols	5	12	0,1	
Potassium	708	1 748	31	
Silicon	20	45	3,1	
Sodium	1 281	3 126	60	98,1
Strontium	0,83	1,5	<0,10	
Sulphate	81	176	23	
Suspended solids	72	159	8	
TDS	6 412	15 322	404	97,4
Total phosphate	4,2	1,3	0,08	

*

*Concentration in mg/ℓ unless otherwise stated

(Note: The pH of the feed was high with the result that the ammonia nitrogen rejection was poor).

Table 19.13 : Chemical composition of the RO feed, permeate and brine after 500 hours of operation (batch).

Constituents*	Feed	Product	Brine	Removal (%)
Total Dissolved solids	5 882	198	20 620	96,6
Total suspended solids	41	1	575	97,6
Ammonia nitrogen	534	69,7	1 802	86,9
Nitrate nitrogen	0,09	0	0,38	100,0
Total Phosphate	5,81	0,152	19,5	97,4
Alkalinity	2 682	311	8 025	88,4
Chloride	2 320	105	9 286	95,5
Sulphate	164	0	397	100
Fluoride	0,948	0	2,89	100,0
Silicon	22,1	0,44	83,1	98,0
Calcium	81,4	0,617	71,5	99,2
Magnesium	133	0,995	437	99,3
Potassium	754	35,9	2 260	95,2
Sodium	1 360	59,5	4 250	95,6
Barium	0,382	0	0,61	100,0
Strontium	0,944	0,007	2,27	99,3
Iron	1,98	0,014	3,51	99,3
Lead	0,147	0	0,297	100,0
Manganese	0,297	0,012	0,49	96,0
Nickel	0,169	0,006	0,525	96,4
Total Organic Carbon	532	7,43	1 266	98,6
Chemical Oxygen Demand	1 500	8	5 400	99,5
Phenols	0.17	0,01	0,5	94,1
Conductivity	1 310	86	3 879	93,4
pH	7,03	8,21	7,68	

*Concentration in mg/ℓ unless otherwise stated

19.3 Spiral Wrap Membranes

The detailed results are shown in Appendix J (see 3.14(b) for experimental procedure).

Permeate flux as a function of time is shown in Figure 19.13(a). The clean water fluxes as a function of time is shown in Figure 19.13(b). Permeate flux started at 838 ℓ/m^2.d (CWF 989 ℓ/m^2.d) and decreased to 767 ℓ/m^2.d after 3 hours of operation. The plant was stopped and the CWF was determined at 984 ℓ/m^2.d. The membranes were preserved in a sodium metabisulfite (SMBS) solution and the CWF was 1 033 ℓ/m^2.d after preservation. Clean water flux was again measured and it was 985 ℓ/m^2.d before the plant was started. Therefore, the CWF remained approximately the same after 3 hours of operation.

Permeate flux was 917 ℓ/m^2.d when the plant was started after 3 hours of operation. This flux was significantly higher than the flux of 767 ℓ/m^2.d when the plant was stopped after 3

hours of operation. This showed that preservation of the membranes in SMBS had the effect to improve membrane performance.

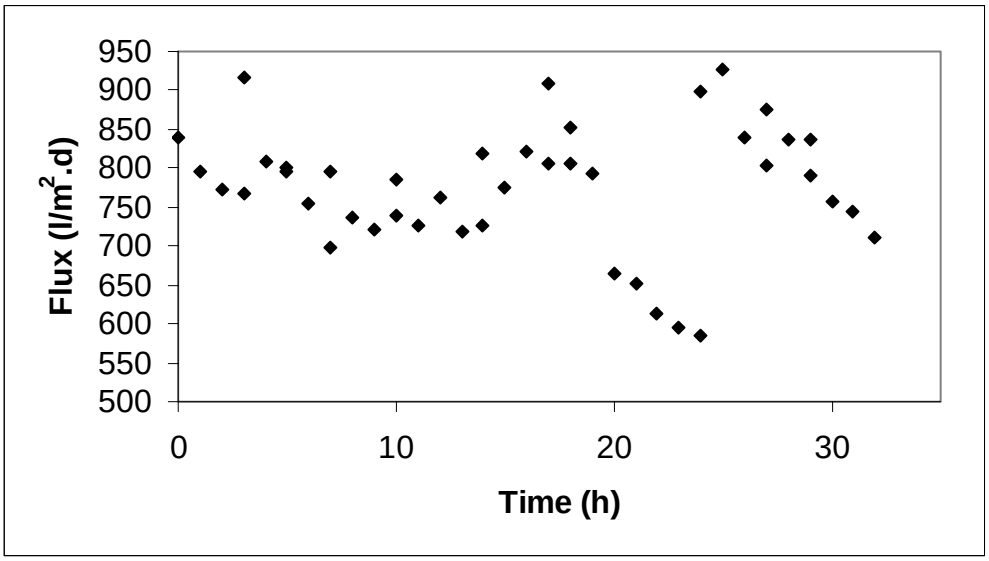


Figure 19.13(a) : Permeate flux as a function of time.

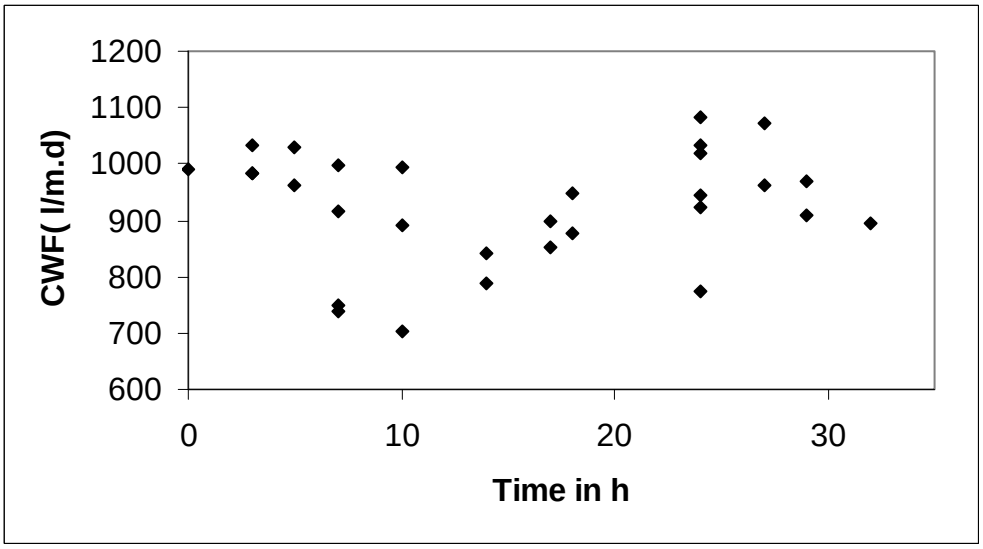


Figure 19.13(b) : Clean water flux as a function of time.

Pemeate flux then decreased from 917 ℓ/m^2.d (after preservation) to 800 ℓ/m^2.d after 5 hours of operation (CWF 962 ℓ/m^2.d). Clean water flux was 1 030 ℓ/m^2.d after preservation and permeate flux was 795 ℓ/m^2.d when the run was started (5 hours of operation). Permeate flux then decreased to 699 ℓ/m^2.d (after 7 hours of operation). Clean water fluxes were determined at 750 and 739 ℓ/m^2.d after 7 hours of operation. Therefore, there was a

significant reduction in CWF (from 1 030 ℓ/m^2.d to 739 ℓ/m^2.d). The membranes were then cleaned with Ultrasil 10 cleaning solution and the CWF increased to 916 ℓ/m^2.d and 997 ℓ/m^2.d (initial CWF 989 ℓ/m^2.d). Therefore, it appears that CWF should be restored with chemical cleaning of the membranes.

Clean water flux was 797 ℓ/m^2.d when the run was commenced (CWF 699 ℓ/m^2.d before cleaning) and decreased to 740 ℓ/m^2.d after 10 hours of operation (CWF 702 ℓ/m^2.d). Preservation of the membranes in SMBS solution and another cleaning with Ultrasil 10 increased the CWF to 995 ℓ/m^2.d (initial CWF 989 ℓ/m^2.d). Therefore, it again appears that it should be possible to restore the CWF with chemical cleaning of the membranes.

Permeate flux was 784 ℓ/m^2.d when the run was started after 10 hours of operation (CWF 740 ℓ/m^2.d before cleaning) and decreased to 726 ℓ/m^2.d after 14 hours of operation. The CWF was 789 ℓ/m^2.d and increased to 840 ℓ/m^2.d after preservation of the membrane. This flux was less than the initial CWF of 989 ℓ/m^2.d.

Permeate flux was 819 ℓ/m^2.d when the run was commenced after 14 hours of operation (726 ℓ/m^2.d before preservation) and decreased to 806 ℓ/m^2.d after 17 hours of operation. Clean water flux was 899 ℓ/m^2.d after preservation of the membranes in SMBS solution and permeate flux was 909 ℓ/m^2.d when the run was commenced (after 17 hours of operation) and decreased to 853 ℓ/m^2.d after 18 hours of operation. The CWF was 877 ℓ/m^2.d and increased to 948 ℓ/m^2.d after preservation of the membranes in SMBS solution.

Permeate flux started at 805 ℓ/m^2.d after 18 hours of operation and decreased to 584 ℓ/m^2.d after 24 hours of operation (CWF 775 ℓ/m^2.d). Therefore, there was a significant reduction in permeate and CWFs after 24 hours of operation and the membranes were cleaned with Ultrasil 10 cleaning solution. Clean water flux increased to 929 and 945 ℓ/m^2.d after cleaning and preservation of the membranes. Therefore, CWF could almost be restored to its initial value of 989 ℓ/m^2.d. Further cleaning of the membranes with acid and Ultrasil (2 times) increased CWF to 1 032 ℓ/m^2.d. Therefore, CWF could be restored (CWF 1 083 ℓ/m^2.d after preservation).

Permeate flux was 900 ℓ/m^2.d when the run was commenced after 24 hours of operation (584 ℓ/m^2.d before cleaning) and decreased to 802 ℓ/m^2.d after 27 hours of operation (CWF 961 ℓ/m^2.d). Clean water flux was 1 073 ℓ/m^2.d after preservation of the membranes and permeate flux was 875 ℓ/m^2.d when the run was commenced after 27 hours of operation (802 ℓ/m^2.d before preservation). Permeate flux then decreased to 838 ℓ/m^2.d after 29 hours of operation (CWF 970 ℓ/m^2.d). Clean water flux was 908 ℓ/m^2.d after preservation of the membranes and permeate flux was 789 ℓ/m^2.d when the run was commenced after 29 hours of operation and decreased to 711 ℓ/m^2.d (838 ℓ/m^2.d initial flux) when the run was terminated after 32 hours of operation (CWF 893 ℓ/m^2.d). Preservation of the membranes in SMBS solution increased the CWF to 1 011 ℓ/m^2.d. Therefore, the RO tests over an approximately 32 hour period have shown that it should be possible to control membrane fouling with regular chemical cleaning of the membranes. However, longer term tests should

be conducted to develop a proper membrane cleaning strategy when using spiral wrap membranes for the treatment of the leachate.

The RO feed inlet and outlet pressures are shown in Figure 19.14. The pressure drop across the membrane remained constant over the test period showing that severe membrane fouling was not experienced.

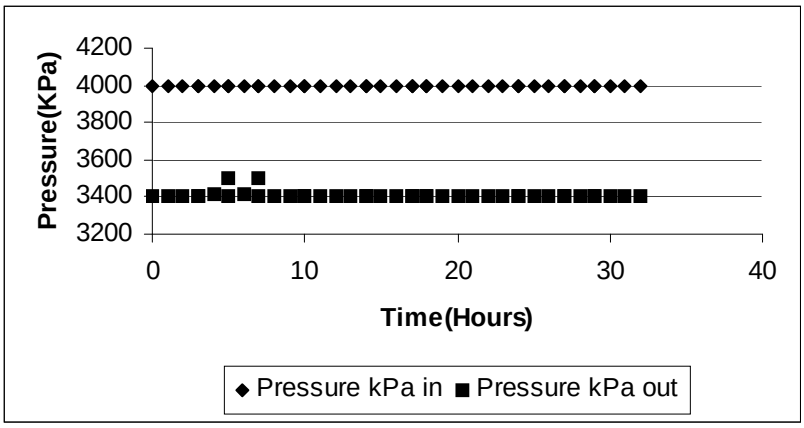


Figure 19.14: Pressure drop across membrane as a function of time.

The conductivity of the RO feed, permeate and brine is shown in Figure 19.15. The feed conductivity varied between approximately 8 and 14 mS/cm over the test period. Permeate conductivity varied between approximately 0,1 and 0,8 mS/cm while the brine conductivity varied between approximately 16 and 24 mS/cm. Permeate conductivity was 0,36 mS/cm when the run was started and was 0,56 mS/cm when the run was terminated after 32 hours of operation. Therefore, there was a slight reduction in permeate quality at the end of the run.

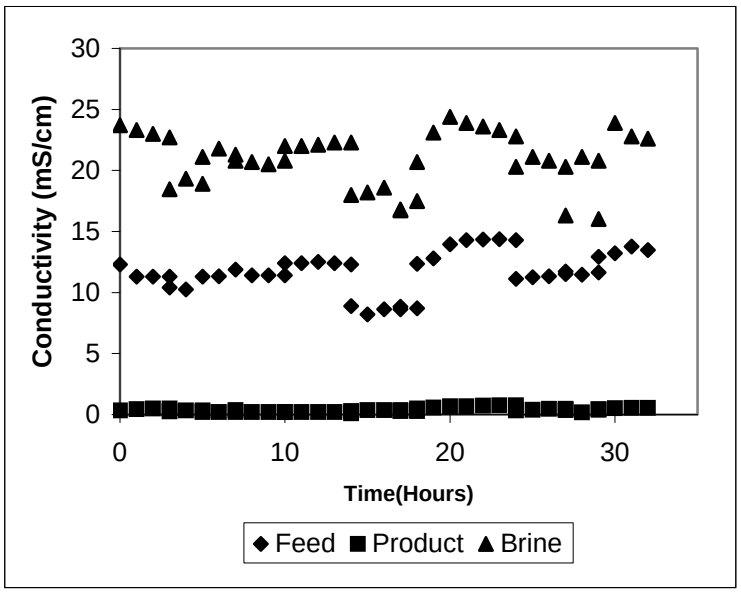


Figure 19.15 : Conductivity of RO feed, permeate and brine as a function of time.

The water recovery and conductivity rejection as a function of time are shown in Figure 19.16. The initial conductivity rejection was 98,9 percent and was 95,9 percent at the end of the run. Therefore, there was a reduction in salt removal that could be ascribed to membrane fouling. The initial water recovery was 53,6 percent and water recovery had declined to 44,2% at the end of run. The lower water recovery at the end of the run could also be ascribed to membrane fouling. However, it is clear from Figure 19.16 that it should be possible to restore water recovery after membrane cleaning (after 7, 14, 17, 24, 27 and 29 hours).

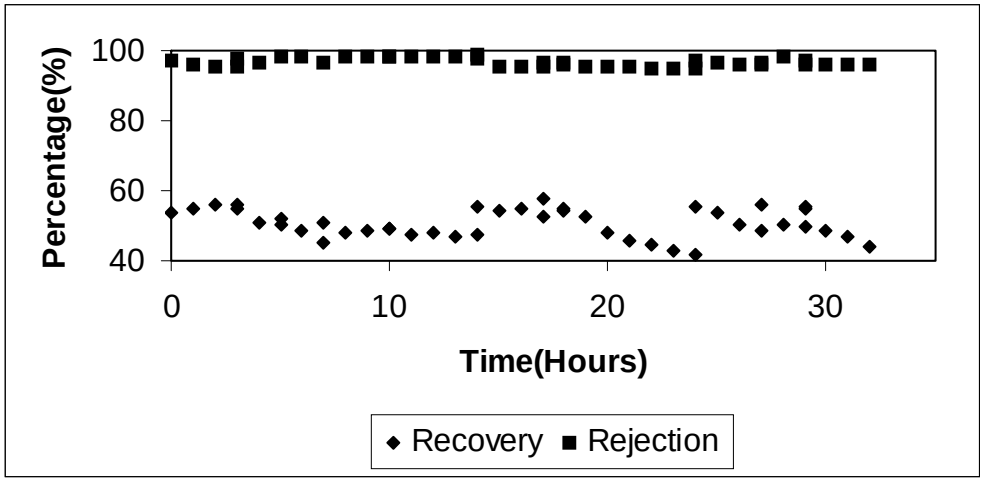


Figure 19.16 : Conductivity rejection and water recovery as a function of time.

The pH of the RO feed, permeate and brine is shown in Figure 19.17. The pH of the feed varied between approximately 7,3 and 8,0 (unadjusted) while the pH of the permeate varied between approximately 6,5 and 8,5. The pH of the permeate was, however, lower than the pH of the feed for most of the time due to the removal of alkalinity from the feed. Brine pH varied between approximately 7,5 and 8,0.

The pressure across the cartridge filters and the sandfilter is shown in Figure 19.18. The pressure at the sand filter outlet varied between approximately 75 and 85 kPa. No pressure gauge was present at the sandfilter inlet. The pressure drop across the cartridge filters varied between approximately 10 and 50 kPa. A cartridge filter was replaced after 7 hours of operation. The filter was darkly coloured. However, it was not really necessary to replace the filter because the pressure drop had not exceeded 70 kPa.

The chemical composition of the RO feed, product and brine is shown in Tables 19.14 and 19.15. An excellent quality permeate is produced. The TDS of the permeate was only 45 mg/ℓ (0,49 mS/m) (99,1% TDS removal, 95,7% conductivity removal). An almost 100 percent removal of organics was obtained (Table 19.14).

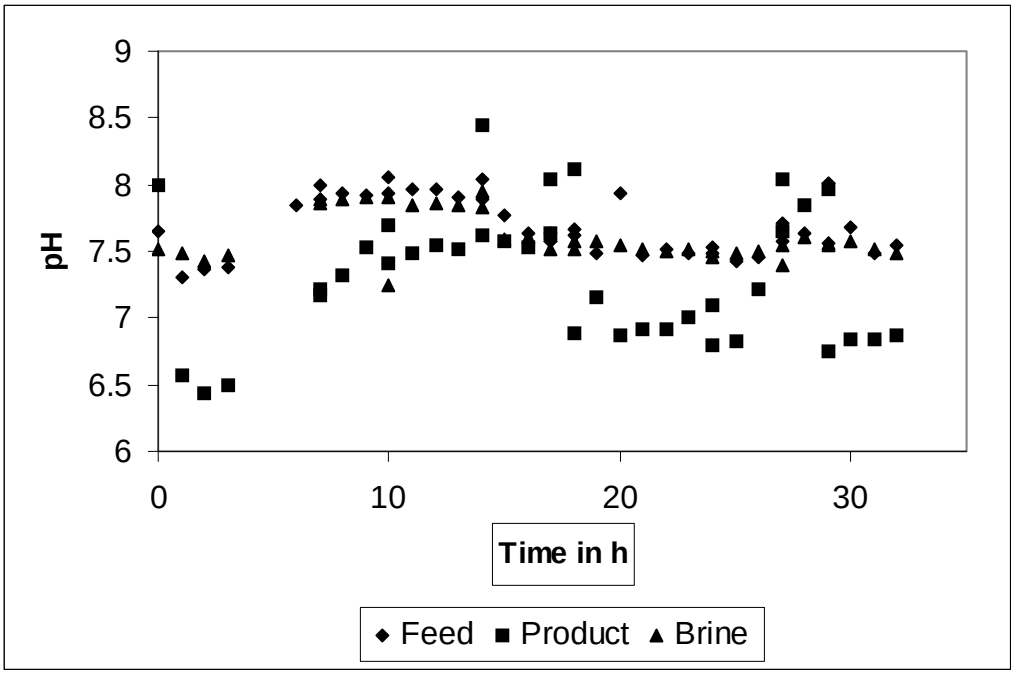


Figure 19.17: pH of RO feed, permeate and brine as a function of time.

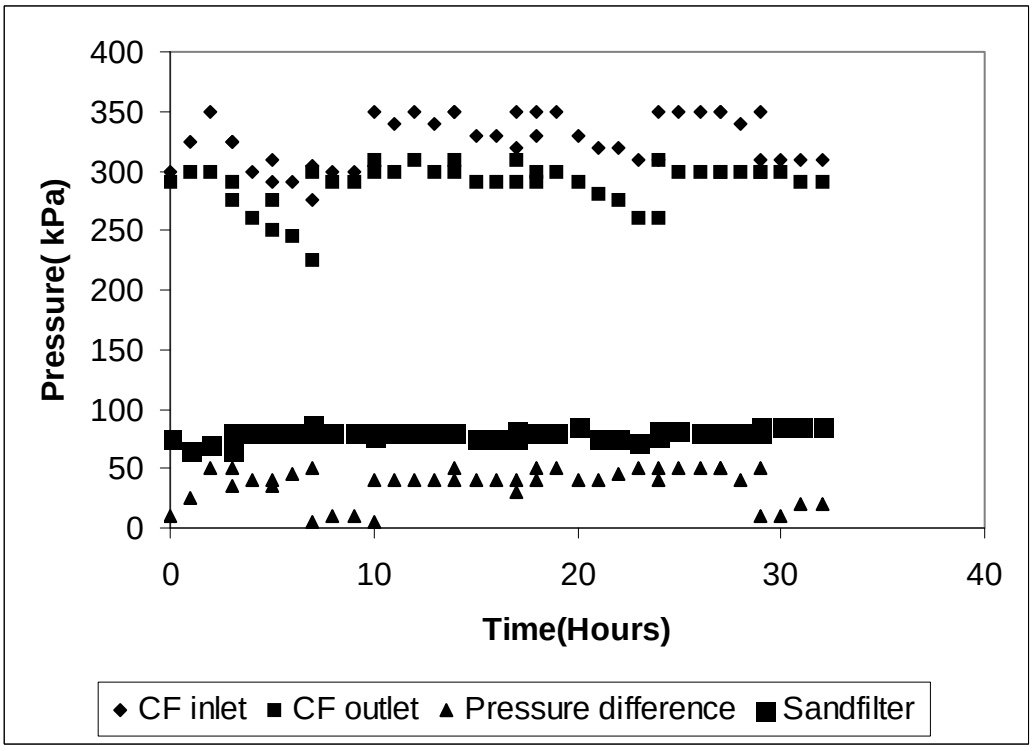


Figure 19.18 : Pressure across the filters as a function of time.

Ammonia-nitrogen was reduced from 589 to 46,6 mg/ℓ in the RO product (92,1% removal). Chloride was reduced from 1 535 to 13,6 mg/ℓ (99,1% removal). The quality of the RO permeate with the exception of ammonia-nitrogen is suitable for discharge to the water environment. The RO brine, however, should be further treated in an evaporator or discharged to landfill.

A similar good quality RO permeate was obtained after 24 hours of operation of the RO unit (Figure 19.5). The TDS was reduced from 3 528 to only 33 mg/ℓ (99,1% removal).

Table 19.14 : Chemical composition of feed, product and RO brine (after 3 hours of operation).

Constituents*	Feed	Product	Brine	Removal (%)
Total Dissolved solids	4 982	45	11 350	99,1
Total suspended solids	47	0	68	100
Ammonia nitrogen	589	46,6	965	92,1
Nitrate nitrogen	2,09	0	2,31	100,0
Total Phosphate	4,67	0,097	8,76	97,9
Alkalinity	3 754	194	7 611	94,8
Chloride	1 535	13,6	3 150	99,1
Sulphate	14,1	0	250,5	100,0
Fluoride	0,894	0	2	100,0
Silicon	22,2	0	49,6	100,0
Calcium	98,2	0	196	100,0
Magnesium	123	0,055	274	100,0
Potassium	639	10,4	1 210	98,4
Sodium	1 160	15,1	2 350	98,7
Barium	0,395	0	0,722	100,0
Strontium	0,976	0,001	1,89	99,9
Iron	3,41	0,003	4,711	99,9
Lead	0,104	0	0,178	100,0
Manganese	0,52	0,011	0,977	97,9
Nickel	0,141	0,012	0,285	91,5
Total Organic Carbon	532	4,02	1 260	99,2
Chemical Oxygen Demand	1 400	0	3 000	100,0
Phenols	0,41	0,02	0,31	95,1
Conductivity	11,3	0,49	22,7	95,7
pH	7,38	6,5	7,47	

*Concentration in mg/ℓ unless otherwise stated

Table 19.15 : Chemical composition of the RO feed product and brine (after 24 hours of operation).

Constituents*	RO Feed	RO Product	RO Brine	Removal %
Alkalinity	2 699	165	3 015	93,9
Ammonia nitrogen	647	48,7	978	92,5
Barium	0,224	0,057	0,4	74,6
Calcium	46,8	0	125	100,0
Chloride	1 277	24	1 860	98,1
COD	900	0	1 900	100,0
Fluoride	0,69	0	1,43	100,0
Iron	3	0	3,8	100,0
Lead	0,245	0,052	0,323	78,8
Magnesium	67,8	0,01	206	99,9
Manganese	0,535	0	1,07	100,0
Nickel	0,112	0,039	2,33	65,2
Nitrate nitrogen	8,88	0	8,8	100,0
Phenols	0,07	0	0,01	100,0
Phosphate (total)	14,0	0,11	32,9	99,2
Potassium	391	5,92	801	98,5
Silicon	11,8	0,006	18,2	99,9
Sodium	839	9,7	1 570	98,8
Strontium	0,557	0	1,14	100,0
Sulphate	54,98	<5,0	84,8	>90,0
TDS	3 528	33	8 105	99,1
TOC	32,4	5,2	46	84,0
TSS	22	0	39	100,0
Conductivity mS/m	6,91	0,3	17,1	95,7
pH	6,3	8,4	7,6	

*Concentration in mg/ℓ unless otherwise stated.

19.4 Economics

The capital and operational cost of a 250 kℓ/d RO plant (feed) to treat the MSWL at Bisasar Road are summarised in Table 19.16. The cost breakdown is shown in Appendix K.

Table 19.16 : Capital and operational costs to treat the MSWL at Bisasar Road.

Plant type	Cost	
	Capital (MR)	Operational (R/kℓ)
Tubular (cellulose acetate)	1,95	11,45
Tubular (polyamide)	8,1[(1)] 6,5[(2)]	16,24
Spiral	0,56	3,51
Disc tube RO	6,2	26,65

(1): Total plant from overseas supplier (2): Only membranes, modules, manifolds and frames from overseas supplier.

20. LIST OF HAZARDOUS WASTE SITES

A list of hazardous waste sites in South Africa is shown in Table 20.1. Effluent from these sites can be treated with RO, ED or evaporation technologies for pollution control and effluent volume reduction.

Table 20.1 : Hazardous waste sites in South Africa.

Head Office	
Holfontein (Springs/Delmas)	H:H
Sappi Enstra Mill	H:H
Bayside Aluminium SPL Pond	H:H
Mondi Merebank Ash Site	H:H
Tutuka Power Station, Eskom	H:H
Chloorkop (Kempton Park)	H:H
AEK CaF_2	H:H
AECI Umbogintwini	H:H
Umgeni Kragstasie	H:h
Aloes I (Port Elizabeth)	H:H
Koedoeskloof (Uitenhage)	H:H
Mossgas (George)	H:H
Margolis (Closed - Bedfordview)	H:H
Ballengeich (Closed)	H:H
Sappi Kraft Tugela Mill (Closed)	H:h
Kynoch Potchefstroom Gypsum Dam	H:h
Wasbank, Dundee Road Products	H:h
CISA Newcastle	H:h
Alusaf Bayside Storage Facility	H:h
Umlazi IV (Natal - Closed)	
Bulbul Drive	H:h
Mossop Western Leathers (Closure)	H:h
Shongweni	H:h
Aloes II	H:H
Frys Metal Deal Party (Closed)	H:H
Bloemfontein Medical Waste Incinerator: Storage Area	H:H
Visserhok CMC	H:H
Sappi Black Liquor Hazardous Lagoon	Haz Lagoon
Visserhok Uitspan (Waste-Tech-Cape Town)	H:H

Waste Disposal Sites with Permits - Highveld Region

TSB-Komati Mill (Nelspruit)	GMB+
Goudkoppies (Johannesburg)	GLB+
Nuffield	GLB+
Holfontein (Springs/Delmas)	H:H
Margolis	H:H
Tutuka Power Station, Eskom	H:H
Sappi Kraft Ngodwana Mill	GLB+
Driekoppiesdam Schoemansdal	GCB+

Waste Disposal Sites with Permits - Transvaal Region

Rosslyn	GMB+
Rayton	GSB+
Chloorkop (Kemptonpark)	H:H
AEK, CaF_2	H:H

Waste Disposal Sites with Permits - Eastern Cape Region

Alexandria Nature Reserve	GCB+
Tsitsikama National Park	GCB+
Aloes I (Port Elizabeth)	H:H
Koedoeskloof (Uitenhage)	H:H

Western Cape Region

De Rust Landgoed	GCB+
Citrusdal	GSB+
Bontebok Swellendam	GMB+
Ceres	GMB+
Darling (Malmesbury)	GMB+
Bellville Park	GLB+
Brackenfell	GMB+
Worcester	GMB+
Mossgas (George)	H:H
Visserhok Uitspan (Waste-Tech)	H:H

Natal Region

Highflats Village	GCB+
Banners Rest (Port Edward)	GMB+
Inanda	GMB+
Margate Oatlands	GMB+
Tongaat	GMB+

Free State Region

Bloemfontein Medical Waste Incinerator: Storage Area	H Storage

KwaZulu / Natal Region

Osizweni (closure)	GSB+
Hilton	GSB+
Madadeni (closure)	GSB+
Kwamgendwa	GSB+
Melmoth	GSB+
St. Lucia	GSB+
Borough of Eshow	GSB+
Hammarsdale	GMB+
Richards Bay TLC	GMB+
Ntuzuma	GMB+
Humberdale	GMB+
Empangeni	GMB+
Canoby Quarry	GMB+
Goswell Aluminium	GMB+
Mondi Alton - Richards Bay	GMB+
Kwadukuza	GLB+
Pinetown South	GLB+
Bisasar Road	GLB+
Cisa Newcastle	H:h
AECI Umbogintwini	H:H
Bulbul Drive	H:h
Wasbank (closed)	H:h
Ballengeich (closed)	H:H
Sappi Kraft Tugela Mill Mtunzini, closed	H:h
Alusaf Bayside SPL (storage facility)	H:h
Frys Metals Deal Party (closed, cont. area)	H:H
Umlazi IV (closed)	H:h
Mondi Merebank Ash Site	H:H
Bayside Aluminium SPL Pond	H:H
Shongweni	H:h

Gauteng Region

Sebokeng Zone 16	GMB+
Sasol Secunda Synthetic Fuels (with H:H portion)	GLB+
Verref-Rietfontein (Springs)	GLB+
Chloorkop (not operative)	H:H
Sappi Enstra Mill	H:H
Holfontein	H:H
Tutuka Power Station	H:H
Sappi Black Liquor Hazardous Lagoon	Lagoon
Margolis (closed)	H:H

Western Cape Region

Friemersheim	GCB+
Op die Berg	GCB+
Denel Somchem	GCB+
Kleinmond Transfer Station	GSB+
Plettenberg Bay	GSB+
Klapmuts (closure)	GSB+
Klapmuts Transfer Station	GSB+
Villiersdorp	GSB+
Wellington (New Site)	GSB+
Paarl	GSB+
Spoornet Knysna Transfer Station	GSB+
Tulbach	GSB+
Voorberg Prison	GSB+
Stanford	GSB+
Caledon	GSB+
Hermanus Transfer Station	GMB+
Karwyderskraal	GMB+
Stellenbosch Municipality	GMB+
Wellington	GMB+
Coastal Park	GLB+
Swartklip Transfer Station	GLB+
Saldanha Staal	GLB+

H : Highly hazardous
h : Less hazardous
G : General
C : Communual
M : Medium
S : Small
L : Large
+ : Positive water balance
- : Negative water balance

21. ODOUR CONTROL

21.1 Introduction to landfill odour control

Control of odour from landfills is becoming a major concern of landfill operators all around the world. Landfill gas (LFG) forms as a natural by-product of the degradation of waste and contains trace amounts of up to a hundred malodorous compounds. There are many difficulties facing landfill operators regarding the control of odour from landfills. The first relates to the fact that the malodours comments of landfills gas are found in minute concentrations. Landfill gas typically composes of 50 - 55% methane, 40 - 45% oxygen and 0 - 5% nitrogen and less than 1% trace components. Due to the low threshold values of some odours, compounds may only need to be present in low parts per billion to cause a problem.

Sources of odour at landfill sites can vary considerably from site to site. Main contributors include, covered areas, tipping/working face, gas collection / transportation /extraction systems, leachate collection / transportation / treatment systems and truck washing bays (for a more comprehensive list of possible sources see McGingley and McGinley, 1998). Many of these possible sources can be eliminated by sound operational procedures. For instance, at truck washing bays, odour problems can be minimised by good drainage and clearing of waste washed off the trucks. Covering and sealing completed sections of the landfill as well as installing a gas extraction system can minimise emissions from covered surfaces. Roberts and Sellwood (1997) suggest the potential for odour problems is greater with wide variety of odorous compounds produced during anaerobic conditions. While this may be true, anaerobic conditions occur in covered sections of the landfill where control measures (discussed above) can be implemented to minimise the effects of these emissions. Odorous emissions from leachate collections systems can be minimised by sound engineering techniques. The most difficult part of the landfilling process in which to control odours is at the working face where fresh waste is off loaded, moved and compacted before being covered. Working faces are typically only covered at the end of each day providing a full working day in which fresh waste odour can enter freely into the atmosphere. Cover used is also often inadequate in terms of the material and depth.

Odorous emissions from landfills cannot be stopped completely from entering the atmosphere. Because of this, control measures to counteract odour are necessary. There are two main considerations when selecting odour control measures. The first the counteractant to be used and the second is the application of the counteractant. Both are equally important. The correct odour counteractant combined with the incorrect application (and vice versa) can result in the control system having limited success.

21.2 Counteractants

Various methods of counteracting odour exist including: masking, absorption, reaction, combination, interference and bacterial manipulation (see Federici, 1998 for more details). In the case of landfills, odour reaction (chemical reaction of airborne odour compounds with odour neutralising compounds forming non-odorous compounds) is the most feasible method other than simply masking the odour.

21.3 Application of Counteractants

If the source of odour on a landfill is uncertain, the best option is to construct fence-line spray units, which spray odour modifying substances into the atmosphere. If the source of odour is identifiable other options include : -

1. Application of counteractant directly to odourous sources such as the working face or leachate chamber.

2. Spraying counteractant into the air near the source via tanks attached to either a pressure pump on the back of a bakkie or more permanently from landfill operating vehicles.

3. Spraying counteractant from a water tanker primarily used for watering roadways to control dust.

22. GUIDELINES FOR LEACHATE CONTROL

22.1 Introduction

Guidelines for leachate control have been compiled by DWAF (Le Roux, 2002). Guidelines for leachate treatment derived from this study will also be presented.

The term of 'leachate' refers to liquids that migrate from the waste carrying dissolved or suspended contaminants. Leachate results from precipitation entering the landfill and from moisture that exists in the waste when it is disposed. Contaminants in the buried refuse may result from the disposal of industrial waste, ash, waste treatment sludge, household hazardous waste, or from normal decomposition. If uncontrolled, landfill leachate can be responsible for contaminating ground water and surface water.

The Department of Water Affairs and Forestry has developed a second edition (1998) of the Waste Management Series (known as the 'Minimum Requirements'), which consist of three sets of documents : -

1. Minimum requirements for the handling, classification and disposal of hazardous waste;
2. Minimum requirements for waste disposal by landfill; and
3. Minimum requirements for water monitoring at waste management facilities.

These Minimum Requirements must be followed and adhered to when applying for a permit from DWAF for a waste disposal facility in terms of section 20 of the Environmental Conservation Act, 1989 (Act 73 of 1989).

The composition of leachate varies greatly from site to site, and can vary within a particular site. Some of the factors affecting composition include : -

- Age of landfill
- Type of waste
- Degree of decomposition that has taken place; and
- Physical modification of the waste (e.g. shredding).

Once ground water is contaminated, it is very costly to clean up. Today's landfills, therefore, undergo rigorous siting, design and construction procedures that provide many safeguards for the control of leachate migration.

A designed lining system, which ensures low-permeability limits the movement of leachate into ground water. Liners are made from low-permeability soils (typical clays) or synthetic materials (e.g. plastic). Landfills can be designed with more than one liner, and a mix of liner types may be used. The Minimum Requirements for Waste Disposal by Landfill from the Department of Water Affairs and Forestry gives detail information on landfill liner designs for the different classes of landfill sites.

Leachate collection systems are installed above the liner and usually consist of a piping system sloped to drain to a central collection point where a pump is located.

Once the leachate has been collected and removed from the landfill, it must undergo some type of treatment and disposal. The most common methods of managing leachate are :

- Discharge to sewage treatment works
- On-site treatment followed by discharge; and
- Recirculation back into the landfill.

22.2 Treatment in a Sewage Works

In some cases, landfill leachate can be added into an incoming wastewater stream at a sewage works, where it is biologically, physically, and/or chemically treated. In South Africa the routine treatment of leachate has tended to concentrate on biological treatment in order to reduce the organic components to acceptable levels. Biological treatment can be preceded by treatment of the inorganic constituents by physical or chemical treatment, in order to make the liquid more acceptable for biological processing, since the best overall treatment efficiencies can generally be achieved by removing the inorganic constituents first, and then removing the organic constituents. The approach protects the subsequent biological and other processes such as adsorption and air stripping from problems caused by metal toxicity, corrosion and scaling. A third stage treatment by using various physical and chemical treatment methods can be used after the biological process to clean up or 'polish' the leachate in order to remove recalcitrant organics or inorganics materials such as high salt levels that cannot be directly discharged into the environment.

- **Physical treatment processes**

  Physical treatment methods are used to remove, separate and concentrate hazardous elements and compounds, both organic and inorganic, from dilute and concentrated waste streams. Most physical treatment methods that have been applied to leachate treatment are conventional technologies and can remove a variety of problem contaminants. Increasingly, membrane technologies such as reverse osmosis and ultrafiltration are being applied. However, most membrane technologies suffer from problems associated with blockage of the membranes and landfill leachates with their relatively high CODs are often not good candidates for these technologies. Reverse osmosis, however, is successfully applied for the treatment of MSWL in Germany.

  Pre-treatment with physical technologies prior to biological treatment have been largely using sedimentation, coagulation and flocculation or filtration in order to remove suspended solids. After biological treatment, the presence of high concentrations of salts normally prevents direct discharge to the environment.

Options for treatment include evaporation or reverse osmosis with the recovery of a brine or solid salt material that often has to be disposed back into the landfill. Clearly, unless this process is managed carefully, it is essentially self-defeating, since the salt can re-enter the leachate and the treatment cycle has to be repeated.

- **Chemical treatment methods**

Chemical treatment methods have been widely used to treat leachate. This includes neutralisation, oxidation, precipitation and wet-air oxidation. Chemical pretreatment of leachate prior to biological treatment has included the addition of an alkali, usually lime, in order to raise the pH and to precipitate out heavy metals or, if the amount of calcium in leachate is a problem, soda ash is added to precipitate calcium carbonate.

Chemical oxidation has also been widely used in South Africa. Hydrogen peroxide is being used at most H sites in South Africa for the mitigation of odours produced by the leachate, since it readily reacts with any sulphides and mercaptan components that normally cause the odour. Hydrogen peroxide is expensive and large amounts would be required to have any significant impact on the concentration of organics in the leachate. In the UK, ozone has been used to oxidize recalcitrant organics such as humic acids, in order to break the molecules and make them more susceptible to biological treatment.

- **Biological treatment**

Biological treatment methods are processes whereby microbes are used to destroy or at least reduce the toxicity of a waste stream. Normally biological treatment of predominantly aqueous wastes such as leachate is accomplished in specially designed bioreactors. A suitable culture of the microorganisms or microbial association, either aerobic or anaerobic, is chosen.

Biological treatment is firmly established as the standard method of waste treatment for some wastes, i.e.

- Domestic sewage
- Waste from food processing
- Hazardous waste e.g. phenols, cyanide, oils
- Leachates

For leachates, a large number of approaches to biological treatment are proposed, but many are unproven and have not yet been shown to be effective on site.

The general type of transformations that can be accomplished biologically include : -

- Degradation of organics to products such as carbon dioxide, methane and water
- Reduction of inorganic compounds, e.g. nitrate
- Complexation of heavy metals, e.g. nickel.

Discharge to a sewage works, however, is not an option in all cases. Care must be taken not to interfere with operations at the sewage works. The contaminants in leachate can sometimes upset sewage works operations.

Typical leachate can often exceed the required discharge limits particularly in terms of COD and salt content. The discharge of heavy metals into the sewer system is, normally strictly controlled. Those metals of concern to the water authority include Fe, Zn, Cd, As and Hg. The last three are normally present in extremely low amounts in leachates, particularly from domestic sites, although high amounts of Fe are often found.

22.3 On-site Treatment

When discharge to a sewage system is not feasible, constructing treatment facilities on-site with the sole purpose of treating leachate may be necessary. The Aloes Class H:H disposal facility in the Eastern Cape is an example of this. These facilities will add to the cost of a new facility, but may be required to meet environmental standards.

On site treatment reduces high concentrations of COD and BOD. Retention times from 10 to 50 days can result in the removal of 90% of COD and ammonia. Nitrification of high concentrations of ammonia can be achieved by extended aeration and at increased temperatures. The addition of phosphoric acid may be required for microbial growth and inputs of sodium hydroxide for pH adjustment. The operating parameters vary, depending on the quality and nature of the leachate and extended trials are required to determine these for a specific leachate. Aerobic treatment results in a reasonable reduction in COD and ammonia and can be accomplished at quite high conductivity and chloride levels. However, the resulting effluents will still have a relatively high COD and high conductivity, which is mainly related to chloride levels. Polishing of the leachate has included the use of artificial reed beds and ozone treatment prior to discharge to a water course.

These methods have been applied widely to the on-site treatment of leachate from domestic waste sites, although waste sites that have accepted limited hazardous waste have also been successfully treated.

It has recently been demonstrated at Holfontein that evaporation/crystallisation technology can be successfully applied for the removal of organics and inorganics from the hazardous leachate. It was also demonstrated that electrodialysis can be successfully applied for the partial desalination of the Holfontein hazardous leachate. Further treatment of the leachate with RO should be able to produce treated water that could be discharged into the water

environment. The brine can be treated in a drier/evaporator. It is also important to note to not to generalize when it comes to MSW and ISW leachates. Leachate quality is often site specific depending on waste streams that have been accepted on site.

Odours should be successfully controlled with aspirating aerators or with the archaea biocatalyst in conjunction with the aerators (Alcock 2002). The aerators do not throw effluent into the air and aerate in that way that they have propellers which pump air down into the waste liquid so that aerosols are not formed.

The following general guidelines derived from literature and experimental work for leachate treatment are also suggested : -

(a) Both MSW and ISW leachate should be treated to comply to the quality requirements laid down for discharge to the water environment and/or to the sewer.

(b) Municipal solid waste leachate should be treated in a SBR unit where only organics is needed to be removed (no TDS removal required).

(c) Municipal solid waste leachate should be treated with RO where only TDS is needed to be removed.

(d) Municipal solid waste leachate should be treated biologically (SBR) as well as with RO where both organics and TDS are needed to be removed.

(e) The process combination for the treatment of small and medium-sized MSW leachate quantities is biological (UF membrane reactor) pretreatment/reverse osmosis/fluidized bed granulator.

(f) Industrial solid waste leachate should be treated with ED and RO where both organics and TDS are needed to be removed.

(g) Industrial solid waste leachate (TDS > 50 000 mg/ℓ) should be treated in an evaporator/crystalliser followed by biological or activated carbon treatment where both organics and TDS are needed to be removed.

(h) Electrodialysis should be more a suitable technology for the treatment of lower TDS (10 000 to 50 000 TDS) leachates than evaporation/crystallisation technology.

(i) Brine from ED and RO plants should be treated in an evaporator/dryer to produce a solid product which is easier to dispose of than a liquid effluent.

(j) Odour control is a well established technology and should be applied at landfill sites.

22.4 Recirculation

Recirculation is another management technique for leachate treatment. When leachate is recirculated through the waste pile, the decomposition process in the landfill speeds up, resulting in a shorter time for the landfill to stabilise. The technique, however, does not eliminate the leachate. Ultimately, the leachate will have to be treated by one of the other methods. Especially in cases where too much leachate is produced for storage thereof in evaporation ponds.

The objectives of recirculation are to : -

- Encourage early establishment and methanogenesis that is promoted by a high moisture content and the movement of moisture. It also promotes degradation of hazardous organic and other biodegradable waste.
- Develop a more uniform quality of leachate in order to facilitate an easier operation of the landfill site.
- Encourage leaching of solid industrial wastes.
- Minimise dry zones in the waste that would remain undergraded for many years.
- Take up absorptive capacity and reduce fluctuations in leachate flow rate.
- Promote enhanced evaporative losses by surface spraying.
- Provide temporary storage of short-lived peak flow rates.

The main advantage of recirculation of leachate for a landfill operator is the increase in the waste stabilisation rate and the considerable decrease in leachate volume that can be obtained.

Many disadvantages are : -

- Surface flooding
- Spraydrift that can cause pollution and health problems
- Malodours
- Clogging of subsurface recirculation systems; and
- Without pretreatment, undesirable inorganic contaminants that can build up (e.g. chloride and ammonia).

23. GENERAL DISCUSSION

It was demonstrated in this study that electrodialysis is a new treatment option that could be considered for the partial desalination of a very difficult-to-treat industrial solid waste leachate of the Holfontein type. The quality of the desalinated leachate should be suitable for discharge into the sewer system, while the ED brine should be further treated in an evaporator/dryer to increase water recovery and to decrease brine volume. Brine volume was determined at approximately 40% of the treated flow, while water recovery of more than 90% should be possible in a 5-stage ED unit. Brine volume should be further reduced by adding less water to the brine. This matter, however, needs to be investigated further, as well as the overall economics of the process.

A treatment option that should be able to produce a treated leachate quality of almost potable water quality from an industrial solid waste leachate of the Holfontein type, is partial desalination of the leachate with ED, followed by further desalination of the ED product with RO. The ED and RO brines can be treated in an evaporator/crystalliser to reduce brine volume. The economics of such a process should be determined.

High concentrations of organics in the desalinated ED product may be a problem when desalting an industrial solid waste leachate of the Holfontein type. However, an interesting treatment option in this case could be to treat the desalinated ED product in a membrane bioreactor prior to discharge to a sewer. Nanofiltration and/or ultrafiltration membranes can be used in the membrane reactor to ensure biodegradation of recalcitrant organics.

The lifetime of ED membranes for the treatment of an industrial solid waste leachate of the Holfontein type is an unknown factor. No information is available in the literature or from suppliers of ED equipment regarding the lifetime of ED membranes for the treatment of such a type of leachate. This study has indicated that the membranes used for the treatment of the Holfontein leachate are reasonably stable towards membrane fouling over an extended period of time (approximately 900 hours). However, no long term performance results of the membranes on pilot scale are available. Therefore, the required information should be obtained through long term pilot studies.

The capital costs of an ED plant is very sensitive to the feed concentration and the current density used in the process. Significantly larger membrane areas are required at lower current densities than at higher current densities. This was encountered in the laboratory studies where much higher current densities (approximately 3 times higher) were used than in the pilot studies, due to a limitation in the pilot equipment. The capital cost estimates in the case of the ED pilot studies were consequently significantly higher. Therefore, the highest current density possible should be used in an ED application without affecting the membranes adversely. This optimum current density should be established in pilot trials as a function of time.

The treatment of the Holfontein leachate with an evaporator/crystalliser followed by biological and activated carbon treatment as suggested by Enviroserv appears to be very promising. Water that should be discharged directly into the water environment should be produced with this combination of technologies. Treatment costs, however, are not available yet. This technology should be more suitable for the treatment of the Holfontein type leachate with its very high TDS and organic concentration. Electrodialysis, however, should be a more suitable technology for lower TDS (20 000 to 50 000 mg/ℓ TDS) industrial solid waste leachates.

Preliminary studies outside the scope of this work have shown that an industrial solid waste leachate from Aloes in Port Elizabeth could be very effectively desalinated with ED. TDS was reduced from 42 185 mg/ℓ to 896 mg/ℓ and the desalinated ED product complied with the quality requirements for discharge to the sewer (chloride reduced from 20 987 mg/ℓ to 345 mg/ℓ). Consequently, the ED process appears to be a very effective process for the partial desalination of lower TDS industrial solid waste leachates.

A new process has been developed for the treatment of industrial solid waste leachates. The Aloes leachate is one example where this process can find application in due course for the desalination/concentration of the leachate. There are a number of other industrial solid waste leachates and other industrial brines which have the potential to be treated with ED. These possibilities should be exploited.

It should be possible to treat the MSWL from Bisasar Road effectively with RO for direct discharge into the water environment. It further appears that tubular RO should be the method of choice. This is a relatively straight forward simple technology to use which does not require any pretreatment for the removal of suspended solids from the leachate. The only pretreatment required is pH adjustment of the feed water and an anti-scalant to prevent the deposition of scalants on the membranes. It was also demonstrated that membrane fouling of both cellulose acetate and polyamide RO membranes could be controlled in pilot studies. However, it appeared that the performance of the polyamide membranes was superior to that of the cellulose acetate membranes for the desalination of the leachate.

Preliminary indications are that good results should be obtained with the desalination of the municipal solid waste leachate at Bisasar Road with spiral wrap RO membranes. The only pretreatment that was applied was sandfiltration followed by cartridge filtration (5 micron) prior to desalination. A small reduction in the permeate flux was experienced over a short test period (32 hours). This indicates membrane fouling. However, it appears that it should be possible to control membrane fouling with chemical cleaning. This should, be evaluated over an extended period of time, as well as to evaluate the effect of the magnetic field over the membrane module for fouling control.

No experience is available in South Africa regarding a combination of biological treatment and reverse osmosis desalination for the treatment of municipal solid waste leachates. It

might be necessary to use this combination for certain applications. Consequently, the necessary experience should be obtained through laboratory or pilot studies.

Reverse osmosis technology should be implemented for the treatment of municipal solid waste leachate in South Africa where required. Electrodialysis technology, however, should be pilot tested over a longer period of time before this technology should be implemented on a large-scale for the treatment of industrial solid waste leachates. This is considered to be necessary so that the optimum conditions can be defined for the successful application of the process.

Guidelines are presented for the treatment of municipal and industrial solid waste leachates. Both MSW and ISW leachates should be treated to comply with the quality requirements laid down for the discharge of effluent to the water environment and/or sewer. Municipal solid waste leachate should be treated in a SBR unit where only organics is needed to be removed (low TDS). However, municipal solid waste leachate should be treated with RO where only TDS is needed to be removed. Where both organics and TDS are needed to be removed from MSW leachate, biological treatment (SBR) followed by RO should be the method of choice. The process combination for the treatment of small and medium-sized MSW leachate quantities is biological (UF membrane reactor) pretreatment/reverse osmosis/fluidized bed granulator.

Industrial solid waste leachate with high TDS could be treated with ED and RO where both organics and TDS are needed to be removed. Industrial solid waste leachate with a very high TDS should be treated in an evaporator/crystalliser followed by biological or activated carbon treatment where both organics and TDS are needed to be removed. Electrodialysis, however, should be a more suitable technology for the treatment of lower TDS (10 000 to 50 000 mg/ℓ TDS) leachates than evaporation/crystallisation technology. Brine from ED and RO plants should be treated in an evaporator/dryer to produce a solid product which is easier to dispose of than a liquid effluent.

The capital and operational costs of ED and RO plants were determined to treat industrial and solid waste leachates. The capital and operational costs are high. The capital costs to treat 80 and 140 kℓ/day leachate of the Holfontein type are estimated at 15,5 and R23,3 million, respectively. Operational costs are estimated at R188,24/kℓ, respectively, for a one year membrane lifetime. Operational costs, however, should be significantly reduced with increasing membrane lifetime (R93,55/kℓ for a 3 year lifetime).

The capital and operational costs of RO plants to treat municipal solid waste leachate are also high. However, the capital costs of South African RO plant are significantly less than that of overseas plants due to the exchange rate. The capital cost to treat 250 kℓ/d MSW leachate with tubular cellulose acetate membranes is estimated at R1,95 million while a tubular polyamide plant from an overseas supplier would cost R8,1 million. Local engineering would reduce this cost to approximately R6,5 million which is still expensive.

Operational costs were estimated at R11,45 and 16,24/kℓ permeate (70% recovery) for the South African and overseas plants, respectively.

The capital costs of a South African spiral wrap RO plant to treat 250 kℓ/d of MSW leachate were estimated at R0,56 million. Operational costs were estimated at R3,51/kℓ permeate (55% recovery). These costs are significantly less than the costs for the tubular RO plants. Very little pretreatment (sand and cartridge filters) was conducted, therefore, the lower operational costs. Operational data were also only collected over a relatively short period (approximately 32 hours). Nevertheless, the capital and operational costs, appear to be attractive and longer term tests should be conducted to confirm the costs.

Disc tube RO is used with great success in Germany for the treatment of MSW leachate. Good results are obtained with this technology. This technology, however, is expensive compared to the South African standards.

24. SUMMARY AND CONCLUSIONS

24.1 Literature Overview

24.1.1 Industrial solid waste leachate (high COD and TDS)

(a) Various methods of leachate treatment may be combined in various modes with other standard chemical engineering unit processes, providing a huge range of variable options are available from which a solution can be engineered to optimise the balance between cost and quality. Some examples of combinations of these processes that have been shown to be successful in previous studies for the treatment of high COD leachates, are as follows : -

(i) Chemical oxidation - biological oxidation - chemical precipitation
(ii) Biological oxidation - chemical oxidation - biological oxidation - chemical precipitation
(iii) Chemical precipitation - chemical oxidation - biological oxidation
(iv) Chemical precipitation - biological oxidation - chemical oxidation - biological oxidation.

In researching leachate treatment requirements for the Holfontein leachate in some detail, it has become apparent that the solution may be found not only in leachate treatment technology, but in chemical process technology as well. Alternatives include : -

(i) The phenosolvan processes for the extraction of phenols from water and other process streams,
(ii) Activated carbon or related chemical dosing technology.

(b) A number of laboratory and other small-scale tests were carried out in 1999 in South Africa to test the application of various treatment options on raw leachate from Holfontein. The general conclusions from the early test work were:

(i) Evaporational/crystallisation is the best technology for the first step in the process. It does not, however, on its own, produce water of the required quality. Further treatment of the condensate produced by the evaporator/crystalliser is required.

(ii) Almost no biological activity occurs in the raw leachate due to the toxic nature of the leachate. To obtain significant biological activity requires substantial dilution of the feed.

(iii) Inadequate reduction in COD and dissolved salts were obtained with electrodialysis treatment of the leachate. membrane fouling was also experienced. However, ED with fouling resistant membranes holds promise for the partial desalination of the leachate.

Pilot studies were subsequently conducted to evaluate COD removal with an evaporator/crystalliser. This showed that with leachate containing 80 000 mg/ℓ COD, the condensate would contain about 14 000 mg/ℓ COD. Dissolved salts in the leachate of 100 000 mg/ℓ were reduced to about 300 mg/ℓ. By raising the pH to 10,5, the COD in the condensate could be reduced to about 6 000 mg/ℓ COD. At a pH of 12,5 the COD was reduced to 1 300 mg/ℓ.

No COD reduction was found with an anaerobic plant on the condensate of the evaporator/crystalliser. An aerobic plant, in contrast, reduced the COD content of the condensate from 3 000 mg/ℓ to about 400 mg/ℓ.

Activated carbon removed the COD from the condensate to about the required level of 65 mg/ℓ for discharge to the water environment.

Ammonia was not removed adequately from the condensate (1 000 mg/ℓ) in either activated carbon columns or in an aerobic biological treatment plant. A number of treatment options to reduce the ammonia are being tested. Treatment options include :

(i) Distillation of the final condensate after activated carbon treatment;
(ii) Air stripping of the condensate prior to biological treatment.

At this stage of the development of the project, both processing options are still being considered. A final decision will be made once a more detailed estimate has been obtained of the capital and operating costs and the risks and uncertainties are better defined.

The treatment options are :

(i) evaporator/crystalliser plus activated carbon; and
(ii) evaporator/crystalliser plus aerobic biological treatment plus activated carbon.

24.1.2 Municipal solid waste leachate (low COD and TDS)

(a) Reverse osmosis plants in operation have proven the suitability of the process for the treatment of landfill leachate and similar waste waters. This, however, does not only apply to the unsurpassed discharge qualities, but also to the high availability of the installations (in general >90%).

Reverse osmosis plants may be operated with or without biological pretreatment. In general a multi-staged RO is required if a biological pretreatment is not installed. With biological pretreatment, a 1-stage RO is in most cases sufficient.

The ideal process combination for the treatment of small and medium-sized leachate quantities is biological pretreatment and removal of solids by ultrafiltration. Sludge separation by ultrafiltration allows the obtaining of a high content of solids in the biological reactor. This makes the installation of small reactor volumes possible. The process combination for the treatment of small and medium-sized leachate quantities is biological pretreatment/reverse osmosis/fluidized bed granulator.

With the new drying technique of evaporation and fluidized bed granulation a solution has finally been found to the much discussed issue of how to properly treat RO concentrate.

(b) Results obtained during the operation of an increasing number of plants under very different conditions prove that RO is a very effective instrument for the purification of landfill leachate if all design criteria and requirements specific for landfill leachate have been taken into consideration, and if an adapted module system as well as correlated technologies are used. This includes high pressure RO, with operating pressure up to 120 bar and/or NF in combination with a controlled crystallisation process, that allows permeate recovery rates of more than 95%.

The elimination of the negative impact of landfill leachate on the environment can be achieved with membrane filtration due to the dramatic minimization of residual waste to be processed or immobilized and due to the high quality of the purified water discharged back to nature. The combination of processes designed for this purpose is one example of a sustainable environmentally friendly development.

(c) The treatment of highly polluted water by RO is a reliable and economic operation and can be considered as state of the art. The process can produce water of any required quality - if necessary in a cascaded operation.

A major problem of waste water treatment is the water recovery rate, which should be near to 100%, realized in a simple, and an energy-efficient process combination. As demonstrated by long-term experiments in pilot plants and on technical scale, this can be achieved by the addition of NF and 200 bar high pressure RO to the 60/120 bar RO stage. The integration of the simple mechanical unit operation crystalliser/hydrocyclone/filtration promises an almost zero discharge process.

(d) In RO/NF, research and development concentrate on shifting the limits of processes to very high water recoveries, i.e. the development of ‘almost zero discharge’ processes. This is also strongly related to module development.

(e) In the course of the treatment of landfill leachates by reverse osmosis (single and multi-level) a concentrate of 15 to 25% of the amount of the raw leachate accumulates, the difference being determined by the chosen concentration factor. In principle, there are three possibilities for the disposal of concentrate :

(i) Return to the landfill;
(ii) Evaporation and drying-out;
(iii) Incineration (for example, in a waste incineration plant).

(f) A leachate treatment pilot plant has been established and commissioned at the Bisasar Road Landfill Site in Durban to access the treatability of the landfill leachate. Complete biological nitrification and denitrification of landfill leachate has been achieved inside the same SBR unit, and the viability of establishing full-scale treatment plants at landfill sites has been assessed. Levels of some 2 000 mg/ℓ of ammonia-nitrogen has shown to be completely and consistently nitrified to nitrate-N, and in turn released as harmless nitrogen gas, resulting in only negligible levels of nitrogenous components in the treated effluent. Sludge build-ups are minimal, and sludge waste has proven to be an infrequent occurrence. Future and ongoing research work involves the assessment of denitrification processes utilising waste molasses, and the feasibility of further polishing treatment aimed in particular at the removal of residual COD levels, using constructed wetlands.

24.2 Characterisation of the Industrial Solid Waste Leachate

The ISWL from Holfontein contains very high inorganic (TDS approximately 105 600 mg/ℓ) and organic (COD approximately 64 000 mg/ℓ) concentrations. The inorganic cations consists mainly out of sodium (67% of cations), ammonium-nitrogen (12,5%) and potassium (11%) while the major anions are chloride (56,4% of anions), sulphate (31,4%) and bicarbonate (12,3%). Significant quantities of magnesium (5,6%) and calcium (3,9%) ions are also present. Other hazardous compounds include phenols, lead, chromium, arsenic, nickel, etc. It is interesting to note that the phenol concentration of the leachate is high (900 mg/ℓ). The BOD/COD ratio is low (0,26). Therefore, it will be difficult to biodegrade the leachate. The relatively high volatile fatty acid concentration (approximately 6 600 mg/ℓ) shows, that some of the organics will be readily biodegradable. Biodegradability of the leachate, however, could be inhibited by toxic compounds in the leachate.

24.3 Biodegradability of the Industrial Solid Waste Leachate

Biodegradable tests (respirometer) have shown that it should be possible to biodegrade the ISWL from Holfontein to some extent. COD removals, however, were low (7,3 to 8%). Significantly higher COD removals (30%) were obtained on the ED treated leachate because most of the salinity had been removed from the leachate. Dilution (5x) of the leachate improved COD removal to 34,9%. The addition of biosupplements to the leachate also improved its biodegrability somewhat. COD removal was improved from 5,4% (no biosupplement) to 7,8 to 8% with the addition of biosupplements. Bio-enhanced treatment of the leachate in an SBR unit has shown that approximately 20 to 45% of the COD could be removed.

24.4 Evaluation of Ash Pretreatment of the Industrial Solid Waste Leachate for Organics and Inorganics Removal

Very good phenol removals (93,6%) were obtained when the ISWL from Holfontein was treated with Iscor ash (200 g/ℓ). COD removal, however, was not that good (14,5%). COD removals increased with increasing ash dosage and similar COD removals were obtained with Lethabo and Sasol ash samples. The ash added high concentrations of calcium to the leachate. No significant removals of magnesium, manganese, barium, strontium and silica were obtained at relatively low ash dosages. However, significant quantities of magnesium, manganese, ammonia and iron were removed at high ash dosage in the case of Sasol ash. Sludge volume comprises approximately 27% of the treated volume in the case of Iscor ash (200 g/ℓ). Iscor ash appears to be the best candidate for treatment of the leachate because it can remove significant quantities of phenol from the leachate.

24.5 Lime, Caustic Soda and Soda-Ash Pretreatment of the Industrial Solid Waste Leachate for Organics and Inorganics Removal

Excellent magnesium (97,1%), manganese (98,5%), barium (90,6%) and iron (86,2%) removals were obtained with caustic soda treatment of the ISWL from Holfontein at high pH (pH 12). Strontium removal was also obtained (60,8%). Significant amounts of COD (46,5%) and phenols (57,8%) could also be removed with caustic soda treatment. However, a high concentration of the sodium was added to the leachate while almost no calcium was removed. High quantities of manganese (99,2%), barium (90,6%) and iron (82,9%) could be removed from the leachate with lime treatment at high pH. However, high concentrations of calcium were added to the leachate. A significant amount of COD (42,5%) could also be removed with lime treatment, while very little phenols were removed. High concentrations of calcium (83,3%) could be removed from the leachate with soda ash treatment. Significant amounts of manganese (86,3%) and barium (75%) could also be removed from the leachate with soda ash and lime treatment. Caustic soda and soda ash treatment of the leachate showed that excellent removals of calcium (83,1%), magnesium (71,2%) and barium (68,8%) could be achieved. However, sodium was added to the leachate while some phenol removal was obtained. Excellent removals of calcium (86,8%), manganese (92,1%) and barium (68,8%) were obtained with treatment of the leachate with Iscor ash, caustic soda and soda ash (200 g/ℓ ash, 12 g/ℓ Na_2CO_3, 6,9 g/ℓ NaOH). Strontium and iron removals, however, were not that good. A significant amount of sodium was also introduced into the leachate. Poor COD removals were also obtained. However, the removal of phenol (92,7% removal) which can attack membranes, was good. Therefore, the abovementioned combination of ash and chemicals were selected for the treatment of the leachate prior to desalination. Chemical treatment costs were determined at R20,40/kℓ for soda ash and R22,08/kℓ for caustic soda. (Note: the chemical dosages are not necessarily the optimum dosages for the desalination of the leachate).

24.6 Coagulation/Flocculation Pretreatment of the Industrial Solid Waste Leachate for the Removal of Suspended Material and Organics

Good suspended solids removals (63 to 82%) could be obtained with treatment of the ISWL from Holfontein with alum, ferric chloride, poly-aluminium chloride and a polyelectrolyte. Good suspended solids removal could also be obtained with filtration (74%) of the leachate. Lowering of the pH also helped to increase the removal of suspended solids (86,3%) in the case of ferric chloride treatment. Organics removals, however, were poor.

24.7 Nanofiltration for the Treatment of the Industrial Solid Waste Leachate

Good organic removals were obtained when the ISWL from Holfontein was treated with nanofiltration (approximately 42 to 59%) (AFC 40, MPT31 and MPT 36 membranes). However, a significant amount of organics permeate the membranes (24 to 37%) showing that there are low molecular mass organics present in the leachate. The monovalent ions (Na^+, Cl^-) permeate the membranes preferentially with the result that a good separation could be obtained between the monovalent and divalent ions (Ca^{2+}, SO_4^{2-}). Approximately 52,5% chloride ions, for example, were present in the permeate against 18% divalent sulphate ions (AFC 40 membranes). Also, significantly more monovalent sodium ions were present in the permeate (70%) than divalent calcium (14,6%) and magnesium (15%) ions. More monovalent chloride ions (56,2%) were present in the permeate than sulphate ions (10,3%) (MPT 31 membranes). The same applied to the monovalent sodium (44%) and calcium ions (31,3%). The chloride ions in the permeate in the case of the MPT 36 membranes were 58,1% while only 25,1% sulphate ions were present in the permeate. The same applies to the monovalent sodium (48%), potassium (42,2%) and the divalent calcium (30,9%) and magnesium ions (33%). Therefore, the nanofiltration permeate should be more suitable for further desalination because most of the scale-forming chemicals have been eliminated. The membranes, however, were fouled by the leachate. Preliminary tests, however, have indicated that it should be possible to clean the membranes with chemicals. Nanofiltration has potential for the treatment of leachate and further investigations should be done in this regard.

24.8 Fouling Potential of the Industrial Solid Waste Leachate for Electrodialysis Membranes

The fouling potential of the ISWL from Holfontein for several ion-exchange membranes was evaluated in a membrane fouling test cell. It was found that most of the anion membranes (Selemion AMV, Tokuyama Soda ACS, Selemion ASV) were fouled by the leachate. However, the Tokuyama Soda AFN anionic membrane, Ionics AR204SZ RA anionic membrane and the Tokuyama Soda AXE anionic membrane were far less fouled by the leachate. The AFN anionic membrane appeared to be more resistant to fouling by the leachate as indicated by the potential drop across the membrane during the fouling studies and by membrane resistance measurements. The AFN anionic membrane was therefore selected for further ED studies.

24.9 Membrane Cleaning Strategies of the Fouled Electrodialysis Membranes with Commercially Available Membrane Cleaning Agents

Long-term membrane fouling studies were conducted in the membrane fouling test cell (approximately 900 hour) to further evaluate the fouling potential of the leachate for the membranes and to evaluate cleaning of the fouled membrane. The study has indicated that the AFN anionic membrane can be fouled by the pretreated ISWL from Holfontein. However, it was demonstrated that it should be possible to clean the fouled membranes with mechanical cleaning, polarity reversal and cleaning of the membrane with salt solution (3%) at high pH (pH 11,5).

24.10 Electrodialysis Treatment for the Desalination/Concentration of the Pretreated Industrial Solid Waste Leachate

Electrodialysis treatment of the pretreated ISWL from Holfontein in a laboratory scale ED unit (AFN and CMX membrane) has shown that the TDS of the leachate could be reduced from 118 485 mg/ℓ to 17 236 mg/ℓ (85,5% TDS removal). Therefore, an excellent removal of TDS could be obtained with ED treatment of the leachate. Approximately 49% of the COD could also be removed from the leachate. Water recovery was 62,5%. This implies that the brine volume comprises approximately 38% of the treated leachate. The salt concentration of the brine is high (215 980 mg/ℓ) and the brine should be further treated for safe disposal. Analysis indicated that there was some reduction in the ion-exchange capacity and gel water content of the anion-exchange membranes after ED treatment. This indicates some degree of membrane fouling. However, the extent of the membrane fouling does not appear to be too serious. It was also shown that the desalinated stream (ED product) was far less toxic than the untreated and pretreated leachate and ED brine. A 16 time dilution of the ED product will be required to avoid acute effects on the aquatic life if the ED product is discharged into the water environment. An 800 times dilution will be required in the case of the ED brine and less than a 159 times dilution on the pretreated leachate.

24.11 Preliminary Economics of the Electrodialysis Process

The capital cost of an 80 kℓ/d ED plant is estimated at R5,6 million (approximately R8,3 million for a 140 kℓ/d plant). Operation costs including electrical energy for ion transport, membrane replacement costs (1 year life time), chemical pretreatment and pumping costs are estimated at R103,41/kℓ. Operational cost is estimated at R68,31/kℓ for a 3-year membrane lifetime. The costs of membrane cleaning chemicals are not included in the operational costs. (Note: costs derived from laboratory-scale tests).

24.12 Performance of ED on Pilot-Scale for the Treatment of Industrial Solid Waste Leachate

Treatment of the pretreated leachate from Holfontein in an ED pilot plant (batch treatment) has shown that the TDS of the leachate could be reduced from 104 260 mg/ℓ to 11 903 mg/ℓ (88,6% TDS removal). COD removal was 48,6% and water recovery was 47,5%. This implies that the brine will comprise approximately 50% of the treated leachate. The TDS of the brine is high (169 805 mg/ℓ), and the brine should be further treated in an evaporator prior to final disposal. This will ensure that a high water recovery and low brine volume can be obtained. It was also shown that the ED product (19 350 mg/ℓ) could be further desalinated to 645 mg/ℓ with RO (96,7% TDS removal). COD was removed from 25 500 to 935 mg/ℓ (96,3% removal). Therefore, a very good quality water could be produced with combination of ED and RO treatment of the leachate. The RO brine which should comprise approximately 20% of the RO feed should also be treated in an evaporator, to increase water recovery and to reduce the brine volume. It also appears that it should be possible to control fouling of the RO membranes with chemical cleaning.

Feed-and-bleed ED pilot tests have shown that the TDS of the leachate could be reduced from 116 235 mg/ℓ to 2 435 mg/ℓ (5 stage ED). Brine volume comprised approximately 41% of the treated leachate. A water recovery of more than 90% should be possible. It also appears that membrane fouling should not be a serious problem.

The capital cost of an 80 kℓ/d (feed) ED plant is estimated at R15,5 million (approximately R23,3 million for a 140 kℓ/d (feed) plant). Operational costs including electrical energy for ion transport, membrane replacement costs (1 year life time), chemical pretreatment and pumping costs are estimated at R188,24/kℓ. Operating costs will be R93,55/kℓ for a 3-year membrane lifetime. The cost of cleaning chemicals are no included in the operational costs.

Significant quantities of magnesium, potassium, aluminium, iron, manganese and titanium can leach from the sludge produced after chemical treatment (ash plus chemicals) of the leachate prior to ED desalination. However, the concentration of these elements in a leachate is not considered to be dangerous when the sludge is disposed of in a landfill.

Organics at low concentrations can also leach from the sludge. These organics are also not considered to be a problem where the sludge is disposed of in a landfill.

24.13 Characterisation of the Municipal Solid Waste Leachate

The electrical conductivities of the Bisasar Road leachate were determined as 1 291 and 1 650 mS/m in two cases. The TDS were determined as 7 070 mg/ℓ in one case. The COD of the Bisasar Road leachate were 2 427 and 2 000 mg/ℓ in two cases. Both the conductivity and the COD of the leachate do not comply to the requirements of the general discharge standard. The BOD/COD ratio was only 0,13 in one case and 0,48 in another case. The low

ratio suggests that it should be difficult to biodegrade the organics in the leachate, while the higher ratio suggests that the leachate should be readily biodegradable.

The ammonia-nitrogen concentrations of the Bisasar Road leachate are high (1 271 and 980 mg/ℓ). The chloride concentrations are also high (1 790 and 2 625 mg/ℓ). Both the ammonia-nitrogen and chloride concentrations are higher than the requirements of the general discharge standard. The sodium (897 and 1 620 mg/ℓ) and potassium (1 022 and 1 150 mg/ℓ) concentrations are also high.

The heavy metals concentration (Cr, 0,17 mg/ℓ, Pb, 0,126 mg/ℓ; Ni, 0,2 mg/ℓ) were higher in one case than in another case (Cr, 0,05 mg/ℓ; Pb <0,2 mg/ℓ; Ni, 0,09 mg/ℓ). The lead concentration does not comply to the general standards for discharge. The barium (0,495 mg/ℓ), strontium (1,09 mg/ℓ), iron (3,16 mg/ℓ) and manganese (0,382 mg/ℓ) concentrations are also high. These concentrations can affect the performance of a membrane process adversely if not properly controlled during treatment.

The chemical composition of the Bisasar Road leachate show that the leachate should be treated prior to discharge into the water environment. The leachate, however, should comply to the quality requirements for discharge to the Durban Metro Sewer if the chloride concentration could be reduced to less than 1 000 mg/ℓ.

The Mariannhill leachate which is a weaker leachate than the Bisasar Road leachate also does not comply to the discharge quality requirements to the water environment and to sewer and should be treated prior to discharge.

The salinity, COD and ammonium-nitrogen concentrations of other leachates in the Durban area are also too high for discharge to the water environment. Some of the heavy metal concentrations are also too high. The chloride concentrations of most of the leachates are too high for discharge to sewer (>1 000 mg/ℓ Cl). Consequently, most of the MSW leachates should be desalinated prior to discharge to either the water environment and the sewer system.

24.14 Fouling Potential of the Municipal Solid Waste Leachate for Tubular Cellulose Acetate and Polyamide Membranes and Membrane Cleaning Strategies

Batch RO tests have indicated that the leachate will foul tubular cellulose acetate and polyamide membranes. However, it was shown in both cases that it should be possible to control membrane fouling. Preliminary results have indicated that the CWF in the case of the cellulose acetate membranes should be restored with chemical cleaning (acid and STPP and EDTA cleaning solution). Less fouling was encountered with the polyamide membranes. The CWF was much higher after treatment than in the case of the cellulose acetate membranes. It also appears that it should be possible to restore the CWF with chemical cleaning (SLS and EDTA cleaning solutions) in the case of the polyamide membranes.

24.15 Performance of Tubular Reverse Osmosis for the Concentration/Desalination of the Municipal Solid Waste Leachate

The performance of tubular reverse osmosis (cellulose acetate and polyamide membranes) for the desalination/concentration of the leachate was evaluated in a pilot scale RO unit in the batch mode of operation. The initial permeate flux was approximately 600 ℓ/m^2.d (cellulose acetate membranes) and the flux decreased as a function of percentage water recovery as a result of the increased osmotic pressure of the feed at higher water recoveries (<40%). Almost identical flux results were obtained with two runs that were conducted (up to approximately 70% water recovery). The initial and CWF at the end of the runs were almost the same. This shows that it should be possible to control membrane fouling with flow reversal and sponge ball cleaning.

The TDS of the leachate could be reduced from 8 975 to 348 mg/ℓ (96,1% removal). Therefore, an excellent quality water could be produced with RO treatment of the leachate. Ammonia-nitrogen, however, was only reduced from 882 to 82 mg/ℓ (90,7% removal). Therefore, ammonia-nitrogen removal was not that good. However, the removal of the other ions like chloride (92,4%), sulphate (99,5%), calcium (98,8%), magnesium (99,7%), potassium (94,9%), sodium (96,4%) and COD (97,7%) were excellent. The quality of the RO product with the exception of ammonia-nitrogen complies to the discharge quality requirements (water environment and sewer).

The initial permeate flux in the case of the polyamide membrane was approximately 1 200 ℓ/m^2.d and the flux also decreased as a function of percentage water recovery as before. The initial and CWF at the end of the runs were approximately the same. Therefore, it again appears that membrane fouling should not be a problem with the treatment of this type of effluent if flow reversal and sponge ball cleaning are applied.

The permeate flux through the polyamide membranes module was significantly higher than through the cellulose acetate membrane module (1 200 ℓ/m^2.d to 600 ℓ/m^2.d for polyamide membranes at end of run and 600 ℓ/m^2.d to 350 ℓ/m^2.d for cellulose acetate membranes at end of run, 70% water recovery). Therefore, more product water should be produced with the polyamide membrane module (0,181 m^2) than with the cellulose acetate membrane module (1,75 m^2).

Higher TDS removals were obtained with the polyamide membranes (97,9%) than with the cellulose acetate membranes (96,1%). Conductivity removals were 96,9% for the polyamide and 93,2% for the cellulose acetate membranes. Similar ammonia-nitrogen removals were obtained with the polyamide membranes (980 to 81 mg/ℓ, 91,7% removal) than with the cellulose acetate membranes (882 to 82 mg/ℓ, 90,7%). Between 98% to 100% removals of chloride, sulphate, calcium, magnesium, potassium and sodium were obtained. Better lead, nickel and phenol removals were also obtained with the polyamide membranes. Therefore, it

appears that the polyamide membranes should perform better for the desalination of the leachate than the cellulose acetate membranes.

24.16 Preliminary Economics of the RO Process for the Treatment of the Municipal Solid Waste Leachate

The capital cost to treat 250 kℓ/d of the MSWL with tubular cellulose acetate membranes was estimated at R2,0 million. Operational costs were estimated at approximately R11,0/kℓ.

The capital cost to treat 250 kℓ/d of the MSWL with polyamide membranes was estimated at R8,1 million. Operational costs were estimated at approximately R15,0/kℓ. (Note: Costs derived from laboratory-scale tests).

24.17 Demonstration of Tubular and Spiral Wrap Reverse Osmosis on Pilot Scale at the Bisasar Road Waste Disposal Site for the Treatment of Municipal Solid Waste Leachate

24.17.1 Tubular Cellulose Acetate Membranes

It appears that it should be possible to control membrane fouling with regular acid (phosphoric) and chemical (EDTA and SLS and/or STPP and EDTA) cleaning. The CWF and the permeate fluxes remained at approximately 500 and 300 ℓ/m^2.d, respectively, after about 500 hours of operation (feed-and-bleed).

The permeate flux after approximately 500 hours of operation (batch tests) was somewhat lower than the initial permeate flux (batch test). The initial permeate flux started at about 600 ℓ/m^2.d and was about 350 ℓ/m^2.d at 75% water recovery. The permeate flux after 500 hours of operation started at about 480 ℓ/m^2.d and was about 350 ℓ/m^2.d at 70% recovery. Permeate flux was about 20% lower after 500 hours of operation. This, however, can be expected to occur in the last module of an RO treatment train where membrane fouling is more likely to occur as a result of a higher brine concentration.

The TDS and conductivity removals were only 77,1 and 64,2%, respectively, after 500 hours of operation (batch test). The TDS and conductivity removals were 96,1 and 93,2%, respectively, on a fresh membrane surface (batch test). Therefore, a significant reduction in salinity removal has occurred as a result of membrane fouling.

24.17.2 Tubular polyamide membranes

It appears that it should also be possible to control membrane fouling with regular acid (hydrochloric) and chemical (Ultrasil 10) cleaning. The CWF and the permeate fluxes remained at approximately 500 and 200 ℓ/m^2.d, respectively, after about 500 hours of operation (feed-and-bleed).

The permeate flux after approximately 500 hours of operation (batch test) was also somewhat lower than the initial permeate flux (batch test). The initial permeate flux started at about 1 250 ℓ/m^2.d and was about 600 ℓ/m^2.d at approximately 70% water recovery. The permeate flux after 500 hours of operation started at 1 000 ℓ/m^2.d and was about 400 ℓ/m^2.d at 70% water recovery. Permeate flux was also about 20% lower after 500 hours of operation as was the case with the cellulose acetate membranes.

The TDS and conductivity removals were 96,6 and 93,4%, respectively, after 500 hours of operation (batch test). The TDS and conductivity removals were 97,9% and 96,9%, respectively, on a fresh membrane surface (batch test). Therefore, the reduction in salinity removal after 500 hours of operation was significantly less than was the case with the cellulose acetate membranes.

24.17.3 Spiral wrap membranes

Some membrane fouling was experienced during treatment of the MSWL with the spiral membranes. Permeate flux started at 838 ℓ/m^2.d (CWF 989 ℓ/m^2.d) and was 711 ℓ/m^2.d when the run was terminated after 32 hours of operation (CWF 893 ℓ/m^2.d). However, membrane cleaning after certain time intervals with Ultrasil 10 and acid was responsible for flux maintenance. Preservation of the membranes in SMBS solution increased the CWF to 1 011 ℓ/m^2.d. Therefore, the RO tests over an approximately 32 hour period have shown that it should be possible to control membrane fouling with regular chemical cleaning of the membranes. However, longer term tests should be conducted to develop a proper membrane cleaning strategy when using spiral wrap membranes for the treatment of leachate.

An excellent quality product water could be produced. TDS was reduced from 4 982 mg/ℓ in the RO feed to only 45 mg/ℓ in the RO product (99,1% removal). Conductivity was removed from 11,3 mS/cm to 0,49 mS/cm in the RO product (95,7% removal). Almost 100% of the COD (1 400 mg/ℓ) was removed. Chloride was removed from 1 535 mg/ℓ to 14,0 mg/ℓ (99,1% removal). Ammonia-nitrogen was removed from 589 mg/ℓ in the RO feed to 47,9 mg/ℓ (92,0% removal) in the RO permeate. Therefore, the RO product complies with the discharge quality requirements (except for ammonia-nitrogen) to the water environment and to the sewer system.

24.17.4 Economics

The estimated capital and operational costs to treat 250 kℓ/d of MSWL with different RO plants are as follows : -

	Capital (MR)		Operational (R/kℓ)
Tubular RO (cellulose acetate)	1,95		11,45
Tubular RO (polyamide)	8,1[(1)]	(6,5)[(2)]	16,24
Spiral RO	0,56		3,51
Disc tube RO	6,2		26,65

[(1)] Total plant from overseas supplier
[(2)] Only membranes, modules, manifolds, frames from overseas supplier.

24.17.5 Odour control

Technologies are available to control odour at landfill sites. These technologies should be applied in practice.

24.17.6 Guidelines for leachate control

Guidelines for the control of leachate are available in the second edition (1998) of the Waste Management Series known as the 'Minimum Requirements'.

Once the leachate has been collected and removed from the landfill, it must undergo some type of treatment and disposal. The most common methods of managing leachate are : -

- Discharge to a sewage treatment work
- On-site treatment followed by discharge; and
- Recirculation back into the landfill.

25. RECOMMENDATIONS

The treatment of industrial solid waste leachate with ED is a new application of the ED process. However, the process should be further researched to optimise the process in terms of : -

(a) membrane lifetime (extisting and newly developed membranes)
(b) removal of organics from the ED product with a membrane reactor
(c) further brine volume reduction by adding less water to the brine during the treatment or by treatment of the brine with evaporation or membrane distillation followed by crystallization
(d) durrent density to reduce membrane area
(e) economics of the process.

The tubular reverse osmosis process should be implemented for the treatment of municipal solid waste leachate in South Africa where required. The spiral wrap RO system using a magnetic field across the membranes holds potential for the treatment of municipal solid waste leachate due to its apparent lower cost and should be further researched to optimise this process for leachate control.

A combination of biological treatment and membrane technology (UF, NF) holds promise for the removal of membrane foulants and for the removal of recalcitrant organics from leachates and should be further investigated.

The performance of evaporation / crystallisation / biological technology for the treatment of the Holfontein type leachate should be monitored so that the process design of this technology could be used for similar leachates if proved to be successful.

26 LITERATURE

Ambachtsheer, R (1999). Bio-enhanced treatment of hazardous landfill leachate. Beta BioSystems, P O Box 25, Albertinia.

Biehler, M J and Hägeles (1995). Treatment processes of sanitary landfill leachates. Natural Resources and Development, **41**, 1995. Institute for Scientific Cooperation, Tübingen, Federal Republic of Germany.

Bilstad, T and Madland, M V (1992). Leachate minimization by reverse osmosis. *Wat. Sci. Technol.*, **25**, 117-120.

Bjorkman, V B and Wechsler, R M (1983). Physico chemical treatment of a high strength leachate. In Proceedings of the 32nd Purdue Industrial Waste Conference. Purdue University, Indiana, USA.

Chian, E S K and De Walle, F B (1976). Sanitary landfill leachates and their treatment. *J. Environ. Engng. Div.,* ASCE **102**, 411-431.

Chianese, A, Ranouro, R. and Verdone, N. (1999). Treatment of landfill leachate by reverse osmosis. *Wat. Res.* **33,** 647-652.

Chiang, L C, Chang, J E and Wen, T C (1995). Indirect oxidation effect in electrochemical oxidation treatment of landfill leachate. *Wat. Res.* **29**, 671-678.

Cossu, R., Polcaro, A M, Lavagnolo, M C, Mascia, M., Palmas, S and Renoldi, F (1998). Electrochemical treatment of landfill leachate oxidation at Ti/PbO_2 and Ti/SnO_2 anodes. *Environ. Sci. Technol*. **32**, 3570-3573.

EnviroServ. (2001). Holfontein leachate treatment plant presentation to DWAF, 4 July 2001.

Federici, N J (1998). *The effectiveness of Odor Counteractants for Control of Odor*. Air & Waste Management Assoc. 91st Annual Meeting and Exhibition, June 14-18, 1998.

Gau, S H and Chang, F S (1996). Improved Fenton method to remove recalcitrant organics in landfill leachate. *Wat. Sci. Technol.*, **34**, 455-462.

Gierlich, H H and Kolbach, J (1998). Treating landfill leachate in European countries. *Pollut. Engng*. Aug. 10-14.

Gombault, E (2002). Leachate treatment at the Holfontein hazardous waste landfill site. Enviroserv Waste Management (Pty) Ltd., P O Box 231, Bedfordview, 2008.

Heinigin, P L A (1995). Proc., Sardinia 95, 5th International Landfill Symposium, CISA, 1995, S. Margherita du Pula Cagliari, Italy, 1995.

Holz, F, Theilen, U and Weber, B (1993). Treatment of landfill leachate by reverse osmosis. Haase Energietechnik GmbH, D-2350 Neumunster, Gadelander Strasse 172, Germany.

Imai, A, Iwami, N, Matsushige, K, Inamori, Y and Sudo, R (1993). Removal of refractory organics and nitrogen from landfill leachate by the micro-organism attached activated carbon fluidized bed process. *Wat. Res.* **27,** 143-145.

Imai, A, Inamori, Y and Sudo R (1998). Effect of pre-ozonation in refractory leachate treatment by biological activated carbon fluidized bed process. *Environ. Technol.*, **19,** 213-223.

Imai, A, Onuma, K, Inamori, Y and Sudo, R (1995). Biodegradation and adsorption in refractory treatment by the biological activated carbon fluidized bed process. *Wat. Res.* **29**, 687-694.

Ince, N H (1988). Light-enhanced chemical oxidation for tertiary treatment of municipal landfill leachate. *Wat. Environ. Res.*, **70,** 1161-1169.

Keenan, J D, Steiner, R L and Fungaroli, A A (1983). Chemical-physical leachate treatment. *J. Environ. Engng. ASCE.* **109**, 1371-1384.

Kenner, R and Peters, T A (1995). WLB Wasser, Luft und Boden 1-2 (1995), 24.

Kim, S M, Geissen, S U and Volgepohl, A (1997). Landfill leachate. *Wat. Sci Technol.*, **35**, 239-247.

Kinman, R N and Nutini, D L (1991). Reverse osmosis treatment of landfill leachate. In Proceedings of the 45th Purdue Industrial Waste Conference. Purdue Industrial Waste Conference. Purdue University, Indiana, U S A.

Knok, K and Jones, P H (1979). Complexation characteristics of sanitary landfill leachates. *Wat. Res.* **13**, 839-846.

Krug, T A and McDougall, S (1989). Preliminary assessment of a microfiltration/reverse osmosis process for the treatment of landfill leachate. In Proceedings of the 43rd Purdue Industrial Waste Conference. Purdue University, Indiana, U S A.

Lema, J M, Mendez, R and Blazquez, R (1988). Characteristics of landfill leachates and alternatives for their treatment : A Review. *Wat. Air & Soil Pollut.* **40,** 223-250.

Le Roux, G (2002). Private communication. TEE@dwaf.gov.za.

Lin, S H and Chang, C C (2000). Treatment of landfill leachate by combined electro-fenton oxidation and sequencing batch reactor method. *Wat. Res.* **34** (17), 4243-4249.

McGinley, C M and McGinley, M A (1998). *Odor Quantification Methods & Practices at MSW Landfills.* Air & Waste Management Assoc. 91st Annual Meeting and Exhibition, June 14-18, 1998.

Minimum Requirements for the Handling, Classification and Disposal of Hazardous Waste, 2nd Edition, 1998, Department of Water Affairs and Forestry.

Minimum Requirements for Waste Disposal by Landfill, 2nd Edition, 1998, Department of Water Affairs & Forestry.

Minimum Requirements for Water Monitoring and Waste Management Facilities, 2nd Edition, 1998, Department of Water Affairs & Forestry.

Percival, L J, Senior, E and Southway (1997). Treatment of a high-strength leachate from a closed co-disposal landfill site in South Africa. *Water SA*, **23**, (4), 411-418.

Peters, T A (1995). Proc., Sardinia 95, 5th International Landfill Symposium, CISA, S. Margherita di Pula-Cagliari, Italy, 1995.

Peters, T A (1996). Water Quality International.

Peters, T A (1998). Purification of landfill leachate with reverse osmosis and nanofiltration. Desalination **119**, 289-293.

Pirbazari, M, Ravindron, V, Badriyha, B N and Kiu, S H (1996). Hybrid membrane filtration process for leachate treatment. *Wat. Res.* **30,** 2691-2706.

Pohlard, F (1996). Water Quality International, 9-10 (1996), 18.

Raphtel, M, Pohl, W and Peters, T A (1994). Lecture abstracts, ACHEMA 1994, Frankfurt.

Rautenbach, R and Linn, Th (1996). High pressure reverse osmosis and nanofiltration, a 'zero discharge' process combination for the treatment of wastewater with severe fouling/scaling potential. *Desalination* **105**, (1996) 6333-70.

Rautenbach, R and Mellis, R (1995). Hybrid processes involving membranes for the treatment of highly organic / inorganic contaminated wastewater. Desalination **101** (1995) 105-113.

Rautenbach, R, Vossenkoul, K, Linn, T and Katz, T (1996). Wastewater treatment by membrane processes - New developments in ultrafiltration, nanofiltration and reverse osmosis. *Desalination* **108**, (1996), 247-253.

Roberts and Sellwood (1997). *Assessing odour from composting facilities*. IWM proceedings March 1997.

Robinson, H (2000). Leachate treatment options for landfill leachate in 1st International Training Seminar, Control Management and Treatment of Landfill Emissions, University of Natal - Civil Engineering Programme, Durban - South Africa, 6 - 8 December 2000.

Schoeman, J J (1992). Electrodialysis of salts acids and bases by electro-osmotic pumping. Thesis submitted in partial fulfilment of the prerequisites for the degree of Doctor of Philosophy in the Faculty of Science, University of Pretoria, March 1992.

Schoeman, J J and Steyn, A (1999). Evaluation of electrodialysis for the treatment of a hazardous leachate. Contract report ENV/P/C 98146, CSIR, Pretoria.

Schoeman, J J, Steyn, A and Hietkamp, S (1999). Evaluation of electrodialysis for the treatment of a hazardous leachate, CSIR Contract Report ENV/P/C 98146.

Simon, G P and Calmon, C (1987). Experimental methods for the determination of non-transport properties of membranes. *Desalination,* **59**, 103, 106.

Slabbert, J L, Oosthuizen, J, Venter, E A, Hill, E, du Preez M and Pretorius, P J (1998). Development of guidelines for toxicity bioassaying of drinking and environmental waters in South Africa. WRC Report No. 358/1/98. Report to the Water Research Commission by the Division of Water, Environment and Forestry Technology, CSIR, Pretoria.

Steensen, M. (1997). Chemical oxidation for the treatment of leachate process comparison and results from full-scale plants. *Wat. Sci. Technol.,* **35**, 249-256.

Strachan, L J, Trois, C, Robinson, H D and Olufsen, J S (2000). Appropriate biological treatment of landfill leachates with full nitrification and denitrification. The Water Institute of Southern Africa's Biennial Conference and Exhibition Sun City, South Africa, 28 May to 1 June 2000.

Thomas, J, Boswell, J E S, and McIntosh, L (1999). Literature investigation regarding leachate treatment, focussing on high COD values. Literature investigation prepared by Bohleweki Environmental (Pty) Ltd, P O Box 11784, Vorna Valley, 1686, for Enviroserv.

Urase, T (2001). High pressure reverse osmosis and low pressure nanofiltration in environmental management in Japan. Paper presented at the 4th WISA-MTD Symposium, Membranes: Science & Engineering, Fleurbaix, Stellenbosch, RSA, 26-27 March 2001.

US EPA (1985). Methods for measuring the acute toxicity of effluents to freshwater and marine organisms. EPA/600/4-85/013, Environmental and Support Laboratory, Office of Research and Development, US Environmental Protection Agency, Cincinnati, Ohio, USA.

Ying, W., Bonk, R R, Loyd, V L and Sojka, S A (1986). Biological treatment of a landfill leachate in sequencing batch reactors. *Environ. Prog.,* **5,** 41-50.

APPENDIX A

HISTORY OF THE QUALITY OF THE HOLFONTEIN LEACHATE OVER A PERIOD OF TIME

Table A1: Chemical composition of the industrial solid waste leachate (Holfontein) over a period of time.

Date	20 Nov 1992	10 Mar 1993	9 Jul 1993	6 Sep 1995	12 Jul 1996	8 Jan 1997	23 Jul 1997	3 Oct 1997	14 Nov 1997	23 Dec 1997	30 Jan 1998	7 May 1998	6 Jul 1998	5 Nov 1998	21 Dec 1998	26 Jan 1999	13 Mar 1999	14 Apr 1999	28 Jun 1999	20 Sep 1999	29-Feb-00*	Average	Standard deviation	Maximum	Minimum
Constituents																									
pH	7.6	6.6	7.5	8	7.7	7.8	7.7	6.9	7.2	7.4	7.5	7.6	7.6	7.6	7.6	7.3	7.5	7.3	7.6	7.7	7.63	7.49	0.31	8.00	6.60
Conductivity (mS/m)	11000	12243	12740	22600	22350	24510	7470	8570	8510	8590	8630	9040	9190	8870	8910	8960	8840	8600	8220	8350	3660	11309.65	5281.09	24510.00	7470.00
TDS	116679	84100	96200	145140	183335	176155																133601.50	41359.40	183335.00	84100.00
Calcium	1302	2250	732	59	4050	2390	1890	2360	3080	2250	2260	1850	1390	1210	1270	1340	1250	1200	690	601	461	1671.20	929.59	4050.00	59.00
Magnesium	1134	1266	1070	446	2120	1310	1210	2160	2800	1950	2300	1870	1280	1120	1340	1200	1090	1070	796	656	387	1409.40	598.21	2800.00	446.00
Sodium	10920	12531	21500	43200	37400	41400	21800	26600	36800	32600	39400	32300	28500	31400	28600	28500	31000	34700	23700	25300	7230	29407.55	8621.10	43200.00	10920.00
Potassium	5507	7751	6000	6600	17200	15000	7650		14300	11500	13100	9290	8430	7970	6970	6770	7420	8030	5190	6080	1700	8987.26	3510.10	17200.00	5190.00
Total Alkalinity as CaCO3	17564	15862	22300	3474	17040	16589	13006	10660	11058	11207	11654	13427	6832	14139	12222	13314	1293	13243	14789	15922	3653	12779.75	4800.44	22300.00	1293.00
Chloride	10292	15472	16200	29600	44797	51934	34242	30143	34298	33900	30604	27208	26949	35106	24616	23836	36538	30248	28997	31055	9097	29801.75	9449.25	51934.00	10292.00
Sulphate	9407	12963	12700	31600	15167	21466	15500	20371	19553	22300	18531	25654	23460	20715	21349	17310	17760	17828	26938	12071	4907	19132.15	5440.14	31600.00	9407.00
Nitrate	2.6	2	0.3	<0.2	89.7	44.3	25.9	151	159	152	131	405	36.5	5.1	27.2	21.2	17.2	21.6	2.2	2.3	0.2	68.22	99.43	405.00	0.30
Fluoride				6.2	0.2	1.1	1.3	1.5	1.7	1.2	0.8	1.3		1.3	1.6	1.9	2	2.4	3.2	2.9	0.8	1.91	1.37	6.20	0.20
Arsenic			1.5	11.8	3.84	4450	0.475	1.67	1.95	7	1.89	1.8		1.54	0.808	1.18	1.15	1.16	1.13	0.964		264.11	1078.68	4450.00	0.48
Boron	72	<0.05	46.7	47.2	26.5	44	30.3	31.3	35.9	40.7	59.6			46.2	53.9	57.1		48.1	40.5	51		45.69	11.79	72.00	26.50
Cadmium	0.24	0.15	0.17	0.02	0.04	0.06	0.05	0.2	0.21	0.09	0.12	0.12	0.06	0.06	0.2	0.07		0.03	0.04	0.01		0.10	0.07	0.24	0.01
Cyanide	17.02	20.33	155	312	2261	48.6	49.6	21.9	96.8	75.4	51.7		275	270	277	224		231	190	373	36	274.96	508.77	2261.00	17.02
Chromium	0.9	2.1	2.6	2.09	10.5	15.8	11.7	11.1	5.36	11.8	6.26	5.62	4.24	4.31	8.06	4.67		4.59	11.5	8.66		6.94	4.16	15.80	0.90
Copper			0.2	0.03	<0.03	0.07	0.08	0.28	0.08	0.29	0.19	0.03	0.03	0.03	0.04	0.03		0.03	0.03	0.03	0.03	0.09	0.09	0.29	0.03
Manganese			12.5	2.22	41.5	49	68.3		70.3	77.9	60.2	59.8	60.2	70.1	47.2	46.4		75.3	16.1	15.1	24.5	48.26	24.46	77.90	2.22
Lead	4.9	40.5	1.33	0.17	0.2	0.17	0.59	3.36	1.51	1.72	1.4	1.12	4.24	0.86	1.03	0.8		0.44	0.55	1.99		3.52	9.05	40.50	0.17
COD	80594	59200	50000	53200	130000	111000	85800	119120	122000	111500	113600	107769	133000	93500	80000	90500	101000	84000	82000	64500	20000	93614.15	24950.92	133000.00	50000.00
Phenol	338	52	422	68	2619	1324	1751	4828	4400	2111	1608	2500	5512	1940	2518	3500	2750	4081	542	440	219	2165.20	1648.25	5512.00	52.00

Phosphate			50	35	2.1	2.6	6.5		26.5	14.8	15.5			20.3	0.2	4.3	1.7	1.8	1.7	2.5		12.37	14.86	50.00	0.20
DOC	1850	2148	11200	14200	43179	32533	6902	40503	37500	27300	44075	30017	29246	24057	28548	37972	26792	6508	27740	28180	182	25022.50	13442.74	44075.00	1850.00
Sulphide			312	440	140	44	52	360	224	208	154			60	72	64		56	48	183	64	161.13	126.57	440.00	44.00
Ammonia as N	570	751	1700	1361	152	3162	2023	2181	2790	2200	2583	1794	700	1370	1428	1805		2378	3729	4	273	1720.05	1000.77	3729.00	4.00
TOX			200	940	520	210	670	660	620	1020	1010	750		1040	820	1690		2300	2000	2610		2016.88	4029	1690	200
Barium			1.2	0.2	1.92	0.27	0.47	2.07	0.88	0.58	0.65	0.38		0.24	0.26	0.31	0.23	0.17	0.09	0.15	0.29	122.24	501.93	2070.00	0.09
Mercury	0.008	0.102	11	0.001	0.002	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.001	0.004	0.001	0.004	0.001	0.001	0.001	0.001	0.56	2.46	11.00	0.001
Chromium VI																									

* Sample data 29 Feb 00 not included in statistical analysis.

APPENDIX B

BIODEGRADABILITY TESTS

BIODEDRADABILITY TESTS ON THE HOLFONTEIN LEACHATE

1. BACKGROUND AND AIM

Environmentek is investigating the use of electro dialysis for the removal of pollutants from hazardous wastes in order to reduce the volumes of waste dams. Our laboratory was requested to evaluate the biodegradability of the waste before treatment and the products formed during the treatment process. The effect of biosupplements on the biodegradation process was also studied. The waste (feed / leachate) was collected from Dam 4 at the Holfontein Waste Site. The feed was pretreated (treated feed) by adding ISCOR ash (200 g/ℓ), soda ash (12 g/ℓ) and caustic soda (7 g/ℓ) to the waste and bubbling the mixture with carbon dioxide. The pre-treated sample was then desalinated (product) by electrodialysis (ED)

2. METHODOLOGY

A Micro-Oxymax Respirometer (Columbus Instruments) was used to monitor the biodegradability of the waste and treatment products. The automated machine can run 10 samples simultaneously using the same oxygen and carbon dioxide gas sensors. The sensors are calibrated before each run and a comprehensive self-testing programme is performed to verify sensor calibration and check for leaks.

Activated sludge collected from the Daspoort Sewage Works in Pretoria was used for this study. Due to the high COD values of the feed it was decided to perform the test on the full strength and 5x dilutions of the samples in case the measurements of the undiluted samples would fall outside the range of the sensors. The samples were incubated at 20°C in 250-mℓ test chambers with continuous shaking for 5 days. The oxygen and carbon dioxide levels were measured at 2 –h intervals in the chamber headspace and recorded in a data file. The data file was then imported into Excel for analysis.

The reaction mixture included the following components:

- 45 g sample,
- 2.5 g activated sludge; and
- 100 mg/ℓ N and 110 mg/ℓ P final concentration.

In the negative control the sample was replaced with saline and in the positive control with 1 g/ℓ glucose.

The following samples were used in the studies:

- Effluent, Holfontein (received 08/08/2001)
- Untreated feed (19/10/2000,run 4)
- Treated feed (8/12/2000)
- Product (12/12/2000, run 5)
- Brine (12/12/200, run 5)

The reaction mixture combinations investigated with the respirometer are summarized in Tables B1 and B2.

Table B1: Samples prepared on 2 February 2001.

Chamber	Sample
1	Raw feed, 5xdilution, autoclaved activated sludge
2	Raw feed, 5x dilution, activated sludge
3	Treated feed, 5x dilution, activated sludge
4	Product, 5x dilution, activated sludge
5	Brine, 5x dilution, activated sludge
6	Raw feed, autoclaved activated sludge
7	Raw feed, activated sludge
8	Treated feed, activated sludge
9	Product, activated sludge
10	Brine, activated sludge

Table B2: Samples prepared on 13 February 2001.

Chamber	Sample
1	Negative control
2	Positive control
3	Product, activated sludge
4	Product, Sybron Biosupplement 1004 TX, activated sludge
5	Product, Sybron Biosupplement 1002 CG, activated sludge
6	Product, Sybron Biosupplement 1003 FG, activated sludge
7	Feed
8	Feed, Sybron Biosupplement 1004 TX, activated sludge
9	Feed, Sybron Biosupplement 1002 CG, activated sludge
10	Feed, Sybron Biosupplement 1003 FG, activated sludge

The COD of all the samples were determined at the beginning and the end of the experiment.

3. RESULTS AND DISCUSSION

Some of the results represented graphically in Figures B1 to B8 are summarized in Table B3.

Biological degradation of the organic compounds are slow. Degradation of the leachate started after approximately 15 hours, with a rapid increase in oxygen consumption for the 5x dilutions of the ED feed and product. Degradation of the 5x dilution of the ED brine started after a contact time of approximately 26 hours. Degradation of the undiluted ED product, feed and brine started after contact times of approximately 22, 36 and 68 hours, respectively. Higher COD percentage removals (34 – 40 %) were obtained for all of the different diluted ED streams compared to the undiluted ED streams (0,2 – 30%). This can most probably be ascribed to the reduction in the toxicity effect of compounds in the leachate for micro-organisms.

The addition of biosupplements had a positive effect on the percentage COD removal for both the ED feed and product as shown in Table 3. In case of the ED feed the COD removal increased from approximately 5,4 % to between approximately 7,8 and 8,1% in the presence of the different biosupplements. With the ED product a COD removal of approximately 31% was achieved and 46% in the presence of the biosupplements compared to approximately 24,8% to before.

Table B3: COD removal during respirometery.

Samples	COD start	COD Removed	COD Removed
	mg/ℓ	(Calc) (mg/ℓ)	(Calc) (%)
Feed + inactivated sludge	57700	133	0,23
Feed	57700	4578	7,93
Treated Feed	41550	3044	7.33
Product	20150	6044	30,00
Brine	55150	444	0.81
Diluted feed + inactivated sludge	11540	2644	22,91
Diluted feed	11540	4022	34,85
Diluted Treated feed	8310	3356	40,38
Diluted Product	4030	1378	34,19
Diluted Brine	11030	3822	34,65
Plus Biosupplements			
Feed	59700	3200	5,36
Feed + 1004 TX	67950	5325	7,84
Feed + 1002 CG	69050	5400	7,82
Feed + 1003 FG	60000	4825	8,04
Product	23000	5700	24,78
Product + 1004 TX	29350	9250	31,52
Product + 1002 CG	27400	10300	37,59
Product + 1003 FG	21450	9775	45,57

CONCLUSIONS

Biological breakdown of organics (COD) in the leachate is slow, with biological activity starting after approximately 15 and 26 hours contact time for the 5x dilution and undiluted leachate, respectively.

COD removals increased with the addition of biosupplements for both the ED feed and ED product.

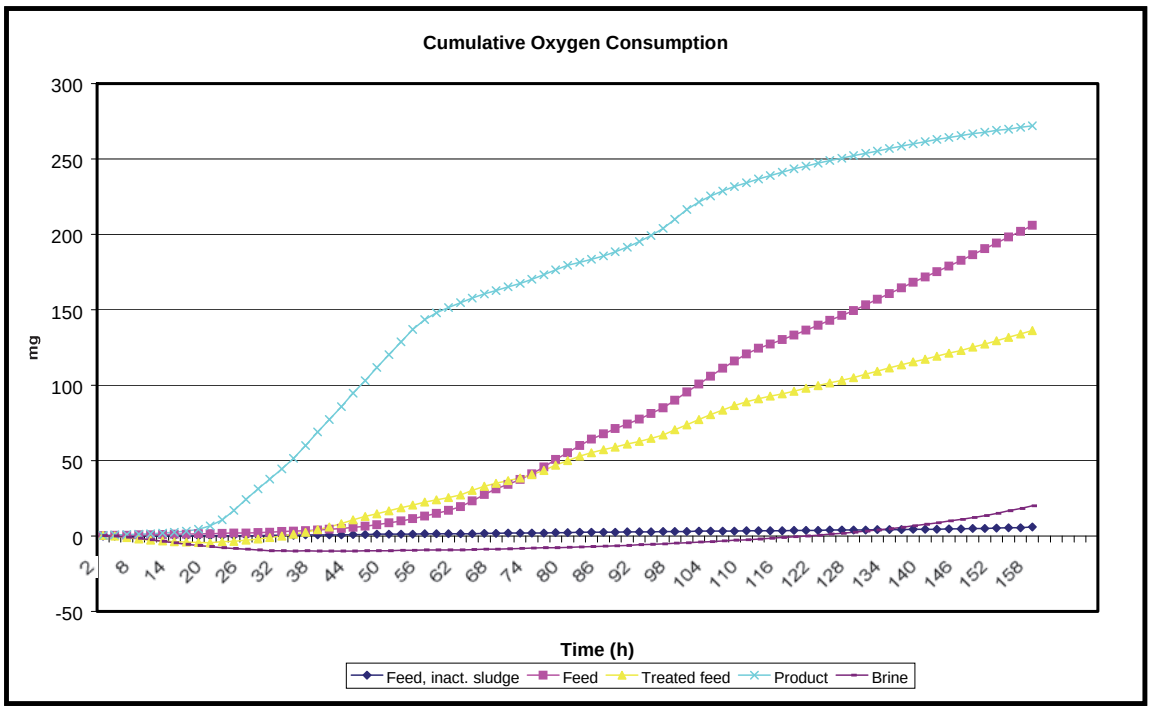


Figure B1. Oxygen consumption as a function of time (leachate, treated leachate, ED product, ED brine).

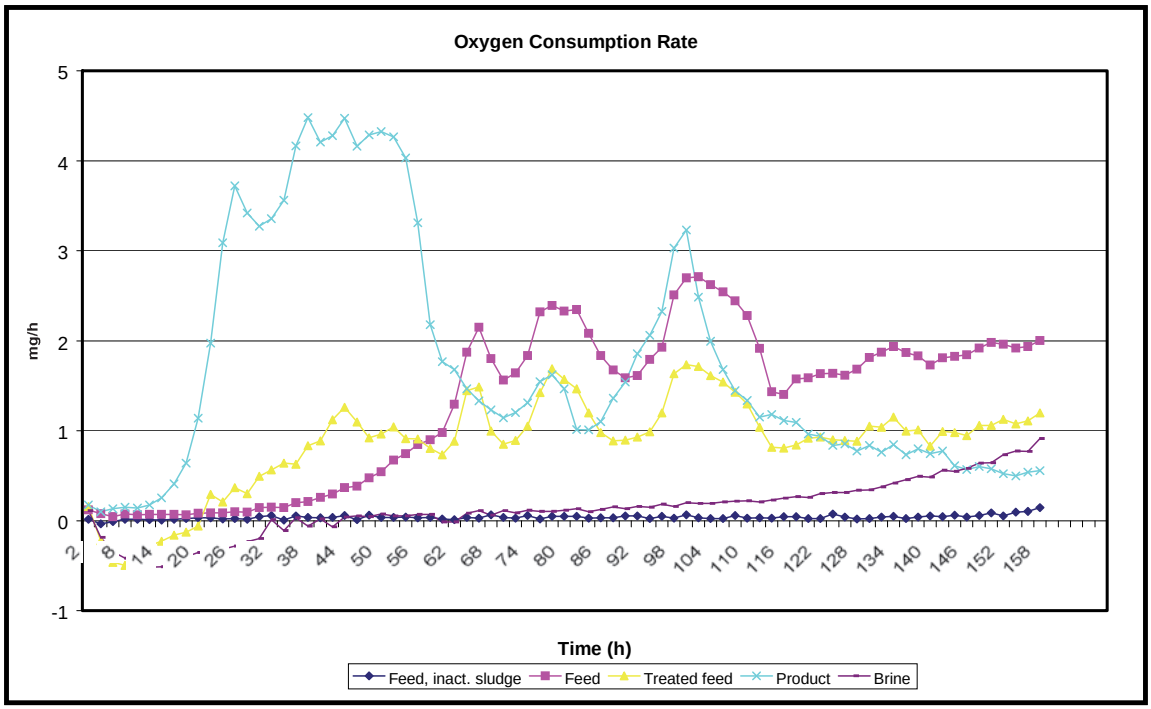


Figure B2. Oxygen consumption rate as a function of time (leachate, treated leachate, product, ED brine).

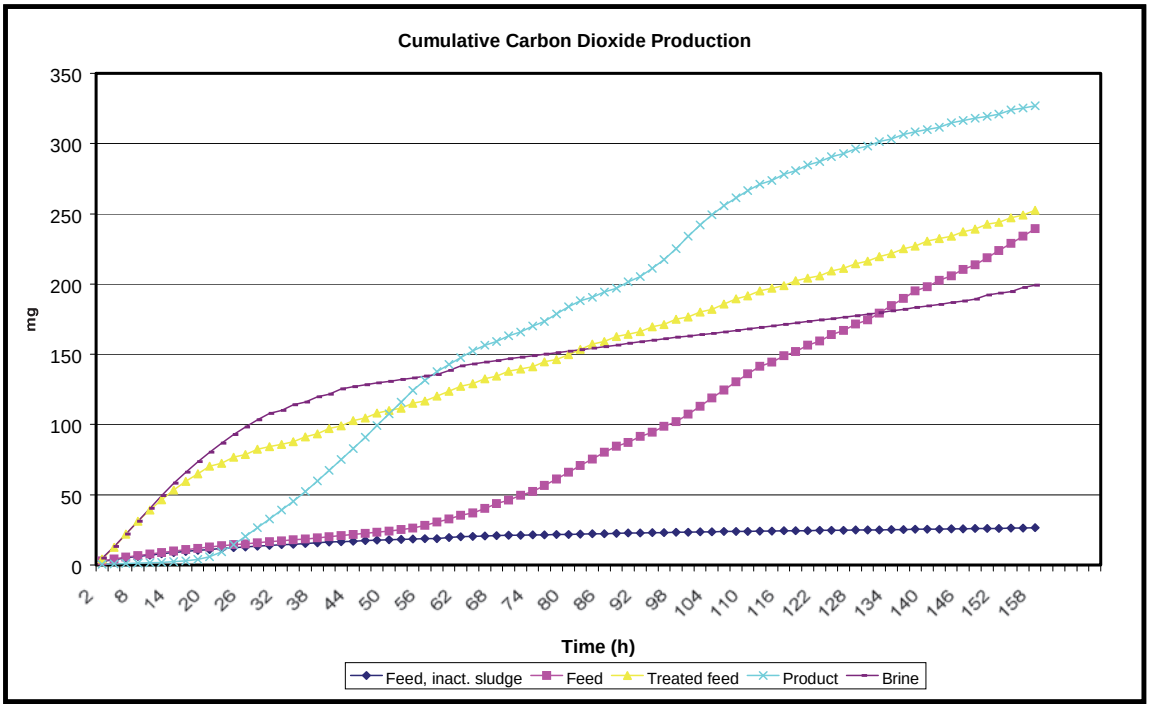


Figure B3. Carbon dioxide production as a function of time (leachate, treated leachate, ED brine).

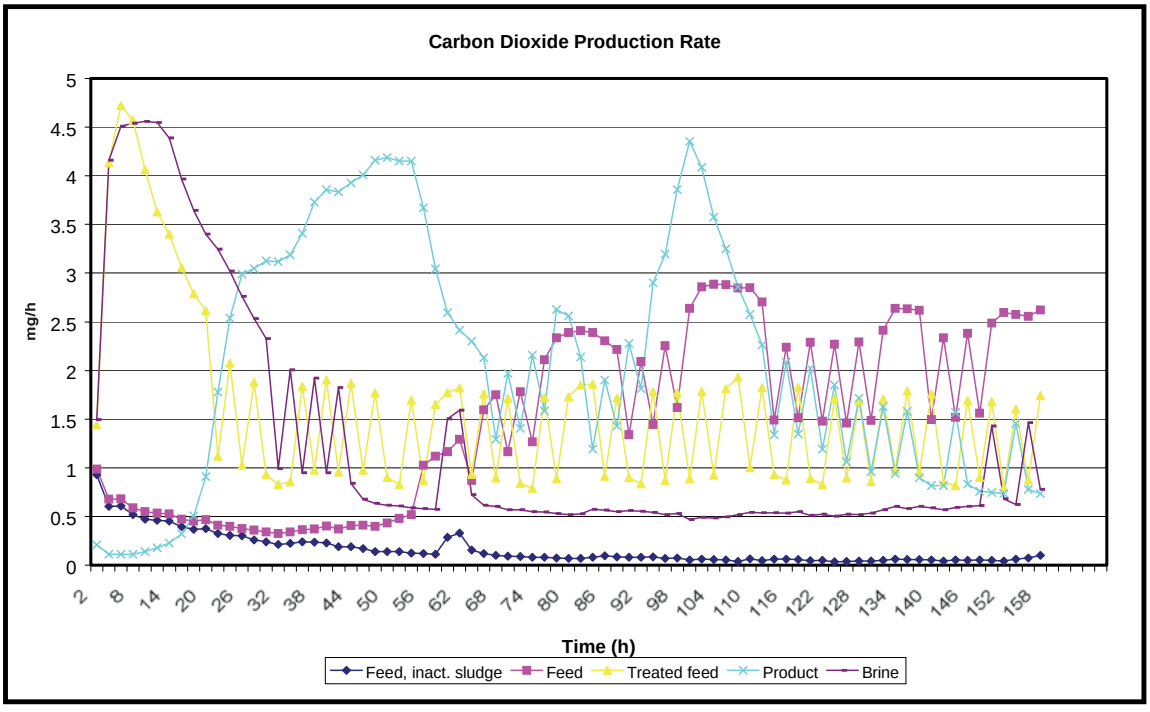


Figure B4. Carbon dioxide production rate as a function of time (leachate, treated leachate, ED product, ED brine).

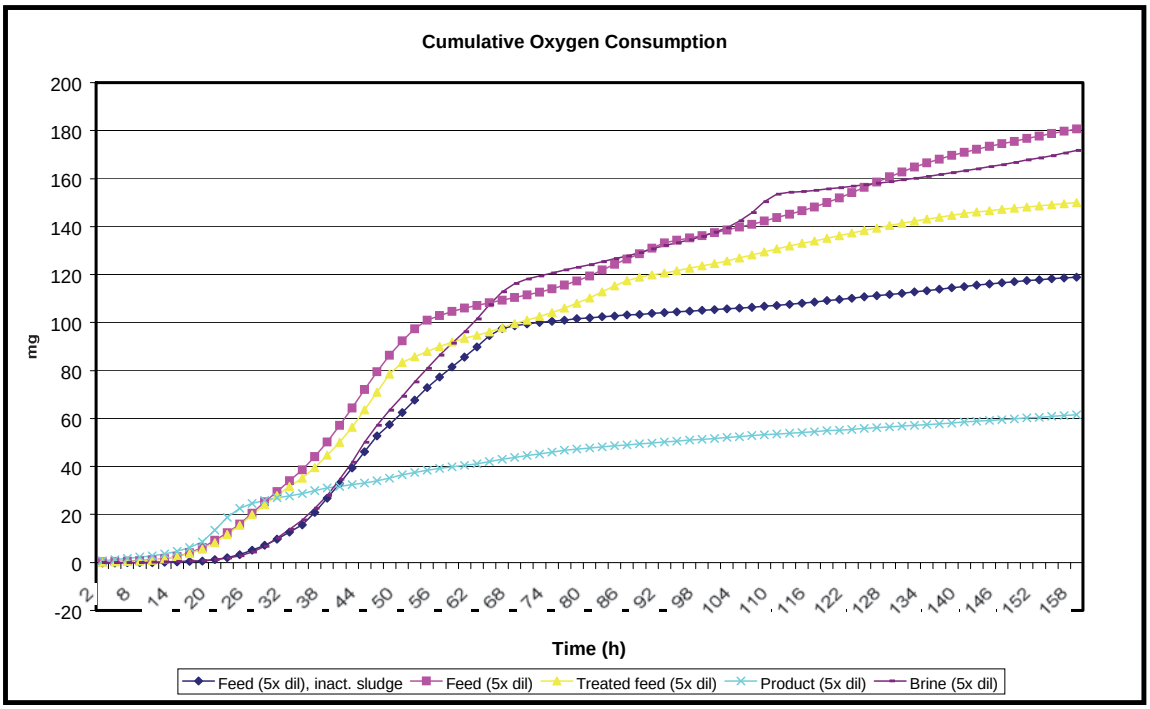


Figure B5. Oxygen consumption as a function of time (5x diluted leachate, treated leachate, ED product, ED brine).

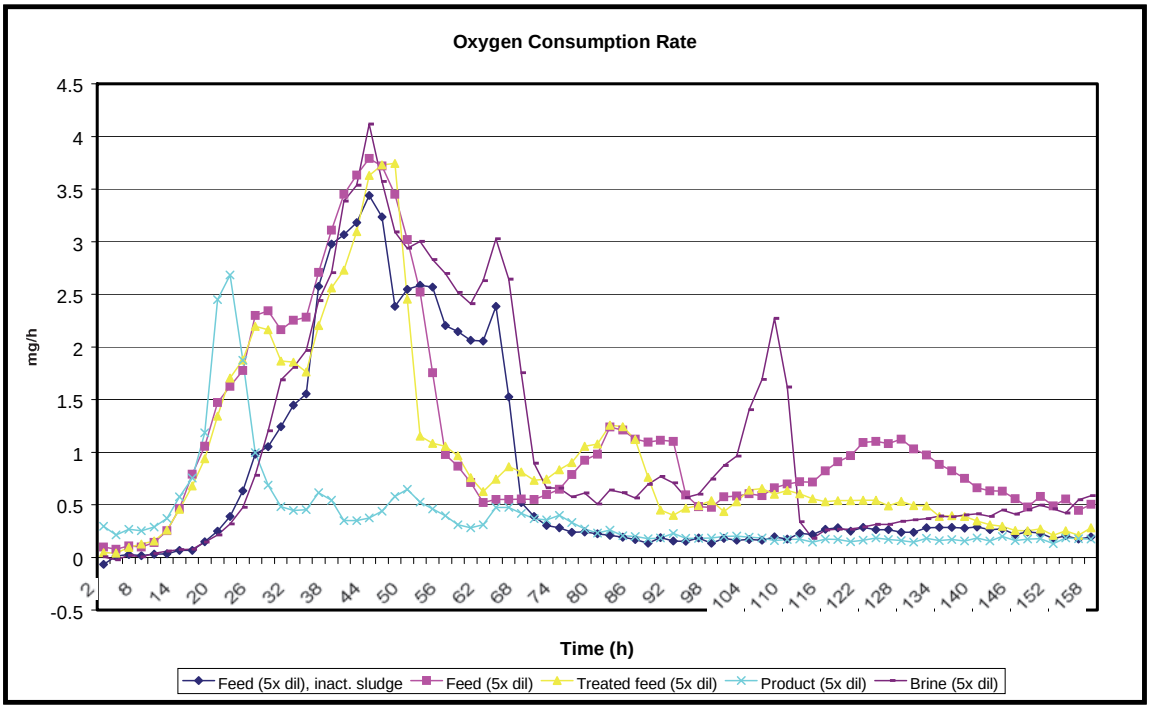


Figure B6. Oxygen consumption rate as a function of time (5 x diluted leachate, treated leachate, ED product, ED brine).

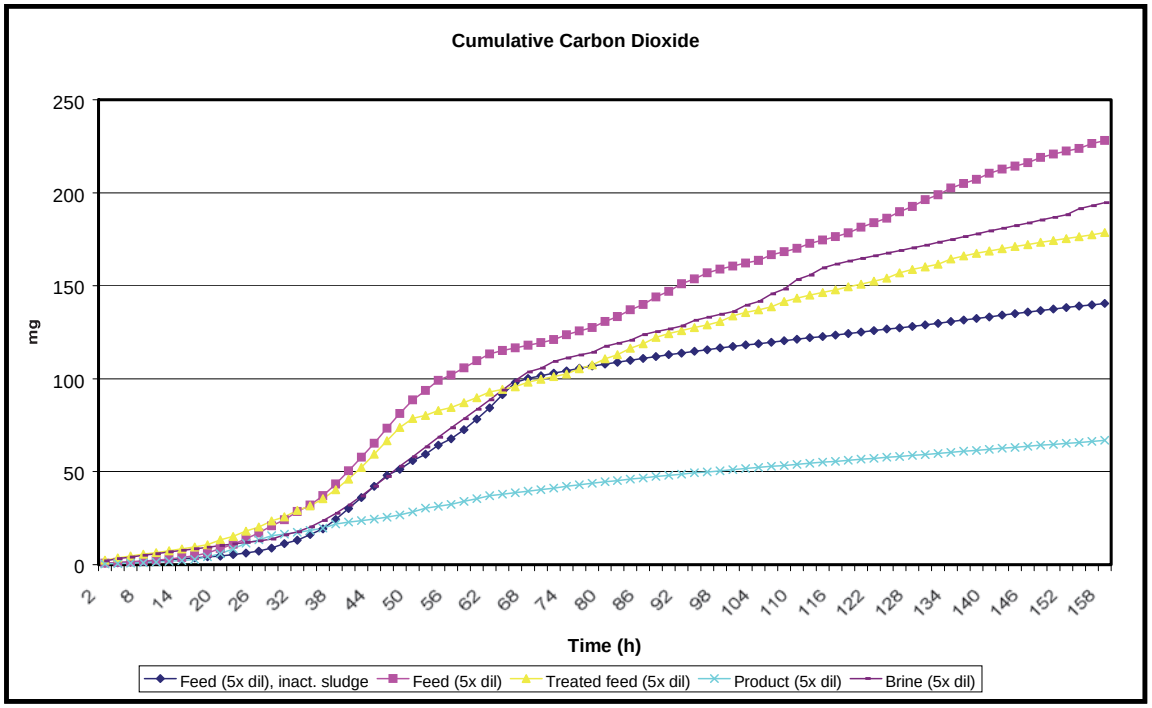


Figure B7. Carbon dioxide production as a function of time (5x diluted leachate, treated leachate, ED product, ED brine).

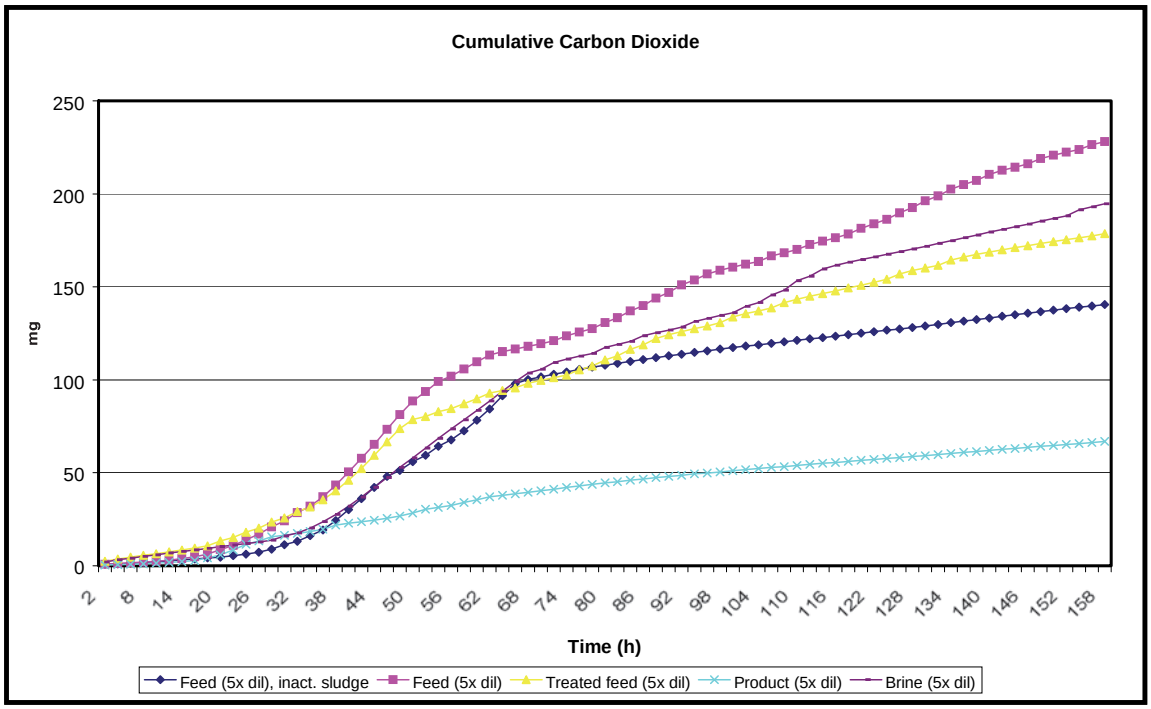


Figure B8. Carbon dioxide production rate as a function of time (5 x diluted leachate, treated leachate, ED product, ED brine).

APPENDIX C

COMPOSITION OF ASH SAMPLES

Table C1: Composition of Lethabo, Sasol and Iscor Ash Samples

Sample Name	Fe_2O_3 %	MnO %	Cr_2O_3 %	V_2O_5 %	TiO_2 %	CaO %	K_2O %	P_2O_5 %	SiO_2 %	Al_2O_3 %	MgO %	Na_2O %	Cl %	S %	C %
Lethabo	3,95	0,05	0,04	0,03	1,67	4,72	0,53	0,26	52,9	32,5	1,3	0,3	0	0,08	0,77
Sasol	4,18	0,03	0,03	0,01	1,65	6,53	0,47	0,15	53,4	28,4	1,0	0,4	0	0,17	2,27
Iscor	16,4	0,18	0,03	0,02	0,59	5,28	0,61	0,12	13,2	10,9	2,9	0,8	0,06	1,33	38,9

Technology Group

Private Bag 40175
2022 Cleveland
Telefax (011) 629-5528
Telephone (011) 629-5111

Tegnologie-groep

Privaatsakg 40175
2022 Cleveland
Telefaks (011) 629-5528
Telefoon (011) 629-5111

SANAS

SANAS ACCREDITED LABORATORY
ISO Guide 25/SABS 0259 (1990)
and EN45001 (1989)

ESKOM

Chemical Sciences

CERTIFICATE OF ANALYSIS

TO: CHEM SERVICES MANAGER
LETHABO POWER STATION
PRIVATE BAG X415
VEREENIGING
1930

THIS REPORT CONTAINS
2
PAGES

REPORT NUMBER
100088662

LETHABO POWER STATION
SUBMITTED ON 13-MAR-2000
REPORTED ON 07-APR-2000
COMPOSITE COAL FOR FEB-2000

AIR-DRIED BASIS

SAMPLE NUMBER		*200229560*
DESCRIPTION		*FEB-2000*
COAL PREP(COMPOSITE)	-	*YES*
INHERENT MOISTURE	%	*5.0*
ASH	%	*40.4*
VOLATILE MATTER	%	*17.9*
FIXED CARBON(BY DIFFERENCE)	%	*36.7*
CARBON	%	*42.17*
HYDROGEN	%	*1.93*
NITROGEN	%	*0.90*
TOTAL SULPHUR	%	*0.70*
CARBONATE(AS CO2)	%	*1.37*
OXYGEN(BY DIFFERENCE)	%	*7.53*
GROSS CALORIFIC VALUE	*MJ/Kg*	*15.44*
CHEMICAL ANALYSIS OF ASHED COAL		
SILICON(AS SIO2)	%	*54.5*
ALUMINIUM(AS AL2O3)	%	*31.6*
IRON(AS FE2O3)	%	*3.8*
TITANIUM(AS TIO2)	%	*1.6*
PHOSPHORUS(AS P2O5)	%	*0.35*
CALCIUM(AS CAO)	%	*4.9*
MAGNESIUM(AS MGO)	%	*1.3*
SODIUM(AS NA2O)	%	*0.0*
POTASSIUM(AS K2O)	%	*0.5*
SULPHUR(AS SO3)	%	*2.4*
MANGANESE(AS MNO)	%	*0.04*

ELEMENTAL ANALYSIS OF ASHED COMPOSITE COAL FROM LETHABO POWER STATION FOR THE PERIOD 01-JAN-1997 TO 31-JUL-1998

SAMPLE DESCRIPTION	SiO2	Al2O3	Fe2O3	TiO2	P2O5	CaO	MgO	Na2O	K2O	SO3	MnO
DEC-96	53.6	31.5	3.6	1.6	0.3	4.3	1.1	0.0	0.4	2.2	0.03
FEB-97	53.0	33.2	3.5	1.6	0.3	4.2	1.2	0.0	0.4	2.6	0.03
MAY-97	53.8	31.4	3.1	1.5	0.3	3.8	1.1	0.0	0.4	1.9	0.01
JUN-97	56.1	32.0	3.5	1.5	0.3	4.4	1.2	0.0	0.4	2.5	0.02
JUL-97	54.8	31.7	3.6	1.6	0.3	4.8	1.3	0.0	0.5	2.8	0.02
AUG-97	51.8	32.3	3.5	1.6	0.3	4.2	1.2	0.0	0.5	2.6	0.02
SEP-97	52.6	32.4	3.0	1.5	0.3	3.6	1.0	0.0	0.4	2.5	0.03
OCT-97	53.5	30.7	3.3	1.5	0.3	3.9	1.1	0.0	0.4	2.4	0.04
NOV-97	55.2	31.8	3.8	1.5	0.4	4.0	1.1	0.0	0.5	2.4	0.04
DEC-97	53.1	32.0	3.4	1.5	0.3	4.1	1.2	0.0	0.5	2.1	0.03
JAN-98	51.7	32.4	3.4	1.6	0.3	4.5	1.2	0.0	0.4	2.0	0.03
FEB-98	53.5	32.6	3.4	1.6	0.3	4.4	1.2	0.0	0.4	2.3	0.03
MAR-98	54.2	31.8	3.5	1.6	0.3	4.4	1.2	0.0	0.4	2.0	0.03
APR-98	54.3	31.8	3.5	1.6	0.3	4.3	1.2	0.0	0.4	2.0	0.03
MAY-98	53.6	30.8	3.7	1.6	0.3	4.5	1.2	0.0	0.5	2.2	0.03
AVERAGE VALUE	53.9	31.8	3.3	1.6	0.3	4.2	1.2	0.0	0.6	2.3	0.03
MINIMUM VALUE	51.7	30.5	0.4	1.5	0.3	3.6	1.0	0.0	0.4	1.9	0.01
MAXIMUM VALUE	56.2	33.2	3.8	1.6	0.4	4.8	1.3	0.0	3.4	2.8	0.04
STANDARD DEVIATION	1.3	0.7	0.8	0.1	0.0	0.3	0.1	0.0	0.7	0.3	0.01
95% CONFID. LIMIT	2.5	1.4	1.5	0.1	0.1	0.6	0.1	0.0	1.4	0.6	0.02

REPORT EXECUTED BY W M DELPORT ON 19-AUG-98

ELEMENTAL ANALYSIS OF ASHED COMPOSITE COAL FROM KENDAL POWER STATION

FOR THE PERIOD 01-JAN-1997 TO 31-MAR-1988

SAMPLE DESCRIPTION	SiO2	Al2O3	Fe2O3	TiO2	P2O5	CaO	MgO	Na2O	K2O	SO3	MnO
DEC-96	50.2	31.4	4.1	1.5	0.6	5.2	1.6	0.0	0.7	2.6	0.03
JAN-97	51.3	32.1	4.4	1.6	0.7	5.5	1.8	0.0	0.7	2.8	0.03
FEB-97	50.8	32.1	4.2	1.6	0.7	5.3	1.7	0.0	0.7	2.6	0.03
MAR-97	49.3	31.0	4.3	1.6	0.7	5.6	1.8	0.0	0.7	2.5	0.02
APR-97	50.1	32.6	4.7	1.7	0.8	5.7	1.9	0.0	0.7	3.6	0.03
MAY-97	49.9	31.5	4.4	1.6	0.8	5.5	1.7	0.0	0.7	3.9	0.03
JUN-97	50.4	31.1	4.5	1.6	0.7	5.2	1.7	0.0	0.7	3.0	0.02
JUL-97	52.0	32.0	4.5	1.6	0.8	5.5	1.8	0.0	0.7	3.3	0.03
AUG-97	52.3	32.0	4.4	1.6	0.8	5.5	1.7	0.0	0.7	2.9	0.03
SEP-97	50.2	30.7	4.3	1.6	0.9	5.3	1.6	0.0	0.7	3.5	0.02
OCT-97	51.1	32.2	4.3	1.6	0.8	5.2	1.7	0.0	0.6	3.3	0.03
NOV-97	49.4	32.3	4.0	1.5	0.7	4.9	1.6	0.0	0.6	3.4	0.04
DEC-97	51.1	31.0	4.1	1.6	0.8	5.2	1.6	0.0	0.7	3.0	0.04
JAN-98	51.2	30.4	4.3	1.6	0.7	4.8	1.5	0.0	0.7	2.7	0.03
AVERAGE VALUE	50.5	31.5	4.3	1.6	0.7	5.3	1.7	0.0	0.7	3.1	0.03
MINIMUM VALUE	48.4	30.4	4.0	1.5	0.6	4.8	1.5	0.0	0.6	2.5	0.02
MAXIMUM VALUE	52.3	32.6	4.7	1.7	0.9	5.7	1.9	0.0	0.7	3.9	0.04
STANDARD DEVIATION	1.0	0.7	0.2	0.0	0.1	0.3	0.1	0.0	0.0	0.4	0.01
95% CONFID. LIMIT	2.1	1.4	0.4	0.1	0.1	0.5	0.2	0.0	0.1	0.8	0.01

REPORT EXECUTED BY W M DELPORT ON 13-MAR-98

ELEMENTAL ANALYSIS OF ASHED COMPOSITE COAL FROM TUTUKA POWER STATION

FOR THE PERIOD 01-JAN-1998 TO 01-May-1999

SAMPLE DESCRIPTION	SiO2	Al2O3	Fe2O3	TiO2	P2O5	CaO	MgO	Na2O	K2O	SO3	MnO
DEC-97	56.5	24.7	5.6	1.3	0.1	4.6	1.9	0.1	0.7	2.1	0.04
JAN-98	56.7	25.1	5.6	1.3	0.1	5.0	2.1	0.1	0.6	4.8	0.04
FEB-98	55.2	24.0	6.0	1.3	0.1	4.9	1.8	0.0	0.6	4.7	0.05
MAR-98	55.5	24.8	5.5	1.3	0.1	4.6	1.8	0.0	0.7	3.1	0.04
APR-98	57.0	25.6	5.1	1.3	0.1	4.2	1.7	0.1	0.7	3.6	0.04
MAY-98	55.6	26.1	5.0	1.2	0.1	4.2	1.6	0.0	0.8	3.1	0.03
JUN-98	56.6	25.3	5.3	1.2	0.1	4.5	1.7	0.0	0.8	3.8	0.04
JUL-98	56.4	26.0	5.4	1.2	0.1	4.4	1.7	0.0	0.8	4.3	0.04
AUG-98	55.0	25.7	5.3	1.3	0.3	4.2	1.5	0.2	0.7	3.0	0.04
SEP-98	55.3	25.4	5.5	1.3	0.2	4.5	1.8	0.1	0.6	5.2	0.04
OCT-98	52.8	26.7	6.1	1.3	0.2	5.4	1.4	0.0	0.6	3.5	0.04
NOV-98	58.2	25.2	5.9	1.3	0.1	4.6	1.9	0.0	0.7	4.3	0.04
DEC-98	58.2	26.3	5.5	1.3	0.1	4.1	1.7	0.0	0.9	4.1	0.04
JAN-99	56.4	27.2	5.0	1.2	0.1	3.5	1.5	0.0	0.8	4.2	0.04
FEB-99	56.6	26.2	5.1	1.2	0.1	3.5	1.5	0.0	0.9	4.1	0.04
AVERAGE VALUE	56.3	25.7	5.4	1.3	0.1	4.4	1.7	0.0	0.7	3.8	0.04
MINIMUM VALUE	52.8	24.0	4.9	1.2	0.1	3.5	1.4	0.0	0.6	2.1	0.03
MAXIMUM VALUE	58.2	27.2	6.1	1.3	0.3	5.4	2.1	0.2	0.9	5.2	0.05
STANDARD DEVIATION	1.3	0.8	0.4	0.0	0.0	0.5	0.2	0.1	0.1	0.8	0.00
95% CONFID. LIMIT	2.7	1.6	0.7	0.1	0.1	1.1	0.4	0.1	0.2	1.6	0.01

REPORT EXECUTED BY W M DELPORT ON 28-APR-1999

APPENDIX D

FOULING POTENTIAL OF THE INDUSTRIAL SOLID WASTE LEACHATE FOR ED MEMBRANES AND EVALUATION OF MEMBRANE CLEANING STRATEGIES OF THE FOULED ED MEMBRANES WITH COMMERCIALLY AVAILABLE MEMBRANE CLEANING AGENTS

Table D1. Fouling potential of the leachate for the AFN anionic ED membrane and membrane cleaning with conventional methods.

Time	Vt	Vanion	Ampere	mS/cm	°C	pH	
0	24	2,619	0,75	90,9	21,5	8,01	
1	21,5	2,098			25,5		
2	20	1,862			27,3		
3	19,5	1,553			27,9		
4	19	1,43			28,3		
5	19	1,384			28,5		
6	19	1,361			28,6		
7	18,5	1,363			28,6		
8	18,5	1,373			28,6		
22	19,5	1,616			25,8		
25	19,5	1,644			26,7		
27	19	1,639			26,9		
29	19	1,688			27,6		
31	19	1,724			28		
46	19,5	2,281			26,6		
48	19,5	2,327			27		
50	19,5	2,369			27,8		
52	19,5	2,325			28,5		
55	19,5	2,338			28,8		
75	19	2,923			28,5		
118	24,5	6,156			28,4		
120	25	6,472			29,4		
121,5	24,5	6,493			29,8	Open stack clean anion membrane manually	
121,5	19,5	1,677			29,2	Slime on feed side of anion membranes	
122	19,5	1,287			29,4		
124	20	1,458			30,3		
126	20	1,964			30,5		
126	17,5	1,363			24,1	Fresh feed	
128	16,5	1,239		90,3	29,8	8,01	
142	16,5	1,397			28,9		
144	17,5	1,353			29,5		
146	17,5	1,326			29,9		
152	17,5	1,81			31		
176	17	1,576			29,4		
178	17,5	1,567			30		
180	18	1,632			31,8		
182	18	1,836			31,2		
184	17,5	1,846			31,5		
186	17,5	1,818			31,9		
200	19,5	1,583			30,1		
202	19,5	1,586			30,5		
204	19,5	1,595			31,2		
206	19,5	1,584			32		
208	19	1,595			32,3		
210	19	1,612			32,1		
224	20,5	1,811			30,4		
226	20,5	1,832			30,7		
228	20,5	1,808			31,3		

230	20	1,726			31,8	
232	19,5	1,643			31,8	
257	19,5	1,667			31,7	
296	21,5	2,621			30,1	
298	21,5	2,74			30,1	Change polarity
298	16,5	1,364			30,1	
300	16	2,553			28,2	
302	16	2,224			30,6	
304	16	2,279			31,1	
318	16,5	1,94			29,5	
320	16,5	2,022			29,7	
322	16	1,941			30,4	
324	17	1,892			30,4	
324	17	1,893			27,2	Fresh feed
326	16,5	1,83			30	
328	16,5	1,842			29,8	
342	18	1,707			28,1	
344	18	1,747			28,4	
346	18	1,707			29	
348	17,5	1,665			29,6	
350	17,5	1,643			29,8	
352	17	1,624			30	
390	19	1,55			28,2	
395	18,5	1,422			29,5	
399	16,5	1,104			29,8	
462	17,5	1,18			27,4	
464	18	1,328			28	
466	18	1,063			28,9	
468	18,5	1,145			29,6	
470	18,5	1,296			29,8	
472	19	1,367			29,8	
486	22	0,853			27,7	Change polarity.
486	19,5	0,902			22,9	Fresh feed
487	18,5	1,464			27	
509	18	1,411			29	
525	18	1,214			26,7	
528	18	1,363			28,5	
533	18	1,346			29,6	
535	18	1,36			29,8	
549	17	1,39			27,2	
555	17	1,231			29	
558	17,5	1,269			29,3	
621	22	2,027			26,7	
628	20,5	1,865			28,5	
630	21	1,617			28,9	
645	22	1,822			27,2	
648	22	1,625			28	
654	22	1,786			29,4	
669	22	1,713			28	
675	22	1,763			29,5	
678	22	1,832			30	
693	23	1,857			28,1	Clean with 3% NaCl, pH 10,5

693	18	0,967			22,5	Fresh feed.
695	16,5	1,219			29	
710	17,5	1,512			27,5	
713	17,5	1,482			27,6	
719	17,5	1,817			29,4	
722	17	0,565			29	After 48 hours shutdown.
725	17,5	1,019			29,5	
740	18	1,285			27,8	
743	18	1,343			28,7	
748	18	1,542			29,8	
764	18,5	1,487			28,4	
768	19	1,496			29	
772	18,5	1,479			27	
773	18,5	1,427			28,5	
788	19	1,729			27,4	
795	18,5	1,736			28,5	
797	18,5	1,769			28,5	
812	19,5	2,007			26,4	
821	19,5	2,057			28,5	
836	20,5	2,135			26,5	
836	25	1,261			17,6	Start after 8 days shutdown/Resistance measurement
837	23	1,239			21,3	
838	22,5	1,098			23,2	
839	22,5	1,333			24	
892	22,5	1,241			23,5	
907	23,5	1,238			22	
908	24	1,385			22,3	

Vt : Total voltage across fouling cell
Varion : Voltage across anion membrane
Current density : 100 mA/cm^2

APPENDIX E

TOXICITY TESTS

Toxicity Testing Laboratory
Water Programme

Telephone : 012 841-3953
Fax : 012 841-2506
e-mail: lslabbert@csir.co.za

TEST REPORT

TOXICITY EVALUATION OF UNTREATED AND TREATED HAZARDOUX WASTE

REPORT NUMBER 0147

Requested by:	Dr Japie Schoeman
Organization:	Environmentek, CSIR
Address:	P O Box 395 PRETORIA 0001
Telephone number:	012 841 2252/2270
Fax number:	012 841 2506
Date of sample receipt:	15 January 2001
Test commencement date:	17 January 2001
Test completion date:	2 February 2001
Project Members:	J L Slabbert and N Ramgopaul
Project Leader	J L Slabbert

Signature : …… ……
JL Slabbert, Head of Laboratory **Date**

1. SAMPLE INFORMATION

1. INTRODUCTION

Electrodialysis was investigated for the removal of pollutants from hazardous waste with the aim to reduce the volumes of waste dams. Our laboratory was requested to assist with a toxicity evaluation of untreated and treated waste samples. The waste (Sample 1) used in the study was collected from Dam 4 at the Holfontein Waste Site. The waste was pre-treated (Sample 2) by adding Iscor ash (200 g/ℓ), soda ash (12 g/ℓ) and caustic soda (7 g/ℓ) to the waste and by bubbling the mixture with carbon dioxide. The pre-treated sample was desalinated (Sample 3) using electro dialysis. During the process, brine (Sample 4) is generated which will, during full-scale operation, be returned to the landfill. All four samples were very dark in colour and had a distinct creosote-like odour.

Moderately hard reconstituted water was used for control testing and dilution of the waste samples (Table 1) (Slabbert *et al.,* 1998).

TABLE E1: Moderately hard reconstituted water[1]

Reagent added[2] (mg/ℓ)	$NaHCO_3$	96.0
	$CaSO_4.2H_2O$	60.0
	$MgSO_4$	60.0
	KCl	4.0
Nominal water quality range	pH	7.4 - 7.8
	Hardness[3]	80 - 100
	Alkalinity	60 - 70

[1]US EPA (1985) [2]Prepared with Milli-Q water
[3]As mg/ℓ $CaCO_3$

2. *DAPHNIA* TOXICITY TEST

Toxicity was established by means of a 48-h *Daphnia pulex* lethality test. The test has been recommended for water and effluent toxicity testing in South Africa and was carried out according to procedures established at the Environmentek laboratories and described in the **Guidelines for Toxicity Bioassaying of Waters and Effluents in South Africa** (Slabbert *et al.*, 1998).

Organisms 24 h or less in age were used for toxicity testing. To obtain the necessary number of young for a test, adult females bearing embryos in their brood pouches were removed from the stock cultures 24 h preceding the initiation of a test and placed in beakers containing moderately hard water (Table E1) and food suspension (trout chow, alfalfa and yeast). Test conditions are summarized in Table E2. Test organisms were transferred to a small intermediary holding beaker and from there to the test beakers.

3. STATISTICAL ANALYSIS

A probit computerized statistical programme (US EPA, 1985) was applied to test data to calculate the *Daphnia* LC_{50} (concentration at which 50% of the organisms died) and 95% confidence limits, and the LC_{10} (minimum effect concentration). The LC_0 (no effect concentration) was derived from the concentrations tested.

TABLE E2: *Daphnia* test conditions[1]

Temperature	20±1°C
Light quality	Laboratory illumination
Photoperiod	Approximately 14 h day light
Feeding regime	No feeding
Oxygen concentration	As obtained
pH	As obtained
Size of test beaker	50 mℓ
Volume of test sample	25mℓ
Number of organisms/beaker	5
Number of replicate beakers	4
Total number of organisms/test	20
Test duration	48 h
Effect measured	Lethality (no movement of body or appendages on gentle prodding)
Interpretation of results	Lethality ≥10% is an indication of toxicity, provided that control lethality <10%

[1] According to US EPA (1985) procedure

4. RESULTS AND DISCUSSION

The results of definitive tests (testing serial dilutions) on the waste samples are summarized in Table E4. Sample 4 (brine) exhibited the highest toxicity, followed in order of magnitude by sample 2 (pre-treated), sample 1 (untreated) and sample 3 (treated). The pH's of all the samples were within the limits required for the sustenance of aquatic life. The oxygen concentration of sample 2 was below the required limit of 2.9 mg/ℓ (40% of 7.2 mg/ℓ) at concentrations <2.5%. The low oxygen level could, therefore, have contributed to the adverse effects at the upper test concentrations. Table E3 presents the LC_0, LC_{10} and LC_{50} values for the wastes.

TABLE E3: LC_0, LC_{10} and LC_{50} values (%) for the tested wastes

Sample	LC_0	LC_{10}	LC_{50}	95% confidence limits	
				Lower limit	Upper limit
1	0.31	0.92	1.72	1.50	2.03
2	<0.63	0.67	1.12	0.92	1.34
3	6.26	9.84	19.23	15.59	23.78
4	0.125	0.33	0.61	0.50	0.75

Toxicity involves an inverse relationship to effect concentrations (the higher the toxicity, the lower the effect concentration). Concentration-based toxicity measurements can be translated into toxic units (TU's) to enable a linear increase between effect concentration and toxicity (Slabbert e*t al*. (1998). One common method of deriving toxic units is to divide the full strength effluent/waste (100%) by the concentration that causes acute or chronic toxicity. The acute toxic units (TU_a's =100/$LC_{50)}$ in Table E5 clearly show how the different samples differ in toxicity (the larger the value, the higher the toxicity).

TABLE E4: Effect of serial dilutions of waste on *Daphnia*

Sample 1					Sample 2					Sample 3					Sample 4				
Concen-tration (%)	pH	Oxygen concen-tration (mg/ℓ)	% Lethality after time:		Concen-tration (%)	pH	Oxygen concen-tration (mg/ℓ)	% Lethality after time:		Concen-tration (%)	pH	Oxygen concen-tration (mg/ℓ)	% Lethality after time:		Concen-tration (%)	pH	Oxygen concen-tration (mg/ℓ)	% Lethality after time:	
			24 h	48 h				24 h	48 h				24 h	48 h				24 h	48 h
10	7.6	9.2	100	100	10	8.0	1.3	100	100	50	7.7	4.2	ns	100	2	8.7	9.2	100	100
5	nd	nd	100	100	5	nd	nd	100	100	25	nd	nd	ns	65	1	nd	nd	50	85
2.5	8.6	7.9	15	85	2.5	8.0	3.9	65	100	12.5	8.2	4.9	ns	15	0.5	8.4	9.7	0	30
1.25	nd	nd	0	70	1.25	nd	nd	20	55	6.25	nd	nd	ns	5	0.25	nd	nd	0	5
0.63	8.3	9.6	0	10	0.63	8.2	7.3	0	10	nd	nd	nd	nd	nd	0.125	7.8	9.8	0	0
0.31	nd	nd	0	0	nd	nd	nd	nd	nd	nd	nd	nd	nd	nd	0.063	nd	nd	0	0
0.16	8.1	9.7	0	0	nd	nd	nd	nd	nd	nd	nd	nd	nd	nd	nd	nd	nd	nd	nd

nd Not determined
ns Not scored at 24 h due to the dark colour. The sample preparations were diluted at 48 h to enable counting of *Daphnia*

TABLE E5: Calculated TU_a's for the wastes

Sample	TU_a
1	58
2	89
3	5
4	164

The LC_0 values in Table E3 indicate the dilution required to avoid acute effects on aquatic systems (e.g. a 16-times dilution in case of sample 3 and a 800-times dilution in case of sample 4).

REFERENCES

Slabbert, J.L., J. Oosthuizen, E.A. Venter, E. Hill, M du Preez and P.J. Pretorius. 1998. Development of guidelines for toxicity bioassaying of drinking and environmental waters in South Africa. WRC Report No 358/1/98. Report to the Water Research Commission by the Division of Water, Environment and Forestry Technology, CSIR, Pretoria.

US EPA. 1985. Methods for measuring the acute toxicity of effluents to freshwater and marine organisms. EPA/600/4-85/013, Environmental and Support Laboratory, Office of Research and Development, US Environmental Protection Agency, Cincinnati, Ohio.

APPENDIX F

FEED-AND-BLEED ED TESTS

Feed and Bleed Run 100%

Run 4 Pretreated Sample 14/01/2001; 200 g/ℓ ISCOR ash, 12 g/ℓ soda ash, 6.9 g NaOH) 13/02/2001

100% 96.7 Active membrane area 204cm^3

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resis	Cur density
(min)	(V)	(V)	(Ampere)	mS/cm	Deg C	pH	(mS/cm)	Deg C	pH	(mS/cm)		pH	(ohm)	(ohm)
0	33,17	22,11	8	88,7	25,3	7,75	121,1	25,6	8,08	88	8,05	4,15	2,76	39,22
30	32,14	21,09	8	83,5	27,9		127,3	28,5				4,02	2,64	39,22
60	31,32	20,34	8	80,6	29,7		131,5	30,6				3,92	2,54	39,22
90	30,42	20,12	8	77,9	31,3		135,5	32				3,80	2,52	39,22
120	29,91	19,78	8	76,9	3 232		138,2	33				3,74	2,47	39,22
150	29,44	19,35	8	77,1	32,5		140,8	33,7				3,68	2,42	39,22
180	29,11	19,11	8	77	32,9	8,21	142,5	34	8,04			3,64	2,39	39,22

30,787 20,27

in out

Feed flow 618 564mℓ/min 54

Brine flow 53 114mℓ/min -61

Recirculation flow 300ℓ/h

Feed and Bleed 80%

Run 5 (Pretreated diluted with tap water, Sample 14/01/2001; 200 g/ℓ ISCOR ash, 12 g/ℓ soda ash, 6.9 g NaOH) 14/02/2001

80% 77mS/cm Active membrane area 204cm3

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resis	Cur density
(min)	(V)	(V)	(Ampere)	(mS/cm)	°C	pH	(mS/cm)	°C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm2)
0	34,19	25,05	8	68,7	25,4	7,56	122,8	25,7	8,08	88,9	8,05	4,27	3,13	39,22
30	33,8	23,29	8	62,7	28,2		127,2	28,9				4,23	2,91	39,22
60	33,01	22,67	8	58,6	30,2		130,4	30,7				4,13	2,83	39,22
90	32,32	21,97	8	56,8	31,4		134,7	32,2				4,04	2,75	39,22
120	31,84	21,73	8	55	32,3		136,5	33,1				3,98	2,72	39,22
150	31,38	21,36	8	54,1	32,9		137,7	33,7				3,92	2,67	39,22
180	31,12	21,15	8	53,9	33,1		139,2	34				3,89	2,64	39,22
210	30,45	20,86	8	53,8	33,7	8,51	139,6	34,5	8,48			3,81	2,61	39,22

32,26 22,26

in out

Feed flow 618 570mℓ/min 48

Brine flow 52 108mℓ/min -56

Recirculation flow 300ℓ/h

Feed and Bleed 56%

Run 6 (Pretreated diluted with tap water, Sample 19/02/2001; 200 g/ℓ ISCOR ash, 12 g/ℓ soda ash, 6.9 g NaOH) 23/02/2001

56% Feed 94.9 54mS/cm Active membrane area 204cm3

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resis	Cur density
(min)	(V)	(V)	(Ampere)	(mS/cm)	°C	pH	(mS/cm)	°C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm2)
0	34,45	24,29	8	48,9	23,6	7,38	120,9	23,7		88,2	8,12	4,31	3,04	39,22
30	35,18	24,99	8	36,3	29		128,7	29,5				4,40	3,12	39,22
60	34,68	25,13	8	32,5	30,8		129,8	31,7				4,34	3,14	39,22
90	35,56	25,63	8	30,7	31,7		130,1	32,5				4,45	3,20	39,22
120	35,11	25,75	8	28,5	32,7		130,8	33,1				4,39	3,22	39,22
150	35,72	25,84	8	27	33,5		130,9	33,9				4,47	3,23	39,22
180	35,55	25,89	8	26,4	33,9		131,7	34,5				4,44	3,24	39,22
210	35,59	25,83	8	26,2	34,2		131,7	34,9				4,45	3,23	39,22
240	35,42	25,75	8	26,4	34,4	7,76	131,5	35,1	7,86			4,43	3,22	39,22

35,25 25,46

in out

Feed flow 606 552mℓ/min 54

Brine flow 52 112mℓ/min -60

Recirculation flow 300ℓ/h

Feed and Bleed 28% Repeat run

27% Feed 98,7 26mS/cm Active membrane area 204cm3

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resis	Cur density
(min)	(V)	(V)	(Ampere)	(mS/cm)	°C	pH	(mS/cm)	°C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm2)
0	50,01	38	6,84	22,1	13,2	7,8	102,4	13,2	7,69	82,4	7,76	7,31	5,56	33,53
30	50,06	37,97	6,05	13,6	16,7	7,6	107,2	17,1	7,68			8,27	6,28	29,66
60	50,02	39,6	4,59	6,9	19,8	7,2	107	20,1	7,71			10,90	8,63	22,50
90	49,99	39,68	4,4	6,4	21,6	7,1	105,9	22	7,76			11,36	9,02	21,57
120	50,08	39,87	4,25	6	22,5	7,1	105	23	7,79			11,78	9,38	20,83
150	50,05	36,69	4,31	6,2	23,4	7,1	104,5	23,7	7,84			11,61	8,51	21,13
180	50,11	40,08	4,33	6,2	23,9	7,2	103,9	24,3	7,87			11,57	9,26	21,23
210	50,13	39,65	4,3	6,2	24,4	7,2	103,3	24,8	7,91			11,66	9,22	21,08
	50,06	38,94												

in out

Feed flow 594 552mℓ/min 42

Brine flow 50 86mℓ/min -36

Recirculation flow 600ℓ/h

Feed and Bleed 28%

Run 7 (Pretreated diluted with tap water, Sample 19/02/2001; 200 g/ℓ ISCOR ash, 12 g/ℓ soda ash, 6.9 g NaOH)

23/02/2001

27% Feed 94,9 26mS/cm Active membrane area 204cm3

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resis	Cur density
(min)	(V)	(V)	(Ampere)	(mS/cm)	°C	pH	(mS/cm)	°C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm2)
0	25,8	19,39	3,3	13,6	22,5	8	95,6	22,8		81,7	8,55	7,82	5,88	16,18
30	26,56	19,99	3,31	12,3	25,1		98,7	25,4				8,02	6,04	16,23
60	26,46	19,85	3,3	11,8	27,3		97,5	27,4				8,02	6,02	16,18
90	26,5	19,81	3,31	11,9	27,8		97,1	28,4				8,01	5,98	16,23
120	26,18	19,49	3,31	11,8	28,9	7,8	97	29	8,38			7,91	5,89	16,23
	26,30	19,71												

	in	out		
Feed flow	600	570	mℓ/min	30
Brine flow	50	76	mℓ/min	-26
Recirculation flow	600ℓ/h			

Feed and Bleed 12 % Repeat run

Run 8 Repeat (Pretreated diluted with tap water, Sample 05/07/2001; 200 g/ℓ ISCOR ash, 12 g/ℓ soda ash, 6,9 g NaOH)

12/07/2001

6% Feed 98,7 6,22mS/cm Active membrane area 204cm3

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resis	Cur density
(min)	(V)	(V)	(Ampere)	(mS/cm)	°C	pH	(mS/cm)	°C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm2)
0	30,12	22,68	2,47	5,9	16,8	7,97	86,2	17,1	8,15	75,5	6,63	12,19	9,18	12,11
30	30,05	23,1	1,62	3,29	19,1	7,66	84,7	19,3	8,12			18,55	14,26	7,94
60	30,65	23,65	1,29	2,29	20,4	7,38	82,5	20,6	8,13			23,76	18,33	6,32
90	30,75	22,25	1,15	1,59	21,5	7,09	79,8	21,7	8,14			26,74	19,35	5,64
120	30,38	22,36	1,16	2,22	22,2	7,25	76	22,5	8,16			26,19	19,28	5,69
150	30,1	22,41	1,15	2,38	22,1	7,34	74	22,4	8,16			26,17	19,49	5,64
180	29,98	22,54	1,16	2,47	22,2	7,38	71,8	22,6	8,18			25,84	19,43	5,69
210	29,99	22,61	1,15	2,4	22,3	7,37	69,8	22,7	8,19			26,08	19,66	5,64
240	29,9	22,79	1,16	2,37	22,4	7,33	67,9	22,8	8,2			25,78	19,65	5,69
270	30,01	22,68	1,16	2,33	22,8	7,35	66,2	23	8,2			25,87	19,55	5,69
	30,19	22,71												

	in	out		
Feed flow	594	582	mℓ/min	12
Brine flow	50	56	mℓ/min	-6
Recirculation flow	600ℓ/h			

Feed Total in	633,65mℓ/min
Brine Total out	56mℓ/min
	91,16231%

Feed and Bleed 12%

Run 8 (Pretreated diluted with tap water, Sample 19/02/2001; 200 g/ℓ ISCOR ash, 12 g/ℓ soda ash, 6,9 g NaOH)

06/03/2001

12% Feed 94,9 11,8mS/cm Active membrane area 204cm3

Time	Total V	Volt (75 cp)	Current	Feed	Feed	Feed	Brine	Brine	Brine	Electrode Rinse		Cell Resist	75 CP Resis	Cur density
(min)	(V)	(V)	(Ampere)	(mS/cm)	°C	pH	(mS/cm)	°C	pH	(mS/cm)	pH	(ohm)	(ohm)	(mA/cm2)
0	45,28	36,92	3,3	6	22,4	8,09	98,8	22,8	8,38	86,2	8,44	13,72	11,19	16,18
30	46,54	37,33	3,28	3,5	26,5		96,8	26,8				14,19	11,38	16,08
60	44,25	36,13	2,81	2,8	28,3		93,6	28,3				15,75	12,86	13,77
90	44	36,13	2,2	2,2	29,9		91,7	30				20,00	16,42	10,78
120	44	36,65	2,33	2,35	30,5		90,5	30,7		87,1		18,88	15,73	11,42
150	44,18	37,15	2,45	2,25	31,3		89	31,3				18,03	15,16	12,01
180	44,09	36,77	2,38	2,3	31		87,3	31,9				18,53	15,45	11,67
210	44,11	36,47	2,34	2,35	32		85,4	32,3				18,85	15,59	11,47
240	44,15	36,67	2,35	2,35	32,1	7,07	84,5	32,6	8,38			18,79	15,60	11,52
	44.51	36.69												

	in	out		
Feed flow	606	588	mℓ/min	18
Brine flow	50	74	mℓ/min	-24
Recirculation flow		600ℓ/h		

Feed Total in 3048mℓ/min

Brine Total out 484mℓ/min

84,12073%

Feed and Bleed Calculations100%

Feed water (in):			Feed water (out):		
Flow rate (litre/min):	0,618		Flow rate (litre/min):	0,564	
Concentration (mg/ℓ)		Mass transfer (mg/min)	Concentration (mg/ℓ)		Mass transfer (mg/min)
Sodium	30 500	18 849,00	Sodium	23 700	13 366,80
Chloride	26 100	16129.80	Chloride	13518	7624.15
Calcium	19	11,74	Calcium	20	11,28
Magnesium	33	20,39	Magnesium	30	16,92
Sulphate	17 436	10 775,45	Sulphate	16 756	9 450,38
Ammonia-N	873	539,51	Ammonia-N	666	375,62
Potassium	6 350	3 924,30	Potassium	3 470	1 957,08
TOC	10 723	6 626,81	TOC	10 339	5 831,20
TDS	116 235	71 833,23	TDS	92 000	51 888,00
Alkalinity	19 348	11 957,06	Alkalinity	18 666	10 527,62
pH	7,75		pH	8,21	
mS/cm	96,7		mS/cm	77	
Ratio: TDS/mS/m	12,02		Ratio: TDS/mS/m	11,95	
			Delta C (TDS)	58 845,00	
Brine water (out):			Brine water (in):		
Flow rate (litre/min):	0,114		Flow rate (litre/min):	0,053	0,061

Concentration (mg/ℓ)		Mass transfer (mg/min)	Feed (in) - (out)	Difference	Cat(ge/min)	An (ge/min)
Sodium	47 700	5 437,80	5 482,20	-44,40	0,23836	
Chloride	55 530	6 330,42	8 505,65	-2 175,23		0,23960
Calcium	41	4,67	0,46	4,21	0,00002	
Magnesium	60	6,84	3,47	3,37	0,00029	
Sulphate	16 635	1 896,39	1 325,06	571,33		0,02761
Ammonia-N	1 434	163,48	163,89	-0,41	0,01171	
Potassium	11 200	1 276,80	1 967,22	-690,42	0,05044	
TOC	8 291	945,17	795,62	149,56		
TDS	150 845	17 196,33	19 945,23	-2748,90		
Alkalinity	19 861	2 264,15	1 429,44	834,71		0,02859
Ph	8,04			TOTAL ge removed	0,30082	0,29579
mS/cm	142,5			**%CE**	**83,04**	**81,65**
Ratio: TDS/mS/cm	10,59					

Total Voltage across		30,8V
Voltage across 75 cp	Voltage	20,3V
	Current	8A
	Time	3h

Loading (TDS)	19,95g/min		membrane area	1,7m^2
	1 196,71g/h		**Loading**	**703,95g/h.m^2**
Water transfer	54g/min		TDS	1 196,71g/h
	3 240g/h		**Water/TDS**	**2,71g/g**
Energy/kℓ	3h		Flow rate	0,618ℓ/min
	8A		Volume	0,11124kℓ
kWh	0,4872	20,3V	**Energy**	**4,38kWh/kℓ**

Feed and Bleed calculations 80%

Run 5

Feed water (in):			Feed water (out):		
Flow rate (litre/min):	0,618		Flow rate (litre/min):	0,57	
Concentration (mg/ℓ)		Mass transfer (mg/min)	Concentration (mg/ℓ)		Mass transfer (mg/min)
Sodium	22 100	13 657,80	Sodium	16 100	9 177,00
Chloride	19 550	12 081,90	Chloride	11 014	6 277,98
Calcium	16	9,89	Calcium	12	6,84
Magnesium	13	8,03	Magnesium	10	5,70
Sulphate	12 111	7 484,60	Sulphate	10 087	5 749,59
Ammonia-N	868	536,42	Ammonia-N	571	325,47
Potassium	3 950	2 441,10	Potassium	2 450	1 396,50
TOC	6 434	3 976,21	TOC	7 133	4 065,81
TDS	79 885	49 368,93	TDS	58 485	33 336,45
Alkalinity	14 907	9 212,53	Alkalinity	13 338	7 602,66
pH	7,56		pH	8,51	
mS/cm	77		mS/cm	53,8	
Ratio: TDS/mS/m	10,37		Ratio: TDS/mS/m	10,87	
			Delta C (TDS)	**80 350,00**	
Brine water (out):			Brine water (in):		
Flow rate (litre/min):	0,108		Flow rate (litre/min):	0,052	0,056

Concentration (mg/ℓ)		Mass transfer (mg/min)	Feed (in) - (out)	Difference	Cat(ge/min)	An (ge/min)
Sodium	40 200	4 341,60	4 480,80	-139,20	0.19482	
Chloride	48 120	5 196,96	5 803,92	-606,96		0,16349
Calcium	35	3,78	3,05	0,73	0,00015	
Magnesium	48	5,18	2,33	2,85	0,00019	
Sulphate	16 699	1 803,49	1 735,01	68,48		0,03615
Ammonia-N	1 561	168,59	210,95	-42,37	0,01507	
Potassium	9 760	1 054,08	1 044,60	9,48	0,02678	
TOC	7 712	832,90	-89,60	922,49		
TDS	138 835	14 994,18	16 032,48	-1 038,30		
Alkalinity	17 076	1 844,21	1 609,87	234,34		0,03220
pH	8,48			TOTAL ge removed	0,23702	0,23183
mS/cm	139,6			**%CE**	**65,43**	**64,00**
Ratio: TDS/mS/cm	9,95					

Total Voltage across		32,3V
Voltage across 75 cp	Voltage	22,3V
	Current	8A
	Time	3,5h

Loading (TDS)	16,03g/min	membrane area	1,7m^2
	961,95g/h	**Loading**	**565,85g/h.m^2**
Water transfer	48g/min	TDS	961,95g/h
	2 880g/h	**Water/TDS**	**2,99g/g**
Energy/kℓ	3,5h	Flow rate	0,618ℓ/min
	8A	Volume	0,12978kℓ
kWh 0,6244	22,3V	**Energy**	**4,81kWh/kℓ**

Feed and Bleed 56%

Run 6

Feed water (in):			Feed water (out):		
Flow rate (litre/min):	0,606		Flow rate (litre/min):	0,552	
Concentration (mg/ℓ)		Mass transfer (mg/min)	Concentration (mg/ℓ)		Mass transfer (mg/min)
Sodium	14 300	8 665,80	Sodium	6 380	3 521,76
Chloride	11 731	7 108,99	Chloride	3 588	1 980,58
Calcium	30	18,18	Calcium	9	4,97
Magnesium	55	33,33	Magnesium	15	8,28
Sulphate	7 290	4 417,74	Sulphate	2 898	1 599,70
Ammonia-N	595	360,57	Ammonia-N	250	138,00
Potassium	2 740	1 660,44	Potassium	874	482,45
TOC	2 300	1 393,80	TOC	2976	1642,75
TDS	51600	31 269,60	TDS	26 310	14 523,12
Alkalinity	9 337	5 658,22	Alkalinity	6 625	3 657,00
pH	7,38		pH	7,76	
mS/cm	54		mS/cm	26,4	
Ratio: TDS/mS/m	9,56		Ratio: TDS/mS/m	9,97	
			Delta C (TDS)	113 835,00	
Brine water (out):			Brine water (in):		
Flow rate (litre/min):	0,112		Flow rate (litre/min):	0,052	0,06

Concentration (mg/l)		Mass transfer (mg/min)	Feed (in) - (out)	Difference	Cat(ge/min)	An (ge/min)
Sodium	44 200	4 950,40	5 144,04	-193,64	0,22365	
Chloride	44 740	5 010,88	5 128,41	-117,53		0,14446
Calcium	60	6,72	13,21	-6,49	0,00066	
Magnesium	123	13,78	25,05	-11,27	0,00209	
Sulphate	20 973	2 348,98	2 818,04	-469,07		0,05871
Ammonia-N	1 617	181,10	222,57	-41,47	0,01590	
Potassium	9 210	1 031,52	1 177,99	-146,47	0,03020	
TOC	5 825	652,40	-248,95	901,35		
TDS	140 145	15 696,24	16,746,48	-1 050,24		
Alkalinity	18 893	2 116,02	2 001,22	114,79		0,04002
pH	7,86			TOTAL ge removed	0,27250	0,24320
mS/cm	131,5			**%CE**	**73,76**	**65,83**
Ratio: TDS/mS/cm	10,66					

Total Voltage across		35,3V
Voltage across 75 cp	Voltage	25,5V
	Current	8A
	Time	6h

Loading (TDS)	16,75g/min	membrane area	1,7m^2
	1004,79g/h	**Loading**	**591,05g/h.m^2**
Water transfer	54g/min	TDS	1004,79g/h
	3 240g/h	**Water/TDS**	**3,22g/g**
Energy/kℓ	6h	Flow rate	0,606ℓ/min
	8A	Volume	0,21816kℓ
kWh 1,224	25,5V	**Energy**	**5,61kWh/kℓ**

Feed and Bleed 28%

Run 7 Repeat

Feed water (in):			Feed water (out):		
Flow rate (litre/min):	0,594		Flow rate (litre/min):	0,552	
Concentration (mg/ℓ)		Mass transfer (mg/min)	Concentration (mg/ℓ)		Mass transfer (mg/min)
Sodium	5 920	3 552,00	Sodium	1 500	855,00
Chloride	4 483	2 689,80	Chloride	595	339,15
Calcium	27,4	16,44	Calcium	5,56	3,17
Magnesium	74,5	44,70	Magnesium	8,68	4,95
Sulphate	3 753	2 251,80	Sulphate	1 296	738,72
Ammonia-N	188	112,80	Ammonia-N	38,3	21,83
Potassium	979	587,40	Potassium	178	101,46
TOC	2 861	1 716,60	TOC	1 045	595,65
TDS	22 895	13 737,00	TDS	6 765	3 856,05
Alkalinity	4 398	2 638,80	Alkalinity	1 384	788,88
pH			pH		
mS/cm			mS/cm		
Ratio: TDS/mS/m	#DIV/0!		Ratio: TDS/mS/m	#DIV/0!	
			Delta C (TDS)	104 595,00	
Brine water (out):			Brine water (in):		
Flow rate (litre/min):	0,086		Flow rate (litre/min):	0,05	0,036

Concentration (mg/ℓ)		Mass transfer (mg/min)	Feed (in) - (out)	Difference	Cat(ge/min)	An (ge/min)
Sodium	30 700	2 333,20	2 697,00	-363,80	0,11726	
Chloride	27 700	2 105,20	2 350,65	-245,45		0,06622
Calcium	117	8,89	13,27	-4,38	0,00066	
Magnesium	331	25,16	39,75	-14,60	0,00331	
Sulphate	15 943	1 211,67	1 513,08	-301,41		0,03152
Ammonia-N	652	49,55	90,97	-41,42	0,00650	
Potassium	5 890	447,64	485,94	-38,30	0,01246	
TOC	7 753	589,23	1 120,95	-531,72		
TDS	111 360	8 463,36	9 880,95	-1 417,59		
Alkalinity	18 014	1 369,06	1 849,92	-480,86		0,03700
pH				TOTAL ge removed	0,14019	0,13474
mS/cm				**%CE**	**69,20**	**66,51**
Ratio: TDS/mS/cm	#DIV/0!					

Total Voltage across		50,06
Voltage across 75 cp	Voltage	38,94
	Current	4,3
	Time	3,5

Loading (TDS)	9,88g/min		membrane area	1,7m^2
	592,86g/h		**Loading**	**348,74g/h.m^2**
Water transfer	42g/min		TDS	592,86g/h
	2 520g/h		**Water/TDS**	**4,25g/g**
Energy/kℓ	3,5h		Flow rate	0,594ℓ/min
	4,3A		Volume	0,12474kℓ
kWh	0,586047	38,94V	**Energy**	**4,70kWh/kℓ**

Run 7

Feed water (in):			Feed water (out):		
Flow rate (litre/min):	0,6		Flow rate (litre/min):	0,57	
Concentration (mg/ℓ)		Mass transfer (mg/min)	Concentration (mg/ℓ)		Mass transfer (mg/min)
Sodium	5 940	3 564,00	Sodium	2 690	1 533,30
Chloride	4 985	2 991,00	Chloride	1 764	1 005,48
Calcium	27	16,20	Calcium	5	2,85
Magnesium	30	18,00	Magnesium	5	2,85
Sulphate	3 119	1 871,40	Sulphate	1 269	723,33
Ammonia-N	219	131,40	Ammonia-N	88,5	50,45
Potassium	1 080	648,00	Potassium	377	214,89
TOC	1 581	948,60	TOC	614,4	350,21
TDS	23 360	14 016,00	TDS	10 360	5 905,20
Alkalinity	4 353	2 611,80	Alkalinity	2 675	1 524,75
pH	8,03		pH	7,81	
mS/cm	26		mS/cm	11,8	
Ratio: TDS/mS/m	8,98		Ratio: TDS/mS/m	8,78	
			Delta C (TDS)	75 765,00	
Brine water (out):			Brine water (in):		
Flow rate (litre/min):	0,076		Flow rate (litre/min):	0,05	0,026

Concentration (mg/ℓ)		Mass transfer (mg/min)	Feed (in) - (out)	Difference	Cat(ge/min)	An (ge/min)
Sodium	24 300	1 846,80	2 030,70	-183,90	0,08829	
Chloride	27 260	2 071,76	1 985,52	86,24		0,05593
Calcium	39	2,96	13,35	-10,39	0,00067	
Magnesium	128	9,73	15,15	-5,42	0,00126	
Sulphate	14 665	1 114,54	1 148,07	-33,53		0,02392
Ammonia-N	1 014	77,06	80,96	-3,89	0,00578	
Potassium	5 500	418,00	433,11	-15,11	0,01111	
TOC	6 238	474,09	598,39	-124,30		
TDS	86 125	6 545,50	8 110,80	-1 565,30		
Alkalinity	12 891	979,72	1 087,05	-107,33		0,02174
pH	8,38			TOTAL ge removed	0,10711	0,10159
mS/cm	97			**%CE**	**69,59**	**66,00**
Ratio: TDS/mS/cm	8,88					

Total Voltage across		26,3V
Voltage across 75 cp	Voltage	19,7V
	Current	3,3A
	Time	2h

Loading (TDS)		8,11g/min	membrane area	1,7m^2
		486,65g/h	**Loading**	**286,26g/h.m^2**
Water transfer		30g/min	TDS	486,65g/h
		1 800g/h	**Water/TDS**	**3,70g/g**
Energy/kℓ		2h	Flow rate	0,6ℓ/min
		3,3A	Volume	0,072kℓ
kWh	0,13002	19,7V	**Energy**	**1,81kWh/kℓ**

Feed and Bleed 12%

Run 8 Repeat

Feed water (in):			Feed water (out):		
Flow rate (litre/min):	0,606		Flow rate (litre/min):	0,588	
Concentration (mg/ℓ)		Mass transfer (mg/min)	Concentration (mg/ℓ)		Mass transfer (mg/min)
Sodium	1 240	751,44	Sodium	515	302,82
Chloride	929	562,97	Chloride	276	162,29
Calcium	22,6	13,70	Calcium	3,31	1,95
Magnesium	24,1	14,60	Magnesium	3,01	1,77
Sulphate	757	458,74	Sulphate	324	190,51
Ammonia-N	39,5	23,94	Ammonia-N	13,05	7,67
Potassium	208	126,05	Potassium	72	42,34
TOC	537	325,42	TOC	328	192,86
TDS	4 701	2 848,81	TDS	1 990	1 170,12
Alkalinity		0,00	Alkalinity	700	
pH			pH	7,07	
mS/cm	11,8		mS/cm	2,35	
Ratio: TDS/mS/m	3,98		Ratio: TDS/mS/m	8,47	
			Delta C (TDS)	59 600,00	
Brine water (out):			Brine water (in):		
Flow rate (litre/min):	0,074		Flow rate (litre/min):	0,05	0,024

Concentration (mg/ℓ)		Mass transfer (mg/min)	Feed (in) - (out)	Difference	Cat(ge/min)	An (ge/min)
Sodium	16 300	1 206,20	448,62	757,58	0,01951	
Chloride	16 369	1 211,31	400,69	810,62		0,01129
Calcium	76,6	5,67	11,75	-6,08	0,00059	
Magnesium	248	18,35	12,83	5,52	0,00107	
Sulphate	8 983	664,74	268,23	396,51		0,00559
Ammonia-N	531	39,29	16,26	23,03	0,00116	
Potassium	3 060	226,44	83,71	142,73	0,00215	
TOC	4 085	302,29	132,56	169,73		
TDS	61 590	4 557.66	1 678,69	2 878,97		
Alkalinity		0,00	0,00	0,00		0,00000
pH				TOTAL ge removed	0,02447	0,01688
mS/cm	84,5			**%CE**	**45,68**	**31,50**
Ratio: TDS/mS/cm	7,29					

Total Voltage across		30,19V	
Voltage across 75 cp	Voltage	22,71V	
	Current	1,16A	
	Time	4h	

Loading (TDS)	1,68g/min		membrane area	1,7m^2
	100,72g/h		**Loading**	**59,25g/h.m^2**
Water transfer	18g/min		TDS	100,72g/h
	1 080g/h		**Water/TDS**	**10,72g/g**
Energy/kℓ	4h		Flow rate	0,606ℓ/min
	1,16A		Volume	0,14544kℓ
kWh	0,105374	22,71V	**Energy**	**0,72kWh/kℓ**

Run 8

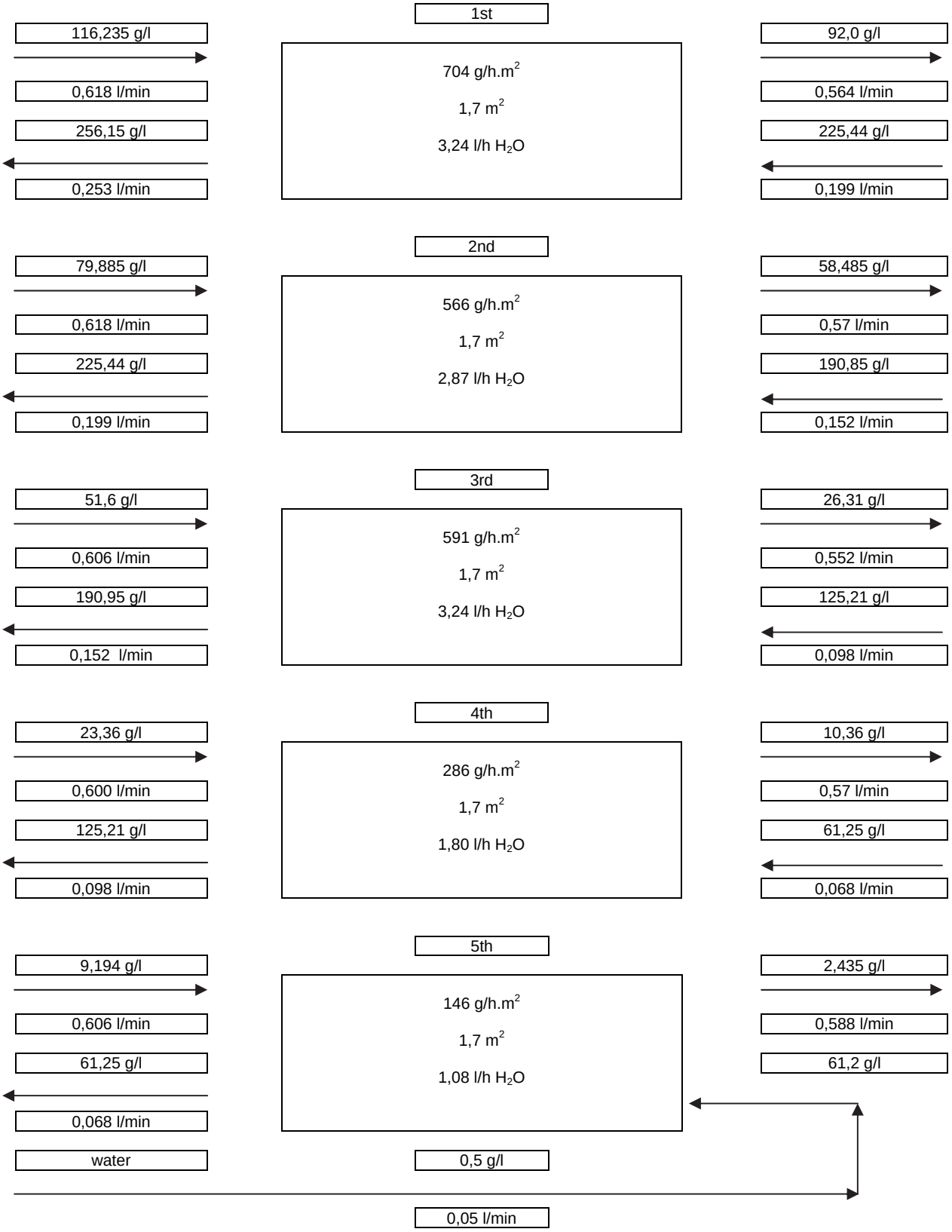
Feed water (in):			Feed water (out):		
Flow rate (litre/min):	0,606		Flow rate (litre/min):	0,588	
Concentration (mg/ℓ)		Mass transfer (mg/min)	Concentration (mg/ℓ)		Mass transfer (mg/min)
Sodium	2 650	1 605,90	Sodium	606	356,33
Chloride	2 283	1 383,50	Chloride	266	156,41
Calcium	26	15,76	Calcium	3	1,76
Magnesium	20	12,12	Magnesium	2	1,18
Sulphate	1 470	890,82	Sulphate	460	270,48
Ammonia-N	93,3	56,54	Ammonia-N	18,4	10,82
Potassium	455	275,73	Potassium	87	51,16
TOC	965	584,79	TOC	363	213,44
TDS	9 194	5 571,56	TDS	2 435	1 431,78
Alkalinity	1 700	1 030,20	Alkalinity	409	240,49
pH	8,09		pH	7,07	
mS/cm	11,8		mS/cm	2,35	
Ratio: TDS/mS/m	7,79		Ratio: TDS/mS/m	10,36	
			Delta C (TDS)	78 035,00	
Brine water (out):			Brine water (in):		
Flow rate (litre/min):	0,074		Flow rate (litre/min):	0,05	0,024

Concentration (mg/ℓ)		Mass transfer (mg/min)	Feed (in) - (out)	Difference	Cat(ge/min)	An (ge/min)
Sodium	24 400	1 805,60	1 249,57	556,03	0,05433	
Chloride	26 373	1 951,60	1 227,09	724,51		0,03457
Calcium	54	4,00	13,99	-10,00	0,00070	
Magnesium	181	13,39	10,94	2,45	0,00091	
Sulphate	12 643	935,58	620,34	315,24		0,01292
Ammonia-N	813	60,16	45,72	14,44	0,00327	
Potassium	4 500	333,00	224,57	108,43	0,00576	
TOC	5 426	401,52	371,35	30,18		
TDS	80 470	5 954,78	4 139,78	1 815,00		
Alkalinity	12 173	900,80	789,71	111,09		0,01579
pH	8,38			TOTAL ge removed	0,06496	0,06328
mS/cm	84,5			**%CE**	**59,86**	**58,31**
Ratio: TDS/mS/cm	9,52					

Total Voltage across		44,5V
Voltage across 75 cp	Voltage	36,7V
	Current	2,35A
	Time	6h

Loading (TDS)	4,14g/min		membrane area	1,7m^2
	248,39g/h		**Loading**	**146,11g/h.m^2**
Water transfer	18g/min		TDS	248,39g/h
	1 080g/h		**Water/TDS**	**4,35g/g**
Energy/kℓ	6h		Flow rate	0,606ℓ/min
	2,35A		Volume	0,21816kℓ
kWh 0,51747	36,7V		**Energy**	**2,37kWh/kℓ**

Feed and brine concentration flows through a 5-stage ED plant.



APPENDIX G

CHEMICAL COMPOSITION OF MUNICIPAL SOLID WASTE LEACHATES

Bisasar Road (A)

Leachate Header Tank

Date	20/8/96	14/1/97	10/7/97	4/12/97	20/1/98	20/2/98	19/3/98	29/4/98	20/5/98	3/6/98	18/6/98	6/10/98	17/12/98	19/7/99	14/12/99	5/7/00	21/1200
Level	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
pH	7	7.5	7.73	7.55	7.9	7.9	8	8.1	7.6	7.65	7.6	7.8	7.65	7.55	7.53	7.49	7.24
Cond mS/m	744	1624	2400	1513	1850	1600	1900	1800	1750	2100	1800	2100	2000	1973	1631	1591	1348
TDS cal mg/l	4028	11368	16800	10591	12950	11200	13300	12600	12250	14700	12600	14700	14000	13811	11417	11137	9436
Alk mg/l	2581	4802	6209	5050	6580	4140	5940	5500	5590	7450	6010	6570	7274	6362	5856	5100	3650
Ca mg/l	525	47.6	69	105	80	72	96	80	80	108	68	32	104	59	87	94	175
Mg mg/l	745	78.5	158	142	96	62	163	120	106	151	127	130	208	135	171	155	144
Na mg/l	768	1160	194	1412	1893	1000	1911	1924	1683	2024	2123	2058	1947	1647	1678	1630	1419
K mg/l	400	721	1354	810	762	1058	1345	1606	1299	1265	1538	1545	1149	1190	979	884	669
Fe mg/l				3,56	14	0.41	9.84	9.7	11.7		10.5	9.85			2.05	3.01	3.82
NO3-N mg/l		<0.05	<0.05	<0.05	13	0.3	3.9	0	ND	<0.05	<0.05	ND	<0.05	<0.05	<0.05	<0.05	<0.05
NH4-N mg/l	400	1917	1338	877	1305	1039	1058	1386	1366	1992	1336	1266	1102	1150	770	946	572
Cl mg/l	1190	2133	3110	2266	3600	2800	3010	3250	3100	3674	3250	3170	3538	2976	1618	1461	1811
SO4 mg/l		<0.16	<0.16	49.8	0.5	2	5	5	1	5.65	0.5	ND	<0.16	16.4	39.7	<0.16	68.1
Cu mg/l	30	<0.05	<0.05	NA						<0.05			<0.05	<0.05	<0.05	<0.05	<0.05
Pb micro g/l	104	56.4	<10	14						6.3			7.3	<40	<4	<4	39
Cd micro g/l	5.8	<10	<1	<1						2			<1	<10	1.1	<1	
Cr micro g/l	24	115	165	77	<1000	<1000	<1000	1290	280	114	300	1570	106	158	64		34
Hg micro g/l	4.4	2.2	0.5	<0.5	<1	<1	<1	<1	<1	3.4	1	<1	0.8	0.5	<0.5	<0.5	<0.5
As micro g/l				27	34	<1	26	60	8	15	TF		226		10.5	2.9	1
B micro g/l		212	409	2566	2000	10	400	800	1000	4378	2800	700	3462	3714	2608	2912	785
CN micr g/l		120	118	TF						<10			64.9	92.5	77.5	336	39.5
COD mg/l		2488	3383	1957	5490	3440	4520	4320	4240	3140	6240	6640	3177	2640	2259	1789	1677
Phenol micro g/l		420	374	149						356			246	146	258	204	129
VFA mg/l	56	<0.1	1.48	1.41						4.17			1.14	311	<0.1	75.4	93.9
Cations meq/l		214.85	155.25	161.96	207.51	153.52	211.79	237.73	216.8	280.62	241.05	231.85	215.29	198.2	171.59	178.56	140.35
Anions meq/l		156.14	211.79	166.25	233.55	162.11	203.97	201.65	199.14	252.61	211.76	220.7	245.14	211.47	163.52	143.15	166.29
SAR		24.02	2.95	21.13	33.79	20.87	27.57	31.8	29.05	29.52	35.13	36.15	25.38	27.05	24.08	24.02	19.25
RSC meq/l		87.25	107.78	84.11	119.33	74.12	100.64	96.16	99.11	131.22	106.39	119.14	123.23	113.22	96.75	84.6	52.45
USDA Class		C4S1NS	C4S1NS	C4S1NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS	C4S4NS

Bisasar Road (B)

Leachate Armco Drain Outlet

Date	11/4/94	24/2/98	11/7/95	20/8/96	14/1/97	10/7/97	4/12/97	20/1/98	20/2/98	18/3/98	29/4/98	20/5/98	4/6/98	18/6/98	6/10/98	17/12/98	19/7/99	14/12/99	5/7/00	21/12/00
Level m	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-	-
pH	7.95	8.38	7.88	7.2	8	8.17	7.96	8	7.4	8	8.2	7.9	8.08	7.8	7.7	7.42	7.93	7.41	7.38	8.42
Cond mS/m	706	827	841	829	710	989	549	660	490	800	810	770	959	690	1100	980	773	684	778	727
TDS cal mg/l	4942	5879	5887	4492	4970	6923	3843	4620	3430	5600	5670	5390	6713	4830	7700	6860	5411	4788	5446	5089
Alk mg/l	2226	2250	2080	2528	1762	2501	1426	1630	1376	2340	2040	2260	3144	2830	2830	3114	1963	404	1980	1990
Ca mg/l	77.5	96.6	92.6	1361	444	85	72	96	64	104	80	80	102	72	60	134	172	84	74	84
Mg mg/l	128	158.2	147	674	634	167	94	130	120	187	154	158	167	185	180	168	225	124	133	324
Na mg/l	668	985.7	876	848	679	1080	606	893	570	1036	1044	894	1104	1315	1395	968	1138	873	947	912
K mg/l	283	306.5	278	368	624	406	220	233	216	393	433	225	448	463	568	376	285	229	285	682
Fe mg/l	3.89	3.75	6.63		4.62		3	15	16	6.16	219.5	41.4		368	116			0.55	0.81	0.66
NO3-N mg/l	0.78	<0.05	<0.05		<0.05	<0.05	<0.05	2.8	4	1.7	0	ND	<0.05	<0.05	ND	<0.05	<0.05	<0.05	<0.05	<0.05
NH4-N mg/l	173	11.5	232		217	342	264	316	169	291	403	350	382	319	440	372	261	298	277	256
Cl mg/l	1332	1647	1600	1560	1110	1629	871	1270	880	1430	1850	1650	1630	1670	1780	1976	1200	1154	673	1088
SO4 mg/l	23.1	11.5	10.1		0.41	36	60.2	1	39	7.5	1.5	5	9.76	95	0.5	23.7	11.9	10.1	0.24	<0.16
Cu mg/l	NA	NA	NA	0.446	<0.05	<0.05							<0.05			<0.05	<0.05	<0.05	<0.05	<0.05
Pb micro g/l	NA	NA	NA	400	<2	<10	<4						7.6			29	<40	<4	<4	<4
Cd micro g/l	<1	<1	<1	17.2	<1	<1	<1						2.7			<1	<10	<1	<1	
Cr micro g/l	4.9	NA	NA	616	14	<10	10	<1000	<1000	<1000	840	<3	32	0.04	560	34	135	25	15	16
Hg micro g/l	<0.5	<0.5	1.1	3.9	<0.5	1.3	<0.5	<1	<1	<1	<1	1	2	4	<1	<0. 5	<0.5	<0.5	<0.5	<0.5
As micro g/l	NA	NA	NA				9.2	85	<1	57	34	19	28			14		5.2	6.9	3.6
B micro g/l	3920	2210	361		117	175	974	4000	570	200	1000	1100	4780	1400	800	2050	1969	783	778	500
CN micro g/l	NA	NA	NA		53.5	<10	<10						<10			<10	35.6	46.7	22	30.1
COD mg/l	683	539	636		540	709	376	1228	818	1380	1754	1392	1043	1608	2137	1804	624	467	1265	581
Phenol micro g/l	<5	<5	<5		<300	<50	<50						105			67.4	<50	<50	<50	<50
VFA mg/l	66	NA	39.2		<0.1	<0.1	<0.1						0.73			0.5	<0.1	24.8	<0.1	<0.1
Cations meq/l	43.63	69.76	78.89		136.18	99.94	62.37	83.53	55.58	96.88	102.08	86.79	105.76	110.82	124.61	96.97	102.45	79.63	82.88	106.16
Anions meq/l	51.56	91.63	86.95		66.52	96.66	54.63	68.6	53.41	87.36	92.94	91.78	109	96.62	106.75	118.44	73.37	40.8	58.56	88.73
SAR	10.84	14.33	13.19		2.8	15.68	11.08	13.97	9.71	14.06	15.74	13.35	15.66	18.67	20.36	13.16	13.46	14.16	15.24	10.11
RSC meq/l	30.15	27.24	25.16		0	32.08	17.22	17.14	14.48	26.27	24.18	28.25	44.09	28.84	38.85	41.81	12.22	0	25	9.04
USDA Class	C4S2NS	C4S2NS	C4S2NS		C4S1S	C4S2NS	C4S2NS	C4S2NS	C4S1NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2S	C4S2NS	C4S4NS

Bisasar Road (C)

Leachate Subsoil drain

Date	9/12/97	20/1/98	20/2/98	17/3/98	29/4/98	20/5/98	3/6/98	18/6/98	6/10/98	17/12/98	19/7/99	14/12/99	5/7/00	21/1200
Level	0	0	0	0	0	0	0	0	0	0	2.34	1.35	4.75	1.71
pH	7.34	7.5	7.6	7.6	7.7	7.4	7.28	7.4	7.4	7.34	7.61	7.41	7.56	7.24
Cond mS/m	867	1300	1020	1070	1100	1000	1176	980	1000	1062	618	595	551	459
TDS cal mg/l	6069	9100	7140	7490	7700	7000	8232	6860	7000	7434	4326	4165	3857	3213
Alk mg/l	2460	2570	2400	2520	2500	2440	3385	2140	2700	3029	1574	1756	1453	810
Ca mg/l	107	128	120	96	88	104	174	104	120	160	102	273	96	81
Mg mg/l	142	163	173	221	187	197	183	246	168	192	114	172	116	76
Na mg/l	788	1250	974	1375	1186	1200	1150	1478	1448	1046	811	900	736	501
K mg/l	272	323	456	476	536	312	448	694	476	390	284	254	205	306
Fe mg/l	13.2	22	18	16.7	21	28.95		40.7	18.3			0.11	0.14	0.08
NO3-N mg/l	<0.05	4.1	4.3	1.1	0	ND	<0.05	<0.05	<ND	<0.05	<0.05	<0.05	<0.05	17.7
NH4-N mg/l	239	476	420	395	454	456	421	392	378	376	207	152	113	80.3
Cl mg/l	1906	2350	2100	2070	2300	2350	2235	2600	1950	2063	1107	1016	413	660
SO4 mg/l	66.8	0.5	1	0.5	3	1	1.18	5	1	<0.16	26.9	64	<0.16	15.5
Cu mg/l	NA						<0.05			<0.05	0.08	<0.05	<0.05	<0.05
Pb micro g/l	<4						<4			<4	<40	7.6	4.6	4.6
Cd micro g/l	39						<1			<1	<10	<1	<1.0	
Cr micro g/l	15	<1000	<1000	<1000	700	70	21	<3	1240	16	9.2	42	5.7	7.5
Hg micro g/l	0.5	<1	<1	<1	1	<1	2.8	1	<1	<0.5	0.5	<0.5	<0.5	<0.5
As micro g/l	15	68	41	53	<1	8	17			9.8		2.5	9.8	4.7
B micro g/l	1426	4000	620	1800	600	600	2868	1500	1300	2096	1566	1418	1245	478
CN micro g/l	<10						<10			<10	67.5	51.7	26.1	76.6
COD mg/l	4617	2940	1776	1920	1848	1688	1151	1704	1858	1091	485	416	334	227
Phenol micro g/l	<50						<50			<50	<50	<50	<50	<50
VFA mg/l	0.7						0.11			1.87	<0.1	<0.1	<0.1	<0.1
Cations meq/l	75.95	117.38	104.43	123.99	117.69	114.34	115.48	135.71	122.13	106.31	71.91	84.47	59.63	45.63
Anions meq/l	104.28	117.9	107.48	108.8	114.85	115.02	130.68	116.14	108.95	118.69	63.22	65.07	40.79	42.58
SAR	11.76	17.29	13.33	17.66	16.42	15.97	14.52	18.04	20.02	13.2	13.12	10.51	11.96	9.61
RSC meq/l	32.21	31.64	27.82	27.49	30.27	27.45	44	17.44	34.23	36.84	17.04	7.37	14.85	5.92
USDA Class	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C40NS	C4S2NS	C4S2NS	C4S2NS	C4S2NS	C4S1NS

Bisasar Road (D)

Randles Cell

Date	10/9/99	14/12/99	5/7/00	21/12/00
pH	7.56	7.46	7.31	7.45
Cond mS/m	1260	736	1090	1303
TDS cal mg/l	8820	5152	7630	9121
Alk mg/l	3967	2636	4165	4380
Ca mg/l	25	171	300	237
Mg mg/l	21	202	224	642
Na mg/l	67	804	1084	1205
K mg/l	830	452	645	645
Fe mg/l		2.81	3.93	0.28
NO3-N mg/l	<0.05	<0.05	<0.05	<0.05
NH4-N mg/l	665	252	458	553
Cl mg/l	1669	1066	612	1201
SO4 mg/l	20.4	13.2	<0.16	6.32
Cu mg/l	<0.05	<0.05	<0.05	0.05
Pb micro g/l	74	17	8.1	45
Cd micro g/l	<10	<1	<1	
Cr micro g/l	113	47	85	95
Hg micro g/l	0.5	<0.5	<0.5	<0.5
As micro g/l		4.2	6.9	7.6
B micro g/l	365	1522	2378	2124
CN micro g/l	64.2	91.1	40.2	417
COD mg/l	1298	1007	3165	3522
Phenol micro g/l	<50	133	1020	93.7
VFA mg/l as Acetate	138	<0.1	298.6	1325
Cations meq/l	74.61	89.9	129.84	172.87
Anions meq/l	126.82	83.02	100.54	162.08
SAR	2.39	9.87	11.54	9.23
RSC meq/l	76.37	27.61	49.94	23.13
USDA Class	C4S1NS	C4S1NS	C4S2NS	C4S1NS

Mariannhill

Date	05/08/97	15/6/98	15/1/99	9/7/99	15/12/99	7/6/00
pH	7	7.55	7.58	8.05	6.94	8.28
Cond mS/m	34	574	1265	1127	558	1006
TDS calc mg/l	238	4018	8855	7889	3906	7042
Alk mg/l	114	2255	5524	4835	2493	2894
Ca mg/l	21	157		166	220	113
Mg mg/l	19	188		178	197	294
Na mg/l	25	404		972	453	992
K mg/l	3.3	369	940	388	320	710
Fe mg/l						1.95
Mn mg/l						0.38
NO3-N mg/l	0.1	<0.05	<0.05	<0.05	<0.05	<0.05
NH4-N mg/l	1.2	254	675	516	164	408
Cl mg/l	37	688	1197	1551	766	1565
F micro g/l		288		512		529
SO4 mg/l	14	<0.16		7.08	3.28	<0.16
Cu mg/l			<0.05			<0.05
Pb micro g/l	<20	74	<4.0			<4
Cd micro g/l	<10	1.2	<1.0			<1
Cr micro g/l	10	402	80			73
Hg micro g/l	0.3	<1.0				<0.5
As micro g/l			62			3.4
B micro g/l	<100	644	1653			2718
CN micro g/l	800	<10	<10			25.1
COD mg/l	63	1481	1605	1237	1066	913
Phenols micro g/l		58				<50
Cations meq/l	3.89	68.66		112.17	67.01	120.25
Anions meq/l	3.62	64.48		140.54	71.51	89.18
SAR	0.85	5.15		12.49	5.35	11.18
RSC meq/l	0	21.84		73.81	22.71	28.13
USDA Class	C2S1S	C4S1NS		C4S2NS	C4S1NS	C4S2NS

Shongweni

ANNEXURE V

FORM TO BE USED FOR CHEMICAL INFORMATION: CONDITIONS 7 AND 11.1

Name of site:			Shongweni Landfill	
Borehole/observation point name/number			Leachate	
Sampling Date (y-m-d)		3/4/00	Time	8h00
Method	Bail	Hand	Time after start of pump	min
	Pump		Depth of Sample	0 m
Date of analysis (y-m-d):		4/4/00	Laboratory	Umgeni Water

General chemistry

Constituent	Unit	Value	Constituent	Unit	Value
pH	(-log[H^+])	7.39	Al	(mg/l)	
EC	(mS/m)	1644	As (III)	(µg/l)	42
TDS	(mg/l)	11508	B	(µg/l)	4274
Ca	(mg/l)	384	Cd	(µg/l)	15
Mg	(mg/l)	195	CN	(µg/l)	478000
Na	(mg/l)	2700	Cr (Total)	(mg/l)	2069
K	(mg/l)	618	Cr (VI)	(mg/l)	
Alkalinity	(mg $CaCO_3$ /l)	3050	Cu	(mg/l)	<0.05
Cl	(mg/l)	1941	Fe	(mg/l)	10.6
SO_4	(mg/l)	183	Mn	(mg/l)	12.6
NO_3-N	(mg/l)	30.9	Pb	(µg/l)	10.1
F	(µg/l)	2280	Zn	(mg/l)	0.31
COD	(mg/l)	30478	Co	(µg/l)	18
NH_4-N	(mg/l)	600	Hg	(µg/l)	0.8
Organic N	(mg/l)		Li	(mg/l)	
Phenol	(µg/l)	5.15	Ni	(µg/l)	122
PO_4	(µg/l)	1822	Se	(µg/l)	<1
E. Coli	count/100 ml		U	(mg/l)	
DOC	(mg/l)		V	(µg/l)	130
TOX	(µg/l)				
TOC	(mg/l)				
PAH	(µg/l)				
PCH	(µg/l)				
VFA	(mg/l)	5026			
VOH	(mg/l)				

Appendix 4
Groundwater & Leachate Analyses of Samples Collected from the Ntuzuma Landfill Site on 17.01.2000

Determinand		Sampling Point			TWQR	TWQR	General
	Matafana River	Umgeni River	Front Berm	Leachate	Aquatic	Domestic	Standard
Major Leachate Constituents							
Ammonia (free & saline) $mg.l^{-1}$	1.7	<1	**15**	30	7	1	-
COD (as O_2) $mg.l^{-1}$	<5	8	**424**	**82**	-	-	75
Electrical Conductivity $mS.m^{-1}$	70	51	**771**	**185**	-	-	>75 $mg.l^{-1}$ change
Nitrate (as NO_3) $mg.l^{-1}$	20	8.5	<1	1.5	-	-	-
pH	7.81	7.91	7.65	7.75	+-5%	6 - 9	5.5 - 9.5
						-	-
Minor Leachate Constituents						-	-
Aluminium (as Al) $mg.l^{-1}$	**0.3**	**1.4**	<0.1	<0.1	0.007	0.15	-
Alkalinity (as $CaCO_3$) $mg.l^{-1}$	61	88	2045	572	-	-	-
Arsenic (as Ar) $\mu g.l^{-1}$	4	5	6	6	-	10	500
Barium (as Ba) $\mu g.l^{-1}$	208	225	1550	700	-	-	500
Boron (as B) $\mu g.l^{-1}$	172	161	754	324	500*	-	100
Cadmium (as Cd) $\mu g.l^{-1}$	<5	5	<5	<5	0.25	5	50
Calcium (as Ca) $mg.l^{-1}$	11	10	68	65	-	32	-
Chloride (as Cl) $mg.l^{-1}$	**252**	89	67	**1773**	-	**100**	100
Chromium (as Cr VI) $mg.l^{-1}$	<0.1	<0.1	<0.1	<0.1	0.007	0.05	0.05
Total Chromium (as Cr) $mg.l^{-1}$	<0.1	<0.1	<0.1	<0.1	0.012	-	0.5
Copper (as Cu) $mg.l^{-1}$	<0.1	<0.1	<0.1	<0.1	0.0003	1	1
Free cyanide (as CN) $\mu g.l^{-1}$	<0.1	<0.1	<0.1	<0.1	1	-	500
Fluoride (as F) $\mu g.l^{-1}$	200	300	300	400	750	2000	1000
Iron (as Fe) $mg.l^{-1}$	**0.26**	**0.91**	<0.1	<0.1	<10%	**0.1**	-
					change		
Lead (as Pb) $\mu g.l^{-1}$	**9**	**14**	13	16	**0.2**	10	0.1
Magnesium (as mg) $mg.l^{-1}$	24	17	563	31	-	30	-
Manganese (as mn) $mg.l^{-1}$	<0.1	<0.1	**0.38**	**2.35**	0.18	0.05	0.4
Mercury (as Hg) $\mu g.l^{-1}$	1	**2**	**<1**	**3**	**0.4**	**1.0**	20
Potassium (as K) $mg.l^{-1}$	2.1	1.7	25	34	-	50	-
Sodium (as Na) $mg.l^{-1}$	75	66	845	116	-	100	<90 $mg.l^{-1}$ change
Soluble orthophosphate	<0.1	<0.1	<0.1	<0.1	<15%	-	-
($H_2 PO_4$ & HPO_4^{2-}) $mg.l^{-1}$					change		-
Sulphate (SO4) $mg.l^{-1}$	45	39	16	14	-	200	-
Total dissolved Solids $mg.l^{-1}$	456	322	6112	1042	<15% change	3000	-
Total Phenols $\mu g.l^{-1}$	<0.1	<0.1	<0.1	<0.1	30	1	100
Zinc (as Zn) $mg.l^{-1}$	<0.1	<0.1	<0.1	<0.1	0.002	3	5

General Standard: The sum of the following metals shall not exceed 1 $mg.l^{-1}$: Cadmium (as Cd), chromium (as Cr), Copper (as Cu), mercury (as Hg) & lead (as Pb)
TWQR: Target Water Quality Range as per the South African Water Quality Guidelines for Aquatic Ecosystems & Domestic Use
*Boron Target Water Quality Range for irrigation use.

TABLE 1

SABS 241 - 1984 SPECIFICATION FOR DOMESTIC WATER SUPPLIES - HAMMARSDALE LANDFILL SITE WATER QUALITY ANALYSIS

DETERMINAND	UNIT	RECOMMEND LIMIT	MAX. ALLOWABLE LIMIT	HDM BH 1	HDM BH 6	HDPD 1	HDPD 2	HDPS 2	HDPS 3	HD L/FILL TP 12	LEACH.
COLOUR		20	ns	<5	<5	<5	<5	<5	5	45	400
TURBIDITY	NTU	1	5	0.1	0.6	2.1	5.7	***160***	*13*	*13*	*14*
pH	pH Units	6-9	5.5 - 9.5	6.7	7.1	7.4	7.3	7.1	7.0	7.8	9.2
CONDUCTIVITY	mS/m	70	300	70	**98***	33	57	**72***	**99***	***385*****	***900*****
TOTAL HARDNESS	mg/l	20 - 300	ns - 650	202	**392**	106	190	154	**368**	***660***	***1000***
CALCIUM HARDNESS	mg/l	ns	ns	106	230	44	94	32	94	160	120
ALKALINITY	mg/l	ns	ns	200	280	104	226	312	254	662	1500
TOTAL DIS. SOLIDS	mg/l	ns	ns	540*	762*	344	568*	736*	920*	2942***	8252***
MAGNESIUM	mg/l	70	100	23	39	15	23	29	42	***120*****	***211*****
CALCIUM	mg/l	150	200	42	92	18	38	13	78	64	48
SODIUM	mg/l	100	400	**102***	92	43	79	**131***	**102***	***742.2*****	***715*****
CHLORIDE	mg/l	250	600	115*	160*	52	67	80	190*	***850*****	***1570*****
SULPHATE	mg/l	200	600	64	80	6.5	15	10	52	**500****	***1900*****
NITRATE	mg/l	6	10	0.8	0.2	1.3	0.10	0.2	0	**8.2***	3.3
FLUORIDE	mg/l	1.0	1.5	0.4	**1.2***	0.3	0.2	0.9	0	0.8	1.0
ZINC	mg/l	1	5	<0.5	<0.5	<0.2	<0.2	<0.2	<0.2	<0.5	<0.5
ARSENIC	µg/l	100	300	<0.1	<0.1	<1	<1	32*	<1	<1	**212*****
CADMIUM	µg/l	0.01	0.02	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1
COPPER	mg/l	0.5	1.0	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1	<0.1
CYANIDE	µg/l	200	1	10	10	1	1	4	2	10	8.0
IRON	mg/l	0.1	1	**0.16***	**0.22****	**0.45****	**0.16****	**14.2*****	**0.55****	**0.86****	***3.0*****
LEAD	µg/l	50	100	<1	<1	<1	<1	<1	<1	<1	<1
MANGANESE	mg/l	0.05	1	0.023	**0.26****	0.04	**0.18****	**7.20*****	**0.53****	**0.058***	**0.73****
MERCURY	µg/l	5	10	<1.0	<1.0	<1.0	<1.0	<1.0	<1.0	<1.0	**20**
PHEN. COMPOUNDS	µg/l	5	10	2.0	<0.1	3.0	2.0	-	-	**7**	***513***

Appendix 7

Borehole data and chemical analyses recorded as per Annexure IV of the Permit

ANNEXURE IV

FORM TO BE USED FOR CHEMICAL INFORMATION: CONDITIONS 5,7,8 AND 10.2

Name of site:			La Mercy Landfill		
Borehole/observation point name/number			Leachate		
Sampling Date (y-m-d)		31/01/2001	Method		Bail x
Sampling time		10h00			Pump
Time after start of pump:			Depth of Sample		
Date of analysis (y-m-d):		1/02/2001	Laboratory		Echalaz & Osborne
General chemistry					
Constituent	Unit	Value	Constituent	Unit	Value
pH	(-log[H^+])	7.53	Al	(mg/l)	
EC	(mS/m)	95	As (III)	(mg/l)	
TDS	(mg/l)		B	(mg/l)	135
Ca	(mg/l)	35	Cd	(mg/l)	<0.1
Mg	(mg/l)	20	free CN	(mg/l)	<5
Na	(mg/l)	95	Cr (Total)	(mg/l)	<0.1
K	(mg/l)	32	Cr (VI)	(mg/l)	
Alkalinity	(mg $CaCO_3$ /l)	144	Cu	(mg/l)	
Cl	(mg/l)	187	Fe	(mg/l)	0.5
SO4	(mg/l)	28	Mn	(mg/l)	
NO3-N	(mg/l)	<0.5	Pb	(mg/l)	<0.1
F	(mg/l)	0.3	Zn	(mg/l)	
COD	(mg/l)	129	Co	(mg/l)	
NH4-N	(mg/l)	1.2	Hg	(mg/l)	10
Organic N	(mg/l)		Li	(mg/l)	
Phenol	(mg/l)	<0.1	Ni	(mg/l)	
PO_4	(mg/l)		Se	(mg/l)	
E. Coli	count/100 ml		U	(mg/l)	
DOC	(mg/l)		V	(mg/l)	
TOX	(mg/l)				
TOC	(mg/l)				
PAH	(mg/l)				
PCH	(mg/l)				
VFA	(mg/l)				
VOH	(mg/l)				

Appendix 8

Previous Audit Data

Leachate from leachate sump

Determinand	***Unit***	***02.04.98***	***15.06.98***	***27.01.2000***
Level m	m	0	0	0
pH		7.50	8.34	7.16
Conductivity	$mS.m^{-1}$	51.5	876	278
Total Dissolved Solids	$mg.l^{-1}$	360.5	6132	1534
Alkalinity	$mg.l^{-1}$	196	3000	788
Calcium (as Ca)	$mg.l^{-1}$	16	103	27
Magnesium (as mg)	$mg.l^{-1}$	22	318	93
Sodium (as Na)	$mg.l^{-1}$	44	805	259
Potassium (as K)	$mg.l^{-1}$	16	437	75
Nitrate (as NO_3)	$mg.l^{-1}$	<0.05	<0.05	<1
Ammonium (as NH_4)	$mg.l^{-1}$	9.3	394	<1
Chloride (as Cl)	$mg.l^{-1}$	67.1	197	430
Sulphate (as SO_4)	$mg.l^{-1}$	8.9	<0.16	8.20
COD	$mg.l^{-1}$	97.8	772	348
Volatile Fatty Acids	$mg.l^{-1}$	<0.1	0.15	-
Phenol	$\mu g.l^{-1}$	<50	<50	-
Mercury (as Hg)	$\mu g.l^{-1}$	<0.5	<0.5	-
Boron (as B)	$\mu g.l^{-1}$	<20	1404	398
Cadmium (as Cd)	$\mu g.l^{-1}$	<1	<1	<1
Chromium (as Cr)	$\mu g.l^{-1}$	<3	39	<100
Lead (as Pb)	$\mu g.l^{-1}$	<4	<4	13
Cyanide (as CN)	$\mu g.l^{-1}$	<10	<10	-
Cations	$meq.l^{-1}$	5.62	105.97	**-**
Anions	$meq.l^{-1}$	6	65.55	**-**
SAR		1.68	8.86	-
RSC	$meq.l^{-1}$	1.32	28.78	-
USDA	class	C2S1M	C4S1NS	-

Table 7 : Leachate from leachate dam

Determinand	**Unit**	**02.04.98**	**15.06.98**	**27.01.00**	**19.02.01**
Level	m	0	0	0	0
pH		7.5	8.34	7.16	7.85
Conductivity	mS.m-1	51.5	876	278	101
Total Dissolved Solids	mg.l-1	360.5	6132	1534	707
Alkalinity	mg.l-1	196	3000	788	265
Calcium	mg.l-1	16	103	27	47
Magnesium	mg.l-1	22	318	93	33
Sodium	mg.l-1	44	805	259	79
Potassium	mg.l-1	16	437	75	26
Nitrate (as NO_3)	mg.l-1	<0.05	<0.05	<1	2
Ammonium (as NH_4)	mg.l-1	9.3	394	<1	3.8
Chloride (as Cl)	mg.l-1	67.1	197	430	152
Sulphate	mg.l-1	8.9	<0.16	8.2	26
COD	mg.l-1	97.8	772	348	45
Volatile Fatty Acids	mg.l-1	<0.1	0.15	-	<0.1
Phenol	µg.l-1	<50	<50	-	<10
Mercury	µg.l-1	<0.5	<0.5	-	3
Boron	µg.l-1	<20	1404	398	110
Cadmium	µg.l-1	<1	<1	<1	<10
Chromium	µg.l-1	<3	39	<100	<5
Chromium IV	µg.l-1				<10
Lead	µg.l-1	<4	<4	13	<10
Cyanide	µg.l-1	<10	<10	-	<5
Cations	meq.l-1	5.62	105.97	22.35	9.47
Anions	meq.l-1	6	65.55	28.11	10.277

APPENDIX H

EXPERIMENTAL CONDITION USING CELLULOSE ACETATE MEMBRANES FOR THE TREATMENT OF THE MUNICIPAL SOLID WASTE LEACHATE AT BISASAR ROAD

CWF flux conducted with Durban tap water					Membrane area		1.75	m2		Membrane	CA	Page No		Dosing concentration		12.5	mg/l				
																				Dosing rate	
Date	Time	Time	Pressure		Temp	Flux	Flux Normal	Feed		Feed in tank		Brine to waste		Product		Brine		Recovery	Rejection	Measured	Calculated
		(h)	(kPa) in	(kPa) out	(deg C)	(ml/min)	(l/m2.d)	(mS/cm)	(pH)	(mS/cm)	(pH)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ml/min) out	(l/5 s)	(%)	(%)	(ml/min)	(ml/min)
08/08/2001	11:00	CWF	4000	3900	18.5	740	708										1.19				
	12:00	CWF	4000	3900	18.5	740	708	Tank 1									1.25				
Start	13:45	0	4000	3900	26	680	546	1867	6.32	1928	6.51			110.6	5.21	300		69.39	94.26		12.3
	14:45	1	4000	3900	28	640	487			2140	6.81	2200	6.8	119.7	5.4	280		69.57	94.41		11.5
	15:45	2	4000	3900	28.5	610	458			2390	7.03	2490	7.06	139	5.7	270		69.32	94.18		11.0
	16:45	3	4000	3900	28	575	438			2640	7.09	2680	7.22	159	5.84	260		68.86	93.98		10.4
	17:45	4	4000	3900	27.5	550	424			2800	7.17	2830	7.3	173	5.94	225		70.97	93.82		9.7
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																				
09/08/2001	Rinsed membrane with tap water																				
	09:30	CWF	4000	3900	18	700	677										1.17				
Start	10:00	4																			
	10:30	4.5	4000	3900	21.5	500	447			2670	7.52	2760	7.53	168	6.41	230		68.49	93.71		9.1
	12:30	6.5	4000	3900	25.5	500	406			2970	7.3	3020	7.41	198	6.19	190		72.46	93.33		8.6
	14:30	8.5	4000	3900	26.5	490	388			3210	7.43	3250	7.44	221	6.27	200		71.01	93.12		8.6
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																				
10/08/2001	Rinsed membrane with tap water																				
	08:25	CWF	4000	3900	18.5	690	660										1.2				
Start	09:00	8.5	4000	3900	20.5	440	403			3240	7.39	3300	7.41	221	6.3	192		69.62	93.18		7.9
	11:00	10.5	4000	3900	25	455	374			3420	7.39	3520	7.47	246	6.4	169		72.92	92.81		7.8
	13:00	12.5	4000	3900	26.5	455	360			3590	7.43	3680	7.46	265	6.54	183		71.32	92.62		8.0
	Rinced with tap water 30 min / preserve (50ml/20l) Formaldehyde																				
13/08/2001	Rinsed membrane with tap water																				
	09:20	CWF	4000	4400	17.5	730	713										1.17				
Start	09:40	12.5	4000	4300	18	455	440			3450	7.63	3530	7.62	227	6.87	193		70.22	93.42		8.1
	10:10	13	4000	4200	20.4	470	431			3460	7.57	3590	7.57	229	6.67	176		72.76	93.38		8.1
	12:10	15	4000	4200	24.2	480	403			3680	7.6	3780	7.59	277	6.88	150		76.19	92.47		7.9
	14:10	17	4000	4100	24.5	490	408			3883	7.59	3880	7.57	269	6.86	200		71.01	93.07		8.6
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																				

14/08/2001	Rinsed membrane with tap water																			
	09:00	CWF	4000	4350	18	745	720	Tank 2									1.23			
Start	09:10	17	4000	4300	18.7	445	424	19.19	6.27	3650	7.53	3780	7.56	251	6.69	200		68.99	93.12	8.1
	09:40	17.5	4000	4100	20.7	450	410			3680	7.5	3780	7.51	248	6.52	204		68.81	93.26	8.2
	11:40	19.5	4000	4150	24.2	450	378			3770	7.47	3920	7.48	255	6.54	195		69.77	93.24	8.1
	13:40	21.5	4000	4200	25	470	387			3960	7.5	4090	7.49	290	6.67	190		71.21	92.68	8.3
	15:40	23.5	4000	4100	24.5	450	375			4040	7.48	4180	7.53	314	6.8	220		67.16	92.23	8.4
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																			
15/08/2001	Rinsed membrane with tap water																			
	09:10	CWF	4000	4250	18.5	745	713										1.18			
Start	09:20	23.5	4000	4300	20	430	398			3900	7.57	4050	7.57	279	6.79	214		66.77	92.85	8.1
	10:00	24	4000	3800	22	430	380			3950	7.55	4080	7.54	291	6.67	200		68.25	92.63	7.9
	12:00	26	4000	4100	24.5	440	367			4040	7.56	4160	7.53	291	6.72	184		70.51	92.80	7.8
	14:00	28	4000	4100	25	440	362			4170	7.61	4260	7.59	304	6.96	180		70.97	92.71	7.8
	15:45	29.75	4000	4200	25	430	354			4260	7.61	4340	7.64	315	7.15	170		71.67	92.61	7.5
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																			
16/08/2001	Rinsed membrane with tap water																			
	09:10	CWF	4000	4350	18.8	750	713	Tank 3									1.15			
Start	09:20	29.75	4000	4350	19	410	388	1882	6.35	4110	7.66	4240	7.65	298	7.37	180		69.49	92.75	7.4
	11:20	31.75	4000	4200	24	440	371			4070	7.51	4260	7.59	306	6.96	190		69.84	92.48	7.9
	13:20	33.75	4000	4150	25	420	346			4210	7.55	4290	7.61	311	7.09	189		68.97	92.61	7.6
	14.05	34.5	4000	4300	25.4	420	342			4200	7.59	4310	7.59	311	7.13	181		69.88	92.60	7.5
	Rinced with tap water 30 min / preserve (50ml/20l) Formaldehyde																			
17/08/2001	Rinsed membrane with tap water																			
	08:30	CWF	4000	4350	18	715	691										1.22			
Start	08:45	34.5	4000	4350	19.5	420	393			3900	7.55	4120	7.66	225	7.73	175		70.59	94.23	7.4
	09:15	35	4000	4350	21	440	398			3990	7.58	4110	7.59	288	6.86	180		70.97	92.78	7.8
	11:15	37	4000	4300	24.8	435	360			3960	7.54	4150	7.56	297	6.92	184		70.27	92.50	7.7
	13:15	39	4000	4150	26	440	353			4110	7.59	4230	7.6	310	7.14	180		70.97	92.46	7.8
	15:15	41	4000	4150	26	440	353			4170	7.58	4290	7.69	333	7.39	180		70.97	92.01	7.8
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																			
20/08/2001	Rinsed membrane with tap water																			
	08:45	CWF	4000	4300	18.5	750	717	Tank 1									1.19			
Start	09:00	41	4000	4350	19.8	410	381	2000	6.31	3840	7.35	3950	7.39	277	7.37	160		71.93	92.79	7.1
	10:00	41.5	4000	3300	21	420	380			3810	7.31	3960	7.34	262	6.21	184		69.54	93.12	7.6
21/08/20	Rinsed membrane with tap water																			

01																					
	09:25	CWF	4000	2600	19.5	710	665										1.2				
Start	09:30	43.5	4000	4300	19.8	415	386			3900	7.42	3990	7.45	140	6.84	185		69.17	96.41		7.5
	10:00	44	4000	3900	22.5	425	372			3880	7.38	4020	7.47	273	6.32	200		68.00	92.96		7.8
	12:00	46	4000	3800	26	455	365			3970	7.37	4100	7.41	299	6.39	182		71.43	92.47		8.0
	14:00	48	4000	4100	27	440	344			4130	7.41	4240	7.48	306	6.64	180		70.97	92.59		7.8
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																				
22/08/2001	Problem with pressure gauge and sponge ball.																				
23/08/2001	Rinsed membrane with tap water																				
	13:00	CWF	4000	3850	19.25	650	612										1.26				
Start	13:15	48																			
	14:00	48.75	4000	3850	22.6	350	305			3980	7.45	4180	7.53	360	6.72	118		74.79	90.95		5.9
	16:00	50.75	4000	3950	27.3	370	287			4140	7.5	4240	754	368	6.85	152		70.88	91.11		6.5
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																				
24/08/2001	Rinsed membrane with tap water																				
	09:20	CWF	4000	3950	19	645	610										1.22				
		Rinse 30 min spongeball recycle																			
	10:15	CWF	4000	3950	19.2	645	608										1.16				
	1 Hour HNO_3 clean - 30 min rinse																				
	12:40	CWF	4000	3950	19	645	610										1.19				
	STPP and EDTA clean 1 hour. 30 min water rinse																				
	14:50	CWF	4000	3950	19.3	680	639										1.23				
	Preserve (50ml/20l) Formaldehyde																				
28/08/2001	Rinsed membrane with tap water																				
	08:50	CWF	4000	3950	19.2	660	622										1.22				
Start	09:00	50.75																			
	10:00	51.75	4000	3950	22.6	385	336			3870	7.5	3970	7.51	409	6.66	174		68.87	89.43		7.0
	12:00	53.75	4000	3950	25.7	400	323			3980	7.47	4020	7.52	410	6.79	171		70.05	89.70		7.1
	14:00	55.75	4000	3950	26.5	400	317			4050	7.52	4130	7.57	426	6.95	150		72.73	89.48		6.9
	16:00	57.75	4000	3950	27	400	313			4100	7.54	4220	7.6	439	7.07	152		72.46	89.29		6.9
	19:00	60.75	4000	3950	27.2	375	292	Tank 2		4250	7.58	4340	7.6	469	7.48	170		68.81	88.96		6.8
	24:00:00	65:75						1970	6.31												
	Preserve (50ml/20l) Formaldehyde																				
29/08/2001	Rinsed membrane with tap water																				
Plant stopped automatically overnight. Antiscalent keeps on running																					
	12:30	CWF	4000	3950	19.7	640	596										1.24				

Start	12:45	65.75																				
	13:00	66	4000	3950	22	440	389			3160	6.96	3250	7	285	6	202		68.54	90.98		8.0
	14:00	67	4000	3950	24.8	440	364	Tank 3		3250	7.21	3320	7.25	272	6.1	186		70.29	91.63		7.8
	16:00	69	4000	3950	27	440	344	1940	6.23	3510	7.42	3580	7.44	311	6.44	174		71.66	91.14		7.7
	18:45	71.75	4000	3950	27	420	328			3740	7.35	3810	7.42	344	6.48	180		70.00	90.80		7.5
	22:45	75.75	4000	3950	27	400	313			3950	7.41	4050	7.45	376	6.55	178		69.20	90.48		7.2
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																					
31/08/2001	Rinsed membrane with tap water																					
	09:40	CWF	4000	3950	19.6	660	616										1.28					
Start	09:45	75.75																				
	11:00	77	4000	3950	23.7	380	323			3880	7.61	3990	7.65	390	6.74	162		70.11	89.95		6.8
	12:00	78	4000	3950	25	380	313			3930	7.65	4000	7.67	395	6.73	170		69.09	89.95		6.9
	14:30	80.5	4000	3950	25.6	310	251			3990	7.69	4070	7.7	505	6.95	154		66.81	87.34		5.8
	15:30	81.5	4000	3950	25.9	310	249			4000	7.69	4100	7.75	506	7.06	135		69.66	87.35		5.6
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																					
03/09/2001	Rinsed membrane with tap water																					
	09:15	CWF	4000	3950	20	665	616										1.24					
Start	09:30	81.5																				
	11:15	83.25	4000	3700	23.2	385	331			3700	7.76	3800	7.78	404	7.22	163		70.26	89.08		6.9
	12:30	84.5	4000	3900	25.5	415	337			3770	7.66	3840	7.65	375	6.85	165		71.55	90.05		7.3
	14:30	86.5	4000	3950	26.6	410	324			3890	7.72	3970	7.72	401	7.14	160		71.93	89.69		7.1
	16:15	88.25	4000	3950	26.7	400	315			3940	7.75	4050	7.79	419	7.41	180		68.97	89.37		7.3
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																					
04/09/2001	Rinsed membrane with tap water																					
	09:25	CWF	4000	3950	20.2	660	608										1.18					
Start	09:30	88.25																				
	11:45	90.5	4000	3950	26.6	400	316	Tank 2		3880	7.52	3980	7.6	390	6.7	185		68.38	89.95		7.3
	13:45	92.5	4000	3950	27.8	395	302	1950	6.3	3970	7.18	4070	7.56	402	6.67	176		69.18	89.87		7.1
	15:45	94.5	4000	3950	27.9	390	298			4040	7.52	4150	7.6	409	6.69	176		68.90	89.88		7.1
	17:00	95.75	4000	3950	21.6	380	339			4070	7.5	4180	7.58	411	6.66	176		68.35	89.90		7.0
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																					
05/09/2001	Rinsed membrane with tap water																					
	09:25	CWF	4000	3950	20.3	655	602										1.22					
Start	09:30	95.75																				
	12:00	98.25	4000	3950	26.5	400	317			4040	7.49	4150	7.58	408	6.68	148		72.99	89.90		6.9
	14:00	100.25	4000	4000	27.5	350	270			4101	7.57	4190	7.62	496	6.82	160		68.63	87.91		6.4
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																					
06/09/20	Rinsed membrane with tap water																					

01																				
	08:55	CWF	4000	3900	20.5	660	604										1.24			
Start	09:00	100.25																		
	09:45	101	4000	3950	24.3	385	322			3760	7.64	3890	7.73	373	6.94	168		69.62	90.08		6.9
	12:30	104.25	4000	3950	27	410	321			3890	7.49	3980	7.58	376	6.66	172		70.45	90.33		7.3
	15:45	107.5	4000	3950	28.2	390	295			4070	7.52	4160	7.59	408	6.65	170		69.64	89.98		7.0
	20:30	112.25	4000	3950	27.8	370	283			4170	7.52	4340	7.6	429	6.71	164		69.29	89.71		6.7
07/09/2001	03:45	119.5	4000	3950	27.4	340	263			4360	7.54	4470	7.62	449	6.77	162		67.73	89.70		6.3
	06:30	122.25	4000	3900	27	340	266			4430	7.64	4510	7.68	460	6.81	157		68.41	89.62		6.2
	08:30	124.25	4000	3950	27.5	340	262	Tank 3		4440	7.62	4530	7.65	460	6.87	152		69.11	89.64		6.2
	09:00	124.75	4000	3950	27.6	340	262	1993	6.35	4450	7.71	4570	7.81	463	7.05	154		68.83	89.60		6.2
	Rinsed with tap water 30 min																			
	09:45	CWF	4000	3950	20.8	620	564										1.24			
Start	10:00	124.75																		
	13:00	127.75	4000	3950	27.6	355	273			4340	7.5	4430	7.56	434	6.6	154		69.74	90.00		6.4
	15:00	129.75	4000	3950	28	350	266			4390	7.48	4480	7.56	447	6.63	151		69.86	89.82		6.3
	17:00	131.75	4000	3950	27.5	340	262			4430	7.48	4540	7.58	447	6.67	155		68.69	89.91		6.2
	22:00	136.75	4000	3950	27.8	340	260			4500	7.49	4590	7.57	456	6.65	152		69.11	89.87		6.2
08/09/2001	00:00	138.75	4000	3950	27.8	330	253			4470	7.49	4610	7.58	457	6.64	152		68.46	89.78		6.0
	08:15	147	4000	3950	27.5	320	247			4570	7.49	4660	7.58	470	6.71	150		68.09	89.72		5.9
	11:00	149.75	4000	3950	28.1	330	250			4450	7.46	4670	7.56	472	6.75	152		68.46	89.39		6.0
	Rinsed with tap water 30 min																			
	11:55	CWF	4000	3950	20.6	615	562	Tank 1									1.21			
Start	12:00	149.75						2030	6.33											
	13:00	150.75	4000	3950	27.3	335	260			4450	7.43	4540	7.52	446	6.61	148		69.36	89.98		6.0
	19:30	157.25	4000	3950	27.8	335	256			4610	7.41	4730	7.5	465	6.58	141		70.38	89.91		6.0
	20:30	158.25	4000	3950	28	330	251			4610	7.38	4730	7.5	464	6.58	141		70.06	89.93		5.9
09/09/2001	00:15	162	4000	3950	27.5	325	251			4680	7.42	4790	7.49	481	6.59	140		69.89	89.72		5.8
	09:15	171	4000	3950	27.6	315	242			4690	7.38	4880	7.52	503	6.63	140		69.23	89.28		5.7
	Rinsed with tap water 30 min																			
	10:00	CWF	4000	3950	21.6	620	554										1.22			
Start	10:15	171																		
	12:15	173	4000	3950	27.7	340	261	Tank 2		4540	7.38	4670	7.48	451	6.52	152		69.11	90.07		6.2
	18:45	180.5	4000	3950	28.8	340	253	2020	6.28	4590	7.3	4740	7.42	469	6.44	150		69.39	89.78		6.1
	23:00	184.5	4000	3950	28.2	335	254			4650	7.34	4780	7.44	473	6.51	148		69.36	89.83		6.0
10/09/2001	07:30	193	4000	3950	27.5	330	255	Tank 3		4730	7.38	4860	7.47	485	6.57	150		68.75	89.75		6.0
	09:45	195.25	4000	3950	28.2	320	242	2020	6.28	4550	7.36	4820	7.49	484	6.6	148		68.38	89.36		5.9
	12:00	197.5	4000	3950	28.4	325	245			4690	7.35	4820	7.48	484	6.61	147		68.86	89.68		5.9
	Rinsed with tap water 30 min																			
	12:50	CWF	4000	3950	20.6	600	548													1.2	
Start	13:00	197.5																		

	15:00	199.5	4000	3950	28.6	340	255			4550	7.27	4660	7.38	456	6.43	138		71.13	89.98		6.0
	16:30	201	4000	3950	28.9	340	252			4480	7.2	4700	7.4	463	6.46	136		71.43	89.67		6.0
	22:00	206.5	4000	3950	29	330	244	Tank 1		4660	7.23	4840	7.39	493	6.48	132		71.43	89.42		5.8
								2000	6.3												
11/09/2001	06:30	215	4000	3900	28.5	320	240			4770	7.27	4920	7.44	515	6.58	132		70.80	89.20		5.7
	08:30	217	4000	3900	28.5	320	240			4820	7.39	4960	7.43	519	6.62	131		70.95	89.23		5.6
	10:45	219.25	4000	3950	29	320	237			4840	7.33	4950	7.45	521	6.61	128		71.43	89.24		5.6
	Rinsed membrane with tap water																				
	11:20	CWF	4000	3950	21	610	552										1.21				
		Rinse 30 min spongeball recycle																			
	12:05	CWF	4000	3950	21	620	561										1.24				
	1 Hour HNO_3 clean - 30 min rinse																				
	14:00	CWF	4000	3950	20.9	620	562										1.21				
	STPP and EDTA clean 1 hour. 30 min water rinse																				
	16:45	CWF	4000		20.8	550	500										1.24				
12/09/2001	Rinsed with tap water 30 min																				
	09:25	CWF	4000		20	550	509										1.25				
Start	09:30	219.25	4000		20.2	216	199			4870	7.64	4880	7.66	972	6.98	98		68.79	80.04		3.9
	11:00	220.75	4000		22.8	260	226			4350	7.31	4430	7.42	790	6.58	108		70.65	81.84		4.6
	15:00	224.75	4000		28.5	300	225			4380	7.42	4500	7.44	790	6.75	110		73.17	81.96		5.1
	16:00	225.75	4000		26	275	221			4380	7.41	4480	7.55	791	6.77	130		67.90	81.94		5.1
	22:30	232.25	4000		23.9	255	216	Tank 2		4350	7.4	4480	7.54	779	6.72	120		68.00	82.09		4.7
								2030	6.28												
13/09/2001	07:00	240.75	4000		23	230	199			4410	7.46	4460	7.59	770	6.78	127		64.43	82.54		4.5
	09:45	243.5	4000		23	220	190			4370	7.49	4440	7.52	750	6.74	120		64.71	82.84		4.3
	11:45	245.5	4000		23	200	173			4350	7.44	4420	7.55	751	6.69	120		62.50	82.74		4.0
	12:45	246.5	4000		22.5	258	226			4320	7.41	4420	7.53	741	6.63	112		69.73	82.85		4.6
	15:15	249	4000		23.5	265	226			4370	7.54	4420	7.56	741	6.8	111		70.48	83.04		4.7
	22:00	255.75	4000		23.3	260	223			4420	7.45	4500	7.57	757	6.71	110		70.27	82.87		4.6
	Rinsed with tap water 30 min																				
	22:55	CWF	4000		18.6	510	487										1.25				
Start	23:00	255.75																			
14/09/2001	09:30	266.25	4000		23.5	255	218			4300	7.35	4450	7.44	767	6.56	108		70.25	82.16		4.5
	11:30	268.25	4000		24.4	260	217			4380	7.38	4490	7.57	778	6.72	103		71.63	82.24		4.5
	15:00	271.75	4000		24.6	260	216			4410	7.43	4510	7.56	782	6.76	100		72.22	82.27		4.5
	Rinsed with tap water 30 min																				
	14:55	CWF	4000		19	510	483										1.24				
	Rinsed with tap water 30 min / preserve (50ml/20l) Formaldehyde																				

17/09/2001	Rinsed membrane with tap water																				
Start	10:00	271.75						Tank 3													
	12:00	273.75	4000		24.4	290	242	1686	6.3	4200	7.41	4290	7.47	712	6.7	116		71.43	83.05		5.1
	15:00	276.75	4000		26.5	300	238			4130	7.35	4210	7.49	703	6.67	115		72.29	82.98		5.2
	16:00	277.75	4000		26.6	300	237			4090	7.36	4170	7.47	697	6.62	130		69.77	82.96		5.4
	Rinsed with tap water 30 min																				
	16:40	CWF	4000		19.5	520	487										1.23				
Start	16:45	277.75																			
	22:15	283.25	4000		28.2	320	242			3900	7.31	3990	7.44	639	6.57	122		72.40	83.62		5.5
18/09/2001	07:00	292	4000		26	330	265			3820	7.34	3940	7.46	615	6.58	129		71.90	83.90		5.7
	08:30	293.5	4000		26.9	335	263			3770	7.3	3910	7.44	582	6.52	129		72.20	84.56		5.8
	Rinsed with tap water 30 min																				
	09:30	CWF	4000		20	520	481										1.23				
Start	09:30	293.5																			
	13:00	297	4000		26.8	325	255			3700	7.36	3800	7.45	577	6.54	130		71.43	84.41		5.7
	15:45	299.75	4000		27	330	258	Tank 1		3700	7.33	3790	7.46	570	6.57	140		70.21	84.59		5.9
	21:00	305	4000		27	330	258	1430	6.34	3570	7.4	3680	7.48	545	6.6	140		70.21	84.73		5.9
19/09/2001	06:45	314.75	4000		26.5	350	277			3500	7.41	3560	7.47	510	6.59	144		70.85	85.43		6.2
	08:45	316.75	4000		27	350	274			3460	7.44	3530	7.49	505	6.63	144		70.85	85.40		6.2
	09:30	317.5	4000		26.9	345	270			3450	7.4	3530	7.47	504	6.58	143		70.70	85.39		6.1
	Rinsed with tap water 30 min																				
	09:50	CWF	4000		20.2	520	479										1.28				
		Rinse 30 min spongeball recycle																			
	10:40	CWF	4000		20	540	500										1.22				
	1 Hour HNO_3 clean - 30 min rinse																				
	12:45	CWF	4000		20	540	500										1.22				
	STPP and EDTA clean 1 hour. 30 min water rinse																				
	16:35	CWF	4000		20	560	518										1.23				
Start	16:45	317.5																			
	19:45	320.5	4000		27	385	301	Tank 2		3340	7.46	3410	7.54	496	6.68	150		71.96	85.15		6.7
								1436	6.3												
20/09/2001	06:45	331.5	4000		27	390	305			3330	7.48	3410	7.59	510	6.74	160		70.91	84.68		6.9
	08:30	333.25	4000		27	380	297			3360	7.53	3380	7.57	503	6.76	150		71.70	85.03		6.6
	10:00	334.75	4000		27	385	301	Tank 1		3300	7.48	3390	7.56	498	6.72	160		70.64	84.91		6.8
	12:30	337.25	4000		28.4	395	297	1430	6.34	3340	7.57	3390	7.59	499	6.81	154		71.95	85.06		6.9
	Rinsed with tap water 30 min																				
	13:25	CWF	4000		20.2	660	608										1.2				
Start	13:30	337.25																			

	17:00	340.75	4000		27.6	390	300	Tank 2		3160	7.28	3300	7.41	463	6.46	180		68.42	85.35		7.1
	22:00	345.75	4000		27.5	395	305	1436	6.3	3210	7.36	3310	7.46	452	6.49	183		68.34	85.92		7.2
21/09/2001	06:45	354.5	4000		27.5	380	293	Tank 3		3350	7.39	3410	7.43	489	6.54	164		69.85	85.40		6.8
	09:00	356.75	4000		27.7	385	295	1555	6.35	3360	7.37	3450	7.44	485	6.51	162		70.38	85.57		6.8
	11:00	358.75	4000		28	385	293			3340	7.32	3470	7.44	491	6.53	160		70.64	85.30		6.8
	14:15	362	4000		28.5	385	289			3410	7.37	3520	7.46	509	6.58	156		71.16	85.07		6.8
	Rinsed with tap water 30 min																				
	15:05	CWF	4000		20.9	550	499										1.26				
	Preserve with formaldehyde																				
25/09/2001	Rinsed with tap water 30 min																				
	11:40	CWF	4000		21.5	590	528	Tank 2													
Start	11:45	362						1335	6.28								1.26				
		364.75	4000		26.2	390	311			3190	7.19	3260	7.29	638	7.73	155		71.56	80.00		6.8
		365.75	4000		27	395	309			3150	7.26	3230	7.35	443	6.34	159		71.30	85.94		6.9
		372.75	4000		27.5	400	309			3110	7.4	3220	7.5	438	6.53	170		70.18	85.92		7.1
26/09/2001	Rinsed with tap water 30 min																				
	10:00	CWF	4000		21.4	590	529	Tank 1													
Start	10:00	372.75						1359	6.3								1.28				
	11:00	373.75	4000		25.3	395	323			3000	7.31	3130	7.45	417	6.46	160		71.17	86.10		6.9
	12:30	375.25	4000		27.7	415	318			3010	7.23	3120	7.35	421	6.34	170		70.94	86.01		7.3
	16:00	378.75	4000		28.7	420	314			3060	7.27	3160	7.38	418	6.35	164		71.92	86.34		7.3
	21:15	384	4000		28.5	415	312			3060	7.31	3160	7.42	423	6.41	180		69.75	86.18		7.4
27/09/2001	10:00	396.75	4000		28.2	410	310			3070	7.44	3160	7.52	416	6.55	180		69.49	86.45		7.4
	11:45	398.5	4000		28.3	410	310	Tank 2		3080	7.38	3140	7.45	414	6.46	160		71.93	86.56		7.1
	13:30	400.25	4000		28.2	405	307	1350	6.25	3110	7.41	3190	7.52	425	6.51	155		72.32	86.33		7.0
	Rinsed with tap water 30 min																				
	14:45	CWF	4000		21.4	565	507										1.26				
	Preserve with formaldehyde																				
01/10/2001	Rinsed with tap water 30 min																				
	12:15	CWF	4000	3500	22	610	540										1.25				
	Sponge ball cleaning for 30 min (5 min cycle)																				
	13:15	CWF	4000	3500	22	610	540										1.25				
	Wash 1 hour with 2% Citric acid - pH 4.5 (NH4OH)																				
		CWF	4000	3500	22	620	548										1.25				
	Wash 1 hour with 2% Citric acid - pH 4.5 (NH4OH)																				
		CWF	4000	3500	22	620	548										1.25				
	Wash 1 hour with 0.5% P3 50																				
		CWF	4000	3500	22	620	548														

	Preserve with 0.25% NMBS																			
02/10/2001	Rinsed with tap water 30 min																			
		CWF	4000	3500	21.5	620	555										1.25				
	Wash 1 hour with 2% Citric acid - pH 4.5 (NH4OH)																			
		CWF	Pump not functioning. No pressure														0.45				
	Preserve with 0.25% NMBS																			
	Take pump in for repairs.																			
09/10/2001		CWF	4000	3300	22	635	562										1.25				
	Wash 1 hour with 0.1% EDTA, 0.2% SLS @ pH 11.8 (NaOH)																			
	Rinsed with tap water 30 min																			
		CWF	4000	3300	22	655	579										1.25				
	Preserve with 0.25% NMBS																			
10/10/2001	Rinsed with tap water 30 min																			
		CWF	4000	3300	22	640	566										1.25				
	Wash 1 hour with 0.1% EDTA, 0.2% SLS @ pH 11.8 (NaOH)																			
	Rinsed with tap water 30 min																			
		CWF	4000	3300	22.5	670	586	Tank 3									1.25				
Start	12:00	401	4000	3300	23.5	410	350	1360	6.38	2880	7.52	2970	7.35	483	6.69	148		73.48	83.23	7	7.0
	14:00	403	4000	3300	27.5	435	336			2930	7.3	3030	7.43	486	6.54	170		71.90	83.41	7.5	7.6
	16:00	405	4000	3300	28.5	440	330			2970	7.25	3040	7.4	482	6.4	168		72.37	83.77	7.5	7.6
11/10/2001	08:00	Fan belt broken. Pump not running. Replace. Approximately 8 hours run during night as measured on feed tank level.																			
	09:30	413	4000	3300	25.75	395	319			3110	7.32	3170	7.41	530	6.6	166		70.41	82.96	7	7.0
	Rinsed with tap water 30 min																				
		CWF	4000	3300	22.5	620	542										1.25				
		414	4000	3200	26	415	333	Tank 2 + 3		2950	7.21	3040	7.35	491	6.45	182		69.51	83.36		7.5
		416	4000	3200	29	440	326	1340	6.29	2970	7.25	3070	7.36	482	6.36	180		70.97	83.77		7.8
12/10/2001	07:40	435	4000	3200	29	420	311			3100	7.27	3180	7.38	497	6.42	164		71.92	83.97		7.3
	09:45	437	4000	3200	29.5	420	307			3100	7.27	3190	7.4	503	6.46	161		72.29	83.77		7.3
	10:55	438	4000	3200	30	425	306			3140	7.29	3240	7.3	490	6.5	156		73.15	84.39		7.3
	Rinsed with tap water 30 min																				
	11:40	CWF	4000	3300	23.5	640	546										1.25				
	Preserve with 0.25% NMBS																				
15/10/2001	Rinsed with tap water 30 min							Tank 1													
	13:10	CWF	4000	3200	22	640	566	1360	6.25								1.25				
	13:30	439	4000	3250	20	360	333			2960	7.52	3080	7.6	480	6.99	136		72.58	83.78		6.2
	14:20	440	4000	3200	25	400	329			3030	7.49	3070	7.57	480	6.77	140		74.07	84.16		6.8
16/10/2001	07:25	457	4000	3200	29	410	304			3200	7.62	3280	7.64	523	6.96	130		75.93	83.66		6.8
	14:00	463	4000	3400	33	400	263			4270	8.05	4220	8.08	832	8.52	110		78.43	80.52		6.4
	Rinsed with tap water 30 min																				

	Switch plant off. No leachate available due to no balanced leachate.																				
17/10/2001	Rinsed with tap water 30 min																				
	08:45	CWF	4000	3200	22.5	645	564										1.25				
	09:50	464	4000	3200	23.5	400	341			2930	7.44	2990	7.56	480	6.66	160		71.43	83.62		7.0
	07:30	487	4000	3200	38	440	244			3380	7.8	3590	7.89	684	7.81	142		75.60	79.76		7.3
	08:50	488	4000	3300	31.5	420	289			3200	7.7	3260	7.74	590	7.2	152		73.43	81.56		7.2
	Rinsed with tap water 30 min							Tank 1													
	09:30	CWF	4000	3200	23	595	514	1360	6.31								1.2				
	09:40	489	4000	3250	29	420	311			3090	7.63	3170	7.72	489	7.36	148		73.94	84.17		7.1
	10:40	494	4000	3250	29.5	420	307			3050	7.55	3160	7.71	500	6.85	148		73.94	83.61		7.1
19/10/2001	Power failure.																				
	Rinsed with tap water 30 min																				
	09:20	CWF	4000	3250	22	590	522										1.25				
	Chemical clean with EDTA, SLS, pH 11.5 (NaOH)																				
		CWF	4000	3250	22	590	522										1.25				
	Preserve membrane with 0.25% SMBS																				
	No leachate available																				
23/10/2001	Rinsed with tap water 30 min																				
		CWF	4000	3200	22	605	535	Tank 2 + 3													
	08:50	495	4000	3250	21	395	358	1400	6.25	2850	7.52	3000	7.69	710	6.65	170		69.91	75.09		7.1
	13:00	500	4000	3300	27.75	430	330			2800	7.58	2890	7.71	605	6.39	188		69.58	78.39		7.7
	Rinsed with tap water 30 min																				
		CWF	4000	3300	22	550	487														
	Sponge ball cleaning for 30 min (5 min cycle)																				
		CWF	4000	3300	22	560	495														
	Clean with 0.2% phosphoric acid for 1 hour																				
		CWF	4000	3300	22	590	522														
	Clean with STTP, EDTA at pH 10.8																				
		CWF	4000	3300	22	590	522														
	Preserve with 0.25% SMBS																				
24/10/2001	Rinsed with tap water 30 min																				
		CWF	4000	3200	22	590	522														
STOP and do batch run.																					
		CWF	4000	3800	22	620	548														
	Preserve membrane with 0.25% SMBS																				

APPENDIX I

EXPERIMENTAL CONDITIONS USING POLYAMIDE MEMBRANES FOR THE TREATMENT OF THE MUNICIPAL SOLID WASTE LEACHATE AT BISASAR ROAD

CWF flux conducted with Durban tap water					Membrane area		0.81	m2		Membrane	AFC99	Page No		Dosing concentration		11	mg/l	Permatreat 391		0.5 mg/l make-up	
																				Dosing rate	
Date	Time	Time	Pressure		Temp	Flux	Flux Normal	Feed		Feed in tank		Brine to waste		Product		Brine		Recovery	Rejection	Measured	Calculated
		(h)	(kPa) in	(kPa) out	(deg C)	(ml/min)	(l/m2.d)	(mS/cm)	(pH)	(mS/cm)	(pH)	(mS/cm)	(pH)	(mS/cm)	(pH)	(ml/min) out	(l/5 s)	(%)	(%)	(ml/min)	(ml/min)
24-Oct-2001	CWF	0	4000	3800	22.25	980	1862										1.27				
25-Oct-2001	CWF	0	4000	3800	22	970	1854	Tank 1									1.25				
	09:00	0	4000	3800	22	680	1300	13.88	6.58	15.4	6.92	16.1	6.89								15.0
	10:00	1	4000	3800	25.5	660	1159			18.68	7.06	19.28	7.07	0.22	5.45				98.82	15	14.5
	11:00	2	4000	3800	27	610	1030			21.4	7.24	22.3	7.28	0.32	5.9	234		72.27	98.50	15	18.6
	12:00	3	4000	3800	27.25	575	965			23.4	7.27	24.2	7.35	0.4	6.04	222		72.15	98.29	15	17.5
	13:00	4	4000	3850	27.5	550	917			24.7	7.3	25.5	7.38	0.45	6.13	214		71.99	98.18	13.5	16.8
	14:00	5	4000	3850	28	530	872			25.7	7.32	26.8	7.4	0.5	6.24	210		71.62	98.05	13.5	16.3
	15:00	6	4000	3850	28	515	847			26.9	7.37	27.7	7.42	0.54	6.32	208		71.23	97.99	13.5	15.9
	16:00	7	4000	3850	28	490	806	Tank 2		27.8	7.36	28.8	7.44	0.59	6.41	208		70.20	97.88	13.5	15.4
	17:00	8	4000	3850	28	480	789	13.49	7.05	28.5	7.45	29.3	7.53	0.67	6.67	196		71.01	97.65	13.5	14.9
26-Oct-2001	08:00	23	4000	3800	28.5	400	649			33.3	7.74	34.1	7.75	0.95	8.59	176		69.44	97.15	11.5	12.7
Rinse 30 min	CWF	23	4000	3800	22.5	740	1398										1.29				
	Preserve with 0.25 % SMBS						0														
29-Oct-2001	CWF	23	4000	3900	23	845	1577	Tank 3									1.25				
	10:30	28	4000	3800	22.5	390	737	13.9	7.01	32.3	7.83	32.8	7.84	0.8	8.76	165		70.27	97.52	12	12.2
30/Oct/2001	09:00	51	4000	3850	29.5	370	584			35	7.78	35.8	7.82	0.9	8.86	149		71.29	97.43		11.4
Rinse 30 min	CWF	53	4000	3800	22	710	1357										1.25				
	10:00	53	4000	3850	24	380	692			34.9	7.77	34.9	7.89	0.5	8.92	147		72.11	98.57		11.6
31/Oct/2001	10:00	77	5190	5000	30	510	793			35.9	7.77	37	7.76	0.8	8.78	110		82.26	97.77		13.6
Rinse 30 min	CWF	77	4000	3800	24	680	1239	Tank 3									1.25				
	11:30	77	4000	3820	29	350	560	13.9	7.23	35	7.78	35.8	7.83	0.8	8.73	150		70.00	97.71	14.5	11.0
1/Nov/2001	09:45	100	4100	3900	30	290	451			33	7.83	33.7	7.85	0.9	8.88	148		66.21	97.27		9.6
Rinse 30 min	CWF	100	4000	3850	23	590	1101										1.3				
	Preserve with 0.25% SMBS																				
	NO LEACHATE																				
21/Nov/2001	CWF	100	4000	3850	25	775	1378										1.3				
	Wash 1 hour with 0.25% 10 - pH 11,36						0														
Rinse 30 min	CWF	100	4000	3850	25	790	1404	Tank 2									1.3				
Start	15:15	100	4000	3850	25	380	676	14	7.06	31.4	7.88	32	7.85	0.36	8.9	156		70.90	98.85	11	11.8
	16:00	101	4000	3850	27.5	410	683			31.4	7.77	32.2	7.81	0.8	8.69	170		70.69	97.45	11	12.8

	17:00	102	4000	3850	29	410	656			31	7.74	32.3	7.78	0.9	8.76	158		72.18	97.10	11	12.5
22/Nov/2001	09:00	118	4000	3850	32.5	360	520			35.3	7.77	36.1	7.78	1.24	8.71	148		70.87	96.49	11	11.2
Rinse 30 min	CWF	118	4000	3850	25	680	1209										1.3				
	10:00	118	4000	3850	30	360	560			34.9	7.78	35.2	7.82	1.1	8.9	148		70.87	96.85	10.5	11.2
Samples	12:00	120	4000	3850	31	360	544			34.7	7.75	35.5	7.76	1.2	8.85	142		71.71	96.54	10.5	11.0
	14:00	122	4000	3850	31.5	360	536	Tank 3		34.8	7.76	35.6	7.8	1.18	8.9	140		72.00	96.61	10.5	11.0
	16:00	124	4000	3850	31.5	350	521	13.9	6.96	35.1	7.73	35.9	7.75	1.21	8.81	149		70.14	96.55	10.5	11.0
23/Nov/01	08:00	140	4000	3850	31.75	340	502			36.6	7.61	37.2	7.65	1.25	8.59	146		69.96	96.58	10.5	10.7
Rinse 30 min	CWF	140	4000	3850	25	670	1191														
	Preserve with 0.25% SMBS						0														
26/Nov/2001	CWF	140	4000	3900	25	760	1351	tank 3									1.4				
	08:40	140	4000	3900	25	330	587	1380	7.05	35.9	7.81	36.1	7.8	0.8	8.88	146		69.33	97.77	9	10.5
27/Nov/2001	08:40	164	4010	3900	31	320	484			3660	7.76	37	7.85	1.3	8.9	134		70.48	99.96	10	10.0
	CWF	164	4000	3900	25	620	1102	Tank 2									1.5				
	09:40	164	4000	3900	29	340	544	12.6	7.23	36.6	7.77	36.6	7.83	0.5	8.97	140		70.83	98.63	10	10.6
28/Nov/2001	07:30	186	4010	3850	30	290	451			35.2	7.5	35	7.94	1	9.08	120		70.73	97.16	10	9.0
	CWF	186	4000	3900	25	540	960										1.4				
	Preserve with 0.25% SMBS						0														
	Lectures being held @ workshop						0														
10/Dec/2001	CWF	186	4000	3850	25.5	780	1369	Tank 1									1.3				
	10:00	186	4000	3850	27	380	642	12.4	6.98	32.3	7.76	33	7.78	0.8	8.87	144		72.52	97.52	11	11.5
11/Dec/2001	09:40	209	4000	3900	32	340	499			31.8	7.81	32.7	7.84	0.9	8.94	154		68.83	97.17	11	10.9
	CWF	209	4000	3850	25.5	580	1018	Tank3									1.4				
	10:30	209	4000	3850	31	360	544	12.89	7.2	31.8	7.81	31.4	7.87	0.5	8.9	154		70.04	98.43	11	11.3
12/Dec/2001	10:00	233	4100	3850	31.5	280	417			28.2	7.86	29.2	7.89	0.8	9.07	170		62.22	97.16	8	9.9
	CWF	233	4000	3900	26	450	780	Tank 2									1.3				
	11:00	233	4000	3900	30	300	467	12.2	6.98	28.2	7.86	28.1	7.88	0.4	8.96	142		67.87	98.58	11	9.7
13/Dec/2001	10:00	256	4100	3950	33	260	370			26.5	7.85	27	7.83	0.9	8.91	150		63.41	96.60	8	9.0
	CWF	256	4000	3900	26	390	676										1.5				
		Chemical Clean with 0.25 % 10 for 1 hr																			
	CWF	256	4000	3850	27	625	1056	Tank 1									1.5				
	12:30	256	4000	3850	30	400	622	11.9	7.09	26.5	7.85	27.1	7.91	0.6	8.88	178		69.20	97.74	11	12.7
14/Dec/2001	11:30	279	4100	3950	32	300	440			26	7.79	26.3	7.81	0.8	8.8	172		63.56	96.92	30	10.4
		CWF	4000	3850	25	435	773	Tank 2									1.5				
	12:20	279	4000	3850	30	280	436	11.8	7.06	26	7.79	25.7	7.84	0.5	8.74	126		68.97		8	8.9
15/Dec/2001	15:00	306	4100	4000	33.5	240	336			29.4	7.9	29.5	7.94	1.4	8.98	96		71.43		8	7.4
	CWF	306	4000	3850	25	370	658	Tank 3													
	15:40	306	4000	3850	30	250	389	12.2	6.99	29.4	7.9	30.3	7.95	0.8	8.99	100		71.43		8	7.7

16/Dec/2001	15:30	330	4100	3900	35	200	267			30.2	7.88	29.6	7.9	1.3	8.95	90		68.97			
	CWF	330	4000	3850	26	310	537														
	membrane preservation																				
18/Dec/2001	CWF	330	4000	3850	25	330	587										1.25				
clean with	0.1 % 10																				
	CWF	330	4000	3850	25	570	1013	Tank 1 & 2													
	12:00	330	4000	3850	24	315	574	11.9	6.98	29.4	8.04	29.3	7.98	0.5	9.1	130		70.79			
19/Dec/2001	10:50	354	4150	4100	35	205	273			29	7.84	29.1	7.85	1.1	8.8	114		64.26		8	7.0
	CWF	354	4000	3850	26	420	728										1.4				
	11:30	354	4000	3850	31	230	348	13.1	7.07	29	7.84	29.4	7.9	0.8	8.96	100		69.70			
20/Dec/2001	11:30	378	4050	3950	34	125	172			28	7.99	27.9	8	1.4	9.04	94		57.08		8	4.8
	CWF	378	4000	3850	25.5	230	404	Tank 1									1.4				
	12:00	378	4000	3850	32	160	235	13.1	7.16	28	7.99	28.5	8.04	1.2	9.1	65		71.11		8	5.0
21/Dec/2001	15:00	405	4050	3850	35	85	113			26.5	8.13	26.4	8.15	1.2	9.23	60		58.62		8	3.2
	CWF	405	4000	3850	25	270	480										1.4				
	Clean with 0.1 % ultrasil 10																				
	CWF	405	4000	3850	25	500	889										1.4				
	membrane preservation																				
7/Jan/2002	CWF	405	4000	3850	25	700	1244	Tank 2									1.4				
	09:00	407	4000	3850	26	410	711	13.1	7	25.2	8.1	25.5	8.1	0.6	9.23	180		69.49		11	13.0
8/Jan/2002	09:30	431	4100	3950	35	200	267			29.2	8.18	28.9	8.2	1.2	9.32	150		57.14		8	7.7
	CWF	431	4000	3850	25	380	676	Tank 1									1.4				
	10:30	431	4000	3800	36	250	322	13.5	7	29.2	8.18	29.5	8.25	0.9	9.32	120		67.57		9	8.1
9/Jan/2002	10:30	455	4100	3950	36	150	193			26.5	8.23	26.6	8.29	1.2	9.31	98		60.48		8	5.5
	CWF	455	4000	3850	26	280	485	Tank 2									1.3				
	11:00	455	4000	3850	36	170	219	13.1	7.1	26.5	8.23	26.3	8.28	1	9.31	90		65.38			
10/Jan/2002	09:00	477	4100	3950	36	170	219													8	3.7
	CWF	477	4000	3850	26	270	468														
	10:00	477	4100	3850	35	190	253	13.1	7.1	29.9	8.22	29.8	8.27	0.8	9.28	85		69.09		8	6.1
11/Jan/2002	10:00	501	4100	3750	36	155	200			30.1	8.25	29.9	8.28	1.3	9.27	56		73.46			
	CWF	501	4000	3850	26	260	451										1.4				
29/Jan/2002	After preservation						0														
	CWF	501	4000	3850	26.5	430	736	Preservative water brown on rinsing													
Clean membrane with HCl at pH 2																					
	CWF	501	4000	3850	26.5	700	1198														
Clean with 0.1% Ultrasil 10 for 0.5 h																					
	CWF	501	4000	3850	26.5	730	1249														

Clean with 0.25% 10 for 0.5 h																					
	CWF	501	4000	3850	26.5	730	1249														
After preservation							0														
	CWF	501	4000	3850	26.5	730	1249														

APPENDIX J

EXPERIMENTAL CONDITIONS USING SPIRAL WRAP RO MEMBRANES FOR THE TREATMENT OF THE MUNICIPAL SOLID WASTE LEACHATE AT BISASAR ROAD

TABLE A-1: Log sheet from reverse osmosis treatment of leachate pilot plant

Date	Time	Time	Pressure		Temp	Flux	Flux normal	Feed in tank		Brine to waste		Product		Brine	Recovery	Rejection	Cartridge filter pressure		Sand filter
		(h)	kPa in	kPa out	deg C	l/10s	l/m2.d	mS/cm	pH	mS/cm	pH	mS/cm	pH	l/10s			kPa in	kPa out	kPa
30-Jan-02	Install pilot																		
31-Jan-02	CWF	0	3000	2350	27	3.35	989.09	0.11		0.31		0.0012		1.95	63.21	98.91	300	325	75
31-Jan-02	12h30	0	4000	3400	26.3	2.79	838.93	12.3	7.65	23.7	7.51	0.35	7.99	2.42	53.55	97.15	300	315	75
	13h30	1	4000	3400	29.3	2.87	796.08	11.3	7.31	23.3	7.48	0.44	6.57	2.36	54.88	96.11	325	300	65
	14h30	2	4000	3400	30.9	2.92	773.65	11.3	7.37	23	7.43	0.5	6.44	2.3	55.94	95.58	350	300	70
	15h30	3	4000	3400	30.6	2.87	767.10	11.3	7.38	22.7	7.47	0.49	6.5	2.38	54.67	95.66	325	275	65
	Stop plant																		
	CWF	3	3000	2350	27.1	3.34	983.55	0.111		0.341		0.0012		1.94	63.26	98.92			
	Preservation																		
01-Feb-02	CWF	3	3000	2350	27	3.5	1033.38							1.9	64.81		325	290	85
13-Feb-02	CWF	3	3000	2350	26	3.25	984.82							2.2	59.63		310	300	85
	10h00	3	4000	3400	25	2.95	916.83	10.4		18.45		0.25		2.3	56.19	97.60	325	290	80
	11h00	4	4000	3415	25	2.6	808.06	10.25		19.33		0.34		2.5	50.98	96.68	300	260	80
	12h00	5	4000	3500	26	2.64	799.98			18.91		0.33		2.63	50.09		290	250	80
	CWF	5	3000	2350	26	3.175	962.09							2.075	60.48		300	270	85
	Preservation																		
26-Feb-02	CWF	5	3000	2350	26	3.4	1030.27							2.04	62.50		310	300	85
	09h50	5	4000	3400	26.3	2.645	795.33	11.29		21.1		0.2		2.45	51.91	98.23	310	275	80
	10h50	6	4000	3420	26.9	2.545	753.39	11.32	7.85	21.8		0.2		2.67	48.80	98.23	290	245	80
	11h50	7	4000	3500	27.2	2.38	699.00			20.8	7.89	0.2	7.17	2.9	45.08		275	225	80
	CWF	7	3000	2390	26	2.475	749.98							2.86	46.39		300	260	85
	Rinse and preservation																		

Date	Time	Time	Pressure		Temp	Flux	Flux normal	Feed in tank		Brine to waste		Product		Brine	Recovery	Rejection	Cartridge filter pressure		Sand filter
		(h)	kPa in	kPa out	deg C	l/10s	l/m2.d	mS/cm	pH	mS/cm	pH	mS/cm	pH	l/10s			kPa in	kPa out	kPa
27-Feb-02	CWF	7	3000	2350	26.2	2.45	738.60							2.9	45.79		300	280	90
	Chemical clean with 0.1% Ultrasil 10 (100g Ultrasil 10/100l water)																		
	CWF	7	3000	2360	26.2	3.04	916.46							2.41	55.78		350	325	82
	Changed 1 cartridge filter and backwashed sand filter																		
03-Mar-02	CWF	7	3000	2390	26	3.29	996.94							1.94	62.91		310	300	87.5
	09h30	7	4000	3400	26	2.63	796.95	11.87	8	21.3	7.86	0.38	7.22	2.57	50.58	96.80	305	300	87.5
	10h30	8	4000	3400	26.5	2.46	735.88	11.4	7.93	20.7	7.89	0.2	7.32	2.67	47.95	98.25	300	290	80
	11h30	9	4000	3400	27.1	2.45	721.46	11.4	7.92	20.5	7.9	0.2	7.53	2.62	48.32	98.25	300	300	80
	12h30	10	4000	3400	27.5	2.54	740.07	11.4	7.93	20.8	7.9	0.2	7.7	2.6	49.42	98.25	305	300	78
	CWF	10	3000	2400	27.5	2.41	702.19							2.89	45.47		310	300	89
	Rinse and preservation																		
13-Mar-02	CWF	10	3000	2350	25.1	2.88	892.84							2.42	54.34		350	310	90
	Clean with 0.1% Ultrasil 10																		
	CWF	10	3000	2350	25.1	3.21	995.15							2.08	60.68		340	310	85
	10h10	10	4000	3400	25.1	2.53	784.34	12.4	8.05	22	7.24	0.2	7.41	2.64	48.94	98.39	350	310	80
	11h10	11	4000	3400	25.5	2.37	727.37	12.4	7.96	22	7.85	0.2	7.49	2.65	47.21	98.39	340	300	80
	12h10	12	4000	3400	26.1	2.52	761.66	12.5	7.97	22.1	7.86	0.2	7.54	2.71	48.18	98.40	350	310	80
	13h10	13	4000	3400	26.6	2.41	719.05	12.4	7.91	22.3	7.85	0.2	7.51	2.71	47.07	98.39	340	300	80
	14h10	14	4000	3400	26.7	2.44	726.10	12.3	7.95	22.3	7.83	0.3	7.62	2.73	47.20	97.56	350	300	80
	CWF	14	3000	2400	26.7	2.65	788.59							2.51	51.36		350	310	85
	Rinse and preservation																		

Date	Time	Time	Pressure		Temp	Flux	Flux normal	Feed in tank		Brine to waste		Product		Brine	Recovery	Rejection	Cartridge filter pressure		Sand filter
		(h)	kPa in	kPa out	deg C	l/10s	l/m2.d	mS/cm	pH	mS/cm	pH	mS/cm	pH	l/10s			kPa in	kPa out	kPa
19 Mar 02	CWF	14	3000	2400	27.6	2.89	839.80							2.22	56.56		350	310	87.5
	13h10	14	4000	3400	27.6	2.82	819.46	8.9	8.04	18	7.95	0.1	8.44	2.26	55.51	98.88	350	310	80
	14h10	15	4000	3400	29.2	2.79	776.06	8.2	7.77	18.2	7.59	0.36	7.58	2.35	54.28	95.61	330	290	75
	15h10	16	4000	3400	27.2	2.8	822.35	8.62	7.63	18.58	7.6	0.37	7.53	2.33	54.58	95.71	330	290	75
	16h10	17	4000	3400	27	2.73	806.04	8.84	7.64	16.82	7.62	0.38	7.64	2.48	52.40	95.70	320	290	75
	Backwashed sand filter																		
	CWF	17	3000	2350	27	2.89	853.28							2.3	55.68		350	300	85
	Preservation with 0.1% NMBS																		
20-Mar-02	CWF	17	3000	2350	26.8	3.03	899.32							2.1	59.06		350	310	90
	10h00	17	4000	3400	26	3	909.06	8.64	7.58	16.74	7.51	0.3	8.04	2.21	57.58	96.53	350	310	82.5
	11h00	18	4000	3400	25.8	2.8	852.81	8.7	7.62	17.48	7.58	0.29	8.12	2.28	55.12	96.67	330	290	80
	CWF	18	3000	2350	25.8	2.88	877.18							2.38	54.75		340	300	82.5
	Backwashed sandfilter and preserved with 0.25% NMBS																		
21-Mar-02	CWF	18	3000	2350	25.9	3.12	947.85							2.07	60.12		350	310	87.5
	09h05	18	4000	3400	26.9	2.72	805.20	12.34	7.67	20.7	7.52	0.51	6.89	2.29	54.29	95.87	350	300	80
	10h05	19	4000	3400	28.9	2.83	793.78	12.79	7.49	23.1	7.57	0.57	7.15	2.54	52.70	95.54	350	300	80
	11h05	20	4000	3400	29.5	2.41	664.74	13.96	7.93	24.4	7.54	0.67	6.87	2.64	47.72	95.20	330	290	85
	12h05	21	4000	3400	29.2	2.34	650.89	14.28	7.47	23.9	7.51	0.69	6.91	2.75	45.97	95.17	320	280	75
	13h05	22	4000	3400	30	2.25	611.87	14.35	7.52	23.6	7.5	0.74	6.92	2.82	44.38	94.84	320	275	75
	14h05	23	4000	3400	29.4	2.15	594.70	14.38	7.48	23.3	7.51	0.76	7.01	2.89	42.66	94.71	310	260	72.5
	15h05	24	4000	3400	29.4	2.11	583.64	14.3	7.49	22.8	7.5	0.76	7.1	2.98	41.45	94.69	310	260	77.5

Date	Time	Time	Pressure		Temp	Flux	Flux normal	Feed in tank		Brine to waste		Product		Brine	Recovery	Rejection	Cartridge filter pressure		Sand filter
		(h)	kPa in	kPa out	deg C	l/10s	l/m2.d	mS/cm	pH	mS/cm	pH	mS/cm	pH	l/10s			kPa in	kPa out	kPa
	CWF	24	3000	2350	29.1	2.78	775.44							2.43	53.36		320	280	85
	Cleaned with 0.25% Ultrasil 10, including rinse 15 min						0.00												
	CWF	24	3000	2350	28.1	3.22	923.19							1.98	61.92		290	210	85
	Backwashed sand filter and preserved with 0.25% NMBS																		
12-Apr-02	CWF	24	3000	2350	24.2	2.98	944.68							2.27	56.76		350	300	87.5
	Acid clean with HCl pH2 30min																		
	Rinse 15min																		
	Clean with 0.25% Ultrasil 10 30 min																		
	Rinse 15min																		
	CWF	24	3000	2350	24.9	3.27	1018.83							1.87	63.62		350	310	87.5
	Acid clean with HCl pH2 30min																		
	Rinse 15min																		
	Clean with 0.25% Ultrasil 10 30 min																		
	Rinse 15min																		
	CWF	24	3000	2350	25.1	3.33	1032.35							1.9	63.67		350	310	85
	Preservation with 0.25% NMBS																		
03 May 02	CWF		3000	2350	22.9	3.31	1082.73							1.89	63.65		350	320	87.5
	09h55	24	4000	3400	23.8	2.81	899.52	11.13	7.53	20.3	7.45	0.34	6.79	2.26	55.42	96.95	350	310	82.5
	10h55	25	4000	3400	23	2.84	926.78	11.24	7.43	21.1	7.48	0.39	6.83	2.43	53.89	96.53	350	300	82.5
	11h55	26	4000	3400	23	2.57	838.67	11.34	7.46	20.8	7.5	0.46	7.21	2.55	50.20	95.94	350	300	80
	12h55	27	4000	3400	23.7	2.5	802.23	11.72	7.57	20.3	7.39	0.48	7.65	2.66	48.45	95.90	350	300	80
	CWF		3000	2400	22.8	2.93	960.70							2.24	56.67		350	310	85

Date	Time	Time	Pressure		Temp	Flux	Flux normal	Feed in tank		Brine to waste		Product		Brine	Recovery	Rejection	Cartridge filter pressure		Sand filter
		(h)	kPa in	kPa out	deg C	l/10s	l/m2.d	mS/cm	pH	mS/cm	pH	mS/cm	pH	l/10s			kPa in	kPa out	kPa
	Rinse																		
	Preserve with 0.25% NMBS																		
04 May 02	CWF		3000	2350	22.5	3.25	1073.20							1.99	62.02		350	310	90
	10h15	27	4000	3400	23	2.68	874.57	11.51	7.71	16.31	7.55	0.37	8.04	2.13	55.72	96.79	350	300	80
	11h15	28	4000	3400	23.4	2.59	837.15	11.47	7.63	21.1	7.61	0.19	7.84	2.57	50.19	98.34	340	300	80
	12h15	29	4000	3400	24	2.63	837.82	11.64	7.56	20.8	7.54	0.45	7.97	2.63	50.00	96.13	350	300	80
	CWF		3000	2350	23.8	3.03	969.95							2.16	58.38		350	310	85
	Rinse and backwash sandfilter																		
	Preserve with 0.25% NMBS																		
21 May 02	CWF	29	3000	2400	23.4	2.81	908.26							2.29	55.10		310	300	90
	12h30	29	4000	3400	23.7	2.46	789.39	12.92	8.01	16.02	7.56	0.39	6.75	2	55.16	96.98	310	310	85
	13h30	30	4000	3400	25.2	2.45	757.63	13.22	7.68	23.9	7.57	0.53	6.84	2.57	48.80	95.99	310	300	85
	14h30	31	4000	3400	25.8	2.44	743.16	13.78	7.49	22.8	7.52	0.55	6.84	2.76	46.92	96.01	310	290	85
	15h30	32	4000	3400	25.4	2.31	710.75	13.48	7.54	22.6	7.49	0.56	6.87	2.92	44.17	95.85	310	290	85
	CWF	32	3000	2350	23.8	2.79	893.12							2.38	53.97		310	290	90
	Rinse																		
	Preservation with 0.25% NMBS																		
23 May 02	CWF	32	3000	2400	20.3	2.91	1 011.00							2.3	55.85		310	300	90
	Preservation with 0.25% NMBS																		

Membrane area is approximately 27.8 square metres

Clean water flux was conducted with tap water

APPENDIX K

COST BREAKDOWN

1. **TUBULAR CELLULOSE ACETATE RO**

Capital cost (250 kℓ/d)		R 1,95 million
Operational costs		
• Pretreatment and CIP chemicals		R 4,78/kℓ
• Energy		R 0,86/kℓ
• Sponge ball replacement		R 0,03/kℓ
• Maintenance		R 2,28/kℓ
• Membrane replacement		R 3,50/kℓ
	Total	R11,45/kℓ

Membrane lifetime 2 years

2. **TUBULAR POLYAMIDE RO**

Capital cost (250 kℓ/d)		R 6,5 million
Operational costs		
• Pretreatment & CIP chemicals		R 4,95/kℓ
• Energy		R 1,71/kℓ
• Maintenance (labour)		R 2,28/kℓ
• Membrane replacement		R 7,30/kℓ
	Total	R16,24/kℓ

Membrane lifetime 18 months
1 British pound = R16,20

3. **DISC TUBE RO**

Capital cost (250 kℓ/d plant)		R 6,2 million
Operational costs		
• Pretreatment & CIP chemicals		R12,71/kℓ
• Energy		R 2,02/kℓ
• Labour		R 0,29/kℓ
• Parts		R 3,10/kℓ
• Membrane replacement		R 8,53/kℓ
	Total	R26,65/kℓ

1 Euro = R9,7

4. **GRAHAMTEK SPIRAL RO PLANT**

Capital cost		R 0,560 million
Operational costs		
• CIP chemicals		R 1,20/kℓ
• Energy		R 1,42/kℓ
• Membrane replacement		R 0,50/kℓ
• Operator		R 0,18/kℓ
• Sandfilter		R 0,07/kℓ
• Cartridge filter replacement		R 0,07/kℓ
• Media for Exfetron		R 0,07/kℓ
	Total	R 3,51/kℓ

Membrane lifetime 2 years.