THE EFFECT OF NATURAL ORGANIC MATTER ON ULTRAFILTRATION AND REVERSE OSMOSIS MEMBRANE PERFORMANCE AT KOMATI POWER STATION

By

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SUMMARY

Komati Power Station has installed a membrane plant consisting of ultrafiltration, double pass reverse osmosis and continuous electro-deionisation to treat cooling tower blowdowns in order to produce demineralised water and to conduct sidestream chemistry control of the cooling water circuit. This plant has replaced the existing ion-exchange plant that was used for the production of demineralised water and thus serves to reduce the loading of mobile salts in the ash dam (90% reduction) by eliminating regeneration effluent from the ion-exchange plant.

Due to oil contamination in the cooling water circuit (when oil from oil coolers leaks into the cooling water), the membrane plant was also designed to operate on raw water from either the Nooigdedacht or the Vygeboom Dam or a blend of both dams. This is considered to be an emergency intervention under abnormal conditions to prevent possible irreversible fouling of the membranes due to oil in the cooling water. The Nooigtedach Dam water contains high concentrations of organic matter and is also enriched with nutrients due to raw sewage influent into the Dam water. This poses a challenge with regard to treatment of the high fouling feed water on the membrane plant.

Natural organic matter in water has the ability to foul reverse osmosis membranes. This adversely affects the operation of the reverse osmosis process. However, very little information is available regarding the fouling characteristics of natural organic material in the raw and cooling water at Komati Power Station for the reverse osmosis membranes. Therefore, a pilot study was undertaken to determine the influence of natural organic matter on membrane fouling, to optimise the process for the removal of natural organic matter and to assess the ability of two different reverse osmosis membranes to effectively treat the high fouling feed water at Komati Power Station.

The ability of a polyethersulphone hollow-fibre ultrafiltration membrane system was first evaluated to remove natural organic matter in the feedwater, by conducting pilot tests, initially without coagulation of the raw water and thereafter with in-line coagulation for organics removal. Jar tests were conducted in the laboratory to determine the most suitable coagulant and dosage for turbidity and natural organic matter removal. Various coagulants were tested and, based on the results of the jar tests, a coagulant (U3000) was identified based on optimal removal of both total organic carbon and turbidity at a dosing level of 20 mg/L.

During the operation of the ultrafiltration pilot plant, permeate flow; feed pressure and feed temperature were monitored. Performance of the ultrafiltration membrane was monitored in terms of flux versus time for operation with and without a coagulation process. The results indicated that there was very little total organic carbon removal (maximum removal of 4%) without coagulation and a slight decrease in flux. The flux declined as a result of fouling but could be recovered by performing hydraulic backwashes and CEB procedures. Permeate flux, however, could be maintained at about 90 Lmh (from 642 hours of operation). Since most of the organics passed through the ultrafiltration membrane, it was concluded that the loss in flux was due to colloidal fouling of the membrane. This was observed when the operation was carried out using raw water as feed as well as when cooling water was used.

The total organic carbon removal increased to 30% when the plant was operated with in-line coagulation. The flux remained relatively stable during the first 600 hours of operation and only decreased significantly during the last 200 hours of operation as a result of fouling. The reduction in flux prior to cleaning was less than the 15% (maximum flux decline of 9.9% during the test period) which is acceptable according to the industry norm of 15%. It appeared that flux could be maintained at around 90 Lmh which was about the same as when no coagulant was applied. The 30% total organic carbon reduction that was obtained was not sufficient to reduce the organics to the level of 6 mg/L dissolved organic carbon that was specified by the membrane manufacturer for the standard brackish water reverse osmosis membrane.

Two reverse osmosis membranes – the standard brackish water reverse osmosis membrane (BW30-2540) and the extra-low-fouling membrane (BW30XFR-2540) – were assessed in terms of their ability to remove dissolved organic carbon, ease of cleaning of the membrane and the ability to recover flux after cleaning. This was done to establish which membrane is more suited to Komati's high-fouling feedwater.

The evaluation of the performance of the two reverse osmosis membranes was conducted using pre-treated water (filtered water after in-line coagulation, anti-scalant and biocide dosing) as well as using water that was not pre-treated. During operation (under both conditions), the normalised permeate flux, conductivity, dissolved organic carbon and organics absorbing at UV_{254} were monitored.

It was established that in terms of flux decline that the extra low-fouling membrane gave slightly superior performance to that of the standard membrane, achieving longer production runs (up to 5 days compared with 3 days achieved by the standard brackish water membrane) without requiring chemical cleaning. The low fouling membrane achieved better CWF recovery after the cleaning cycles (81.26% Lmh of the virgin membrane on the occasions when there was flux loss) compared to the standard membrane (restored to 77.35% of CWF of the virgin membrane) when using untreated feed water. This performance improved when pre-treated feed water was used and the low fouling membrane's CWF regained after the CIP was 95.89% which was within the industry norm of a flux recovery of 95%, indicating that the CIP had been effective. It was determined that the TOC rejection of the low-fouling membrane was higher (average TOC rejection of 97%, maximum TOC rejection of 99%) than that of the standard membrane (average TOC rejection of 95.3%, maximum TOC rejection of 97%).

Preliminary efforts to optimize the pre-treatment for organics removal in order to reduce organic loading for the RO membranes confirmed that the use of granular activated carbon and use of an organic scavenger resin might not be economically feasible due to the relatively quick TOC breakthrough (8910BV, approximately 18000BV and less than 18000BV for the Filtrasorb 300, Filtrasorb 400 and organic scavenger resin, respectively). Although further investigations should still be conducted, the preliminary results indicate that it would be beneficial to also identify other options that can be further investigated for optimization of organics removal at Komati Power Station.

Decline in the normalised flux as well as the evidence of biofouling were witnessed during the pilot operation suggesting that the membranes were fouled. Autopsies were performed on both membranes to identify foulants responsible for the decline in flux that was observed during the pilot study. The results did not indicate an organic foulant on the membrane surface. Biofouling should however, be monitored in the main plant as this was suspected to have resulted in the flux decline during the pilot study.

The low fouling membrane demonstrated a better capability to treat the Komati raw and cooling water and would be expected to achieve lower operating costs for the plant (CIP costs and membrane replacement costs) while achieving better organics removal and it is therefore recommended that the low-fouling membranes be used at Komati Power Station as they are superior to the standard membrane and the cost of the low-fouling membranes is comparable to that of the standard membrane. While this would provide somewhat better performance than that obtained with the standard brackish water membranes, it is proposed that further investigation into pre-treatment optimization for organics removal as well as more efficient cleaning solutions be investigated to improve the performance and economics of the main water treatment plant at Komati power Station.

Keywords: Organic fouling; biofouling, clean-in-place; flux; dissolved organic carbon; Power Station; cooling water; raw water, natural organic matter; fouling; ultrafiltration, reverse osmosis.

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CONTENTS

LIST OF ABBREVIATIONS

Alum	Aluminium sulphate
ATR	Attenuated Total Reflection
CAM	Camden Power Station
CEB	Chemically Enhanced Backwash
CEDI	Continuous Electro-Deionisation
CIP	Clean-In-Place
CNOM	Colloidal Natural Organic Matter
CWF	Clean Water Flux
DNOM	Dissolved Natural Organic Matter
DOC	Dissolved Organic Carbon
DVP	Duvha Power Station
EDX	Energy-Dispersive X-Ray Spectroscopy
EDTA	Ethylenediaminetetraacetic acid
EEMS	Excitation Emission Matrix Fluorescence Spectroscopy
FTIR	Fourier-Transform Infrared
GAC	Granular Activated Carbon
HPSEC	High-Performance Size Exclusion Chromatography
KHP	Potassium Hydrogen Phthalate
KOM	Komati Power Station
LC-OCD	Liquid Chromatography-Organic Carbon Detection
Lmh	L/m^{2}h
MWTP	Main Water Treatment Plant (Refers to Komati Power Staion Desalination Plant)
MIEX	Magnetic Ion-exchange
MWCO	Molecular Weight Cut-Off
NMR	Nuclear Magnetic Resonance
NOM	Natural Organic Matter
OC	Organic Carbon
PAC	Powdered Activated Carbon

PES	Polyethersulphone
POC	Particulate Organic Carbon
RO	Reverse Osmosis
ROSA	Reverse Osmosis System Analysis
SAC	Spectral Absorption Coefficient
SDI	Silt Density Index
SEC	Size Exclusion Chromatography
SEM	Scanning Electron Microscopy
SUVA	Specific Ultraviolet Absorbance
t_f	Final time
t_i	Initial time
TOC	Total Organic Carbon
UF	Ultrafiltration
UV	Ultraviolet
UV/Vis	Ultraviolet/Visible
ZLED	Zero Liquid Effluent Discharge

LIST OF DEFINITIONS

Cooling tower blowdowns	Refers to the amount of cooling water that must be removed from the cooling water system in order to maintain salt within specified limits
Flux	Flow (amount of permeate produced) per unit area of membrane surface per unit time.
Forebay	Refers to the chamber where the cooling tower outlet channels come together upstream of the cooling water pumps suction.
Hydraulic resistance	Hydraulic resistance, W_m, is a characteristic value for a given membrane that is used and is defined for pure solvent flux. It is given by J = change in transmembrane pressure / W_m
Hydrophilic fraction	Refers to the fraction of NOM (DOC) with the highest polarity (Polar NOM). This fraction passes through the XAD-8 resin during fractionation of NOM.
Hydrophobic fraction	Refers to the fraction of lowest polarity (Non polar NOM). This fraction is adsorbable/retained on the XAD-8 resin.
Pedogenic	Soil derived
Transmembrane pressure	Pressure required to push/pump water through a membrane. It is defined as the pressure gradient of the membrane, or the average feed pressure minus the permeate pressure.
Transphilic fraction	Refers to the fraction of intermediate polarity. This fraction passes through the XAD-8 resin but is retained on the XAD-4 resin.
Zeta potential	The zeta potential (z) is the potential drop between the shear surface and the bulk solution and is related to the particle charge and the thickness of the double layer. The thickness of the double layer is composed of a compact

	fixed layer (includes bound water) of oppositely charge ions adsorbed to the colloid surface and a diffuse layer of a mixture of charged ions. A shear surface separates the fixed and diffuse layers and represents the mobile portion of the colloid. Zeta potential is defined as $4\pi\delta q/D$, where q = charge at shear surface δ = thickness of diffuse layer D = dielectric constant of liquid. (Sawyer *et al*., 2003).

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CHAPTER 1:
INTRODUCTION

1.1 MEMBRANE TECHNOLOGY IN THE WATER TREATMENT INDUSTRY

Membrane technology has become more acceptable as a water treatment technology due to the operational advantages that it offers such as (Crozes *et al*., 1997):

- Limited use of chemicals;
- Less manpower-intensive;
- Highly automated;
- Smaller space requirements.

There are various membrane technologies available in the market and these can be classified based on the driving forces that induce transport of materials across the membranes (Taylor and Wiesener, 1999:11.14-11.15):

- Temperature gradient (e.g. membrane distillation)
- Concentration gradient (e.g. Dialysis, pervaporation, osmosis)
- Pressure gradient (e.g. RO, NF, UF, MF)
- Electrical potential (e.g. Electrodialysis, electro-osmosis)

The various technologies (in terms of advantages, disadvantages, uses and limitations) can be summarised as shown in Table 1 (Asano, 2007:430, 432, 464, 466, 468, 501, 508).

Table 1 Summary of membrane processes

Process	**Application**	**Working Principle**	**Advantages**	**Disadvantages**
Microfiltration (MF)	Removal of turbidity, bacteria and viruses (pathogens)	Separation based on size exclusion, where particles larger than the pore size are retained by the membrane and particles smaller than the pore size pass through.	• Can be used as a pre-treatment step in water treatment. • No sludge residue that require disposal. • Lower chemical consumption.	• Macromolecules not removed (removal limited to particles > ca. 0.2 µm). • Not effective in NOM removal. • Tendency to foul resulting in lower productivity. • Chemical cleaning generates chemical waste stream which requires disposal.
Ultrafiltration (UF)	Removal of turbidity, bacteria and viruses (pathogens)	Relies on size exclusion, where species larger than the pore size of the molecular weight cut-off (MWCO) are retained by the membrane and smaller species pass through.	• Efficient removal of suspended particles and colloids, turbidity, algae, bacteria, parasites and viruses for clarification and disinfection purposes. • Serves as an alternative pre-treatment to further treatment steps, such as softening and nitrate removal. • Lower chemical consumption • No sludge residue for disposal.	• Tendency to foul resulting in lower productivity and increased energy costs. • Chemical cleaning generates chemical waste stream which requires disposal.
Nanofiltration (NF)	Softening, Colour removal, NOM removal, Removal of disinfection by-products precursors. Desalination of brackish water and wastewater.	Relies on a combination of size exclusion and solution/diffusion permeation to achieve separation.	• Requires much lower energy than reverse osmosis (RO) • High rejection of divalent ions. • High rejection of NOM. • High flux.	• Fouling effects reduce performance and result in increased operational costs. • Requires chemical cleaning to restore/maintain performance.

Reverse Osmosis (RO)	Seawater desalination, brackish water desalination, softening, demineralised water production, removal of synthetic organic carbons (SOC)	Relies on a combination of size exclusion and solution/diffusion permeation to achieve separation.	• Provides a barrier to micro-organisms • water reuse (wastewater demineralization) • removes >90% of TDS	• Requires high pressure to achieve high salt rejection • Requires pre-treatment processes to minimize scaling and fouling • Requires chemical cleaning to restore/maintain performance.
Electro-deionisation (CEDI) (Flynn, 2009:8.27-8.30).	Desalination of brackish water, silica and boron removal, polisher for production of ultra-pure water for power generation, pharmaceutical and electronics industries.	Process is based on the use of alternating sheets of anion and cation permeable membranes arranged in parallel between positive and negative electrodes. As water flows down through the compartments, electrical potential drives ions from the product streams to the reject streams. Ion-exchange resin is used to facilitate ion transport.	• Chemical free regeneration • Lower operating costs due to savings in regenerant chemicals • Modular system provides redundancy as each stack can be operated and controlled individually. • Smaller footprint • Continuous process	• Requires pre-treatment (RO) • Does not remove organics.
Electrodialysis and electrodialysis reversal (ED and EDR)	Brackish water desalination, seawater desalination, softening	Electrodialysis is an electro-chemical separation process in which ions are transferred through ion-exchange membranes by means of a direct current (DC) voltage and removed from the process stream.	• Minimal pre-treatment required • Operates at a low pressure • Anti-scalant is not required • Longer membrane life expectancy because foulants are removed continuously during the reversal process • Requires less maintenance	• Limited to 50% salt rejection for a single membrane stack (stage) • Requires larger footprint to produce similar quantity and quality of water if multiple staging is used • Not as effective at removing micro-organisms and many organic contaminants

For the purpose of this study the work was limited to the application of ultrafiltration (UF) and reverse osmosis (RO) for the treatment of raw and cooling water at Komati Power Station. The limitation, however, in terms of these processes, is that membrane performance in terms of flux and efficiency is highly dependent not only on the quality of the source water that is treated but also on the membrane pre-treatment and operation.

According to several researchers (Zularisam *et al*., 2006; Amy, 2008; Fan *et al*., 2001; Kimura *et al*., 2004; Liu *et al*., 2004; Norberg *et al.,* 2007), membrane fouling is the main limitation in membrane filtration efficiency, as it causes a reduction in throughput over time. The reduction in throughput can be defined as a decrease in flux over time of operation due to the increment of hydraulic resistance as a result of scaling and/or a fouling layer on the membrane surface. A decline in throughput may also lead to increased energy requirement to maintain the same system performance. It is therefore important, prior to selection of technologies, to identify the foulants, the nature of these foulants and their impact on the various technologies. This will enable selection of the most effective technology for application as well as the impact on plant performance, before any corrective action on a production plant can be undertaken.

Fouling is influenced by various chemical and physical characteristics of the foulants. The foulant can be characterised according to its molecular structure, surface charge, molecular size and functional group. Water sources (such as untreated surface water, cooling water blowdowns, recovered effluent water, etc.) contain natural organic matter (NOM), microbiological species as well as scale-forming compounds such as calcium sulphate, barium sulphate and calcium carbonate (Gericke, 2011). Concentrations of these foulants or contaminants will vary as they are mainly source dependent. One of the most important identified foulants found in surface water filtration is NOM (Lee *et al*, 2008). It is because of the effect NOM has on the fouling of UF and RO membranes that the NOM component of the composition of the source water is of particular significance to this study.

1.2 BACKGROUND AND OPERATION PHILOSOPHY OF THE KOMATI POWER STATION DESALINATION PLANT

The coal fired Komati Power Station was not designed as a zero liquid effluent discharge (ZLED) plant. However, on return-to-service, it was required to comply with Eskom's ZLED Policy (Retief & Murphy, 2006). Zero liquid effluent discharge refers to a process in which no wastewater (i.e. effluent) is deliberately discharged from a power station to a surface water body or watercourse under normal operating and climatic conditions (Hoffman, 2000). Eskom is committed to ensure that this happens through various strategies such as effluent re-use, effluent minimisation and treatment of effluent. Over and above this, all facilities must be designed and managed in a manner that ensures that all water is used effectively.

Currently (as was the case in the past), demineralised water is produced in an ion exchange plant consisting of cation, anion and mixed bed ion exchange beds. The strong acid cation resins and strong base anion resins are regenerated using sulphuric acid and sodium hydroxide, respectively. The effluent generated from this process contains high concentrations of sodium sulphate. This effluent is discharged to the ash dams at Komati Power Station. The Department of Water and Environmental Affairs expressed concern over the mobilisation of salts, both from the ash dumps and from the unlined dams, contaminating the surrounding ground water (Hanekom, 2005). As a result, there was a requirement to reduce the salt loading in effluent streams significantly. In an endeavour to fulfil these requirements (for ZLED as well as reduction of salt loading), process selection for demineralised water production was very strongly in favour of membrane separation processes.

In order to comply with the ZLED Policy, Komati Power Station has decided to use cooling water blowdowns as source water for demineralised water production. Cooling water blowdown (also referred to as cooling tower blowdown) refers to the amount of cooling water that must be removed from the cooling water system in order to maintain dissolved salt concentrations (total dissolved solids – TDS) within specified limits as detailed in the Eskom Cooling Water Chemistry Standard (Hanekom, 2008).

The raw water source for Komati Power Station is either the Nooigtedacht or the Vygeboom Dam or a blend of both dams. The water supply for Komati Power Station is normally drawn from the Nooigtedacht Dam and is only blended with the Vygeboom Dam water when the Nooigtedacht Dam level water is low. The operator (Rotek Bulk Water Services), reports that in the past there had been incidents where raw sewage manholes from neighbouring sewage treatment plants overflowed and contaminated water in the dams. Consequently, the Nooigtedacht Dam which naturally contains high concentrations of total organic carbon (TOC) mainly in the form of NOM - is enriched with nutrients (from the spilled raw sewage), which contribute to increased microbial activity. As a result, Komati Power Station is faced with a challenge in terms of microbiological activity in the cooling water system. It was proposed that nitrogenous compounds in the cooling water are converted to ammonia as a result of biological activity, although this is not yet proven as current monitoring of the cooling water does not include the determination of organic nitrogen (Northcott & Reynolds, 2010). This proposed mechanism is supported by literature which states that the total nitrogen is comprised of organic nitrogen, ammonia, nitrite and nitrate. The organic fraction consists of a complex mixture of compounds including amino acids, amino sugars and proteins. The nitrogen in these compounds can be readily converted to ammonium through the action of micro-organisms in the water (Tchobanoglous, Burton & Stensel, 2004:62). Thus the organically bound and biologically derived nitrogen is converted to ammonia in the cooling water and poses a problem for the brass tubes of the turbine lubricating and seal oil coolers. It is postulated that the presence of ammonia in the cooling water has therefore led to stress corrosion cracking of the seal oil and lubricating oil cooler tubes and, consequently, oil in the coolers can leak into the cooling water (Northcott & Reynolds, 2010).

An investigation was launched into the problem of ammonia in the cooling water and it was established that the main contributing factors to the corrosion of the tubes were the chemistry control of the cooling water (with regard to a biocide/biodispersant dosing programme) and the construction material of the coolers (Northcott & Reynolds, 2010). The Eskom Cooling Water Chemistry Standard (Hanekom, 2008) stipulates that there must be zero oil in the cooling water at Eskom sites that have RO plants installed. This requirement must be adhered to; otherwise the RO membranes would be irreversibly fouled. The desalination plant pre-treatment does not make provision for the removal of oil; thus, in order to safeguard the desalination plant as well as the cooling water system (cooling towers and condensers), measures are being taken (work is in progress) to ensure that there is no oil contamination in the cooling water. This is achieved by coating the cooler tubes with an epoxy and installation of oil in water detectors on the lubricating and seal oil cooling systems in all the generating units in order to protect them from ammonia attack and oil. The purpose of the detection system is to shut down a leaking cooler in the event of a tube failure. The standby coolers will be put into service while the leaking tubes are plugged/replaced on the affected cooler(s).

In order to protect the desalination plant from oil contamination, should these measures fail, an alternative raw water supply is available in the event of oil ingress into the cooling water system. This system is also equipped with an oil detector that is installed upstream of the suction of the cooling water feed pumps to the desalination plant. The oil detector is interlocked with the feed pumps to stop the pumps when the oil concentration exceeds the limits specified for the RO membranes. This is essential, as the oil can foul the UF and RO membranes irreversibly; therefore, in such situations, the supply of feed water to the desalination plant will be raw water from either the Nooigtedacht or from the Vygeboom Dam (Dladla, 2010 and Garrine, 2010).

1.3 PROBLEM STATEMENT

Komati Power Station is faced with a challenge regarding feed water to the membrane plant. The feed water composition has two components that are undesirable when it comes to membrane treatment processes:

- Organic matter;
- Oil contamination.

The microbiological activity in the cooling water also presents a complex facet of the feed water composition. This is due to interactions between the microbes/bacteria, organic matter and the oil.

Komati Power Station electricity generating units were completely mothballed in 1990 (Galt, 2006). As a result there was no cooling water chemistry data upon starting of the return to service project which meant that feed water quality information was not available during the design phase of the plant. Therefore, these contaminants (NOM and oil) were not considered in the design of the plant and it was only prior to commissioning that the problem was highlighted.

The literature (Zularisam *et al.,* 2006, Amy G, 2008; Fan *et al.*, 2001; Kimura *et al.*, 2004; Liu *et a.l*, 2004; Norberg *et al.,* 2007) shows clearly that organic matter results in the fouling of membranes (both UF and RO, but in varying degrees). UF backwash procedures and clean-in-place solutions (CIPs) (for both UF and RO membranes) can be used to clean the membranes to regain flux and permeability. However, over long periods of operation, UF membrane fouling can no longer be reversed by hydraulic backwash procedures. As the number of UF cycles increase, so does the irreversible fraction that adsorbed on the membrane increases to cause membrane fouling. At this point, in order to recover the flux, the membrane must be chemically cleaned. For RO membranes, CIPs are required to recover the flux of the membrane after fouling. For UF, chemically enhanced backwash procedures are performed twice daily at Komati Power Station. The downside to these cleaning procedures is the generation of an effluent stream that must be dealt with on site and the repeated chemical cleaning of the membranes can shorten the membrane lifespan. This, however, is not a desirable situation at Komati Power Station. Therefore, it is necessary to establish the influence of the contaminants on the UF and RO membranes as well as to optimise the processes for the removal of organics in order to achieve the required make-up water quality for the boilers, while minimising the impact on the water treatment processes as well as on water management.

Due to the complexity of oil measurement, identification and quantification of oil contamination and its impacts on the pilot plant UF and RO membranes were not considered in this work. The feed water used for the pilot testing was therefore free of any oil contamination. This was confirmed by performing total oil and grease analyses of the feed water.

1.4 OBJECTIVES OF THE STUDY

In this study, the influence of organics in the raw (Nooigtedacht or Vygeboom Dam water) and cooling water on UF and RO membrane fouling at the Komati Power Station desalination plant was investigated.

The main objectives of the study were to:

- investigate the impact of natural organic matter in the Komati Power Station raw and cooling water on UF and RO membrane fouling;
- optimise the UF and RO membrane processes to minimise fouling of the membranes.

The specific objectives were to:

- determine the quality (chemical composition) of the source water;
- determine the feed water characteristics (TOC, DOC, UV absorbance at 254nm, SUVA and TOC:DOC);
- determine the capability of the polyethersulphone (PES) hollow fibre UF membranes to reduce the feed water TOC effectively in order to minimise organic loading on downstream processes;
- optimise the pre-treatment of raw and cooling feed water to the membrane process by evaluating the ability of various coagulants (MembraQuest 3000a, PAC, Ferric Chloride, Alum and U3000) to reduce the TOC effectively and, consequently, to select the most suitable coagulant for TOC and turbidity removal for in-line coagulant dosing prior to UF;
- determine the performance of UF and RO membranes through pilot studies when operated under two conditions: without pre-treatment and with pre- treatment (in-line coagulation prior to UF, biocide and anti-scalant dosing of the UF permeate);
- assess two RO membranes (BW30-2540 standard brackish water RO membrane and BW30XFR-2540 extra-low fouling RO membrane) in terms of their ability to reject TOC from raw and cooling water, ability to recover flux after the cleaning

cycles and the frequency of cleaning required by each, in order to determine which membrane is the most economic option for use with high-fouling water;

- determine the effectiveness of an organic scavenger resin and granular activated carbon (GAC) for the removal of NOM in the optimisation of pre-treatment for dissolved organic carbon (DOC) removal from the cooling water.
- perform autopsies on the RO membranes to identify foulants on the membrane surface after operation of the pilot plant.

1.5 EXPECTED BENEFITS OF THE STUDY

The expected benefits associated with this work are:

- Identification of the most suitable membranes for the process that will result in cost savings over the life of the plant.
- Reduced fouling of membranes will result in savings in operating and management costs of the plant (cleaning costs).
- Better ease of cleaning of the membranes (ability to regain flux with less loss of life after cleaning) will result in longer life expectancy and, hence, potential savings in membrane replacement cost over the life of the plant.
- High TOC rejection of membranes will result in lower TOC in the demineralised water which, in turn, will result in a reduction in the formation of organic acids which can cause failure in the steam water cycle and, hence, a more reliable plant with greater availability.

1.6 LAYOUT

The layout of the dissertation is as follows:

Chapter 1: Introduction

This chapter describes the background to the study and the problem statement. It also outlines the objectives as well as benefits associated with the study.

Chapter 2: Literature Review

This chapter provides insight on matters pertaining to natural organic matter and its influences on the fouling of membranes. The nature and chemistry of NOM is briefly described, followed by a discussion of the types and mechanisms of fouling, operational influences of the fouling of membranes, cleaning of membranes as well as techniques available for the characterisation of membrane foulants.

Chapter 3: Experimental

This chapter details the procedures used for the determination of the raw, concentrated (evaporated) raw and cooling water feed quality, raw and cooling water feed characteristics, jar tests for the removal of turbidity and organics in the feed water, treatment of the feed water using UF and RO pilot plants, optimization of the pre-treatment for the removal of organics using an organic scavenger resin and GAC, membrane autopsies, as well as the analytical techniques.

Chapter 4: Results and Discussion

Results obtained from the processing of the data are presented. Information obtained from the results is discussed in terms of the set objectives, findings from literature as well as industry standards and practices.

Chapter 5: Conclusion and Recommendations

The conclusions drawn in terms of set objectives are presented and recommendations are made for the Komati Power Station desalination plant with the view to improving/optimising the processes at the desalination plant in order to obtain optimum organic removal from the raw and cooling water feed to the UF and RO membranes at a reasonable cost.

Chapter 6

This chapter lists the reference documents used

Chapter 7

The appendices are given in this chapter.

CHAPTER 2:
LITERATURE REVIEW

2.1 INTRODUCTION

The literature review provides insight into the nature and chemistry of NOM; types of membrane fouling; the mechanisms by which fouling can occur; a brief overview of processes used for the removal of NOM such as coagulation/flocculation, UF, NF and RO; operational influences on membrane fouling; chemical cleaning of RO membranes, as well as some techniques for identifying the foulants discussed.

2.2 INTRODUCTION TO NOM: THE CHALLENGE FACING WATER AND WASTEWATER TREATMENT PLANTS

Water throughout the world contains natural organic matter as a result of the interactions between the hydrological cycle and the biosphere and geosphere. According to various researchers (Thanjekwayo, 2008; Bolto *et al.*, 2004; Kim and Yu, 2005; Amy, 2008; Aoustin *et al.*, 2001; Wang and Hsieh, 2001), NOM poses a challenge in water and wastewater treatment by:

- forming harmful disinfection by-products in drinking water;
- fouling membranes and ion-exchange resins;
- encouraging the deterioration of water quality in distribution systems because of bacterial growth;
- reducing the effectiveness of adsorption processes (e.g. activated carbon) for removal of pollutants, by competing with pollutants for adsorption sites;
- producing aesthetically undesirable effects in drinking water, such as colour, bad taste and odour.

It is evident that NOM removal and/or reduction in water supplies is critical in order to minimise the harmful effects of NOM and, hence, to ensure the success of water treatment processes. Efforts to reduce or remove NOM in water supplies are, however, complicated by the complexity of NOM itself, as well as by its interaction with synthetic organics and inorganic pollutants in water and its interaction with membranes and adsorbents. Thus, an understanding of the nature and chemistry of NOM is required in order to be able to optimise water treatment processes for the removal of NOM and, subsequently, to minimise the fouling of membranes used in the treatment of water.

2.3 NATURE AND CHEMISTRY OF NOM

2.3.1 Nature of NOM

NOM in natural surface waters can exist as dissolved natural organic matter (DNOM) or as colloidal natural organic matter (CNOM). DNOM refers to those compounds that are completely soluble under all pH conditions, fully solvated by water and consists of a wide range of compounds that generally share properties of relatively low molecular weight (Thanjekwayo, 2008). A major component of DNOM, which has been the focus of many researchers, is humic substances. Humic substances consist of humic and fulvic acids. DNOM is also reported to include simpler organic compounds: the major classes being carbohydrates, carboxylic acids, amino acids and hydrocarbons, which are reported to constitute about 20% of DNOM.

CNOM is formed by components that contain high molecular weight compounds and aggregates of smaller molecules (Thanjekwayo, 2008). These colloidal materials are classified as either hydrophilic or hydrophobic colloids. Although not well-characterised, CNOM has ionisable functional groups of the same type as DNOM (primarily carboxyl), which may bind metals and other positively charged substances, thus facilitating a means of transport for these metals and charged species (Thanjekwayo, 2008).

2.3.2 NOM Fractions in Surface Waters

NOM in surface water can be found in three fractions (hydrophobic, hydrophilic and transphilic), as can be seen in Figure 1 (NOM fractionation based on DOC):

- Hydrophobic:
  - Acids (humic and fulvic acids, aromatic acids, phenols and tannins);
  - Bases (proteins and aromatic amines);
  - Neutrals (hydrocarbons, high molecular weight aldehydes, ethers and furans).
- Hydrophilic:

  - Acids (hydroxy acids and sugars);
  - Bases (amino acids, purines and pyrimidines);
  - Neutrals (polysaccharides, low molecular weight alcohols, aldehydes and ketones).
- Transphilic fractions.

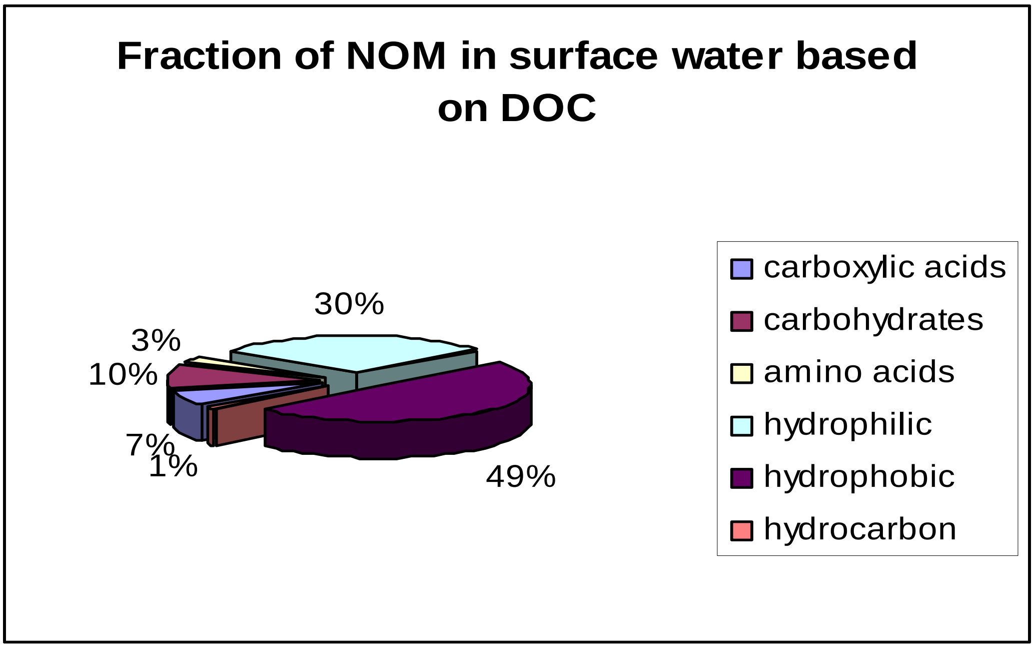


Figure 1: Fractions of NOM in surface water based on DOC (Zularisam *et al.*, 2006)

The hydrophobic fraction of NOM represents almost 50% of the DOC and comprises of larger molecular weight compounds. The hydrophilic fraction of NOM represents 25 to 40% of the DOC and comprises of lower molecular weight compounds (polysaccharides, amino acids, protein, etc.) and is normally described as the non-humic fraction. The transphilic fraction of NOM makes up 25% of the DOC in natural water with molecular weight between the hydrophobic and hydrophilic fractions (Zularisam *et al.*, 2006; Fan *et al.*, 2001).

2.3.3 Chemistry of NOM

2.3.3.1 Structure and Properties of NOM

The chemical structure of NOM has been the subject of numerous investigations over a long period because of the chemical complexity of humic substances. Some of the similar characteristics of humic and fulvic acids are as follows and they provide a general picture of the structure of NOM. Humic acid (Figure 2) and fulvic acid (Figure 3) are chosen (over hydrophilic acids and non-humic acid) as representatives of NOM because they are a major fraction of any NOM.

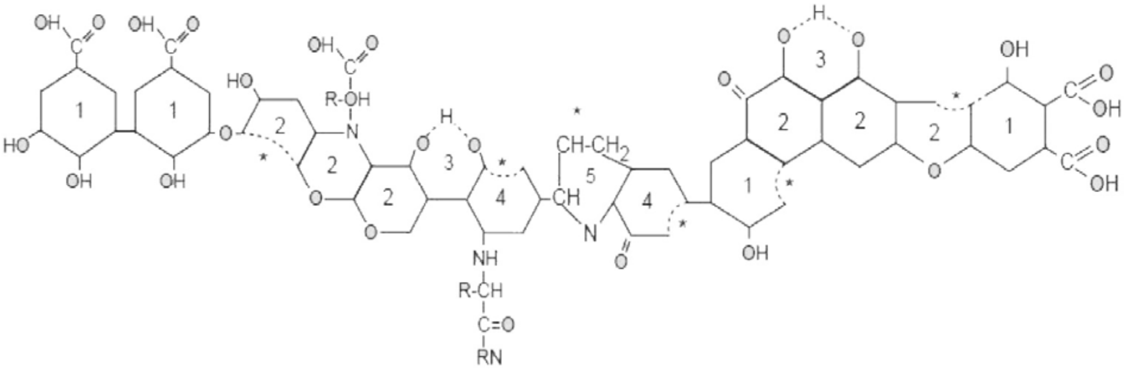


Figure 2: An example of a model structure of humic acid (Zularisam *et al*., 2006)

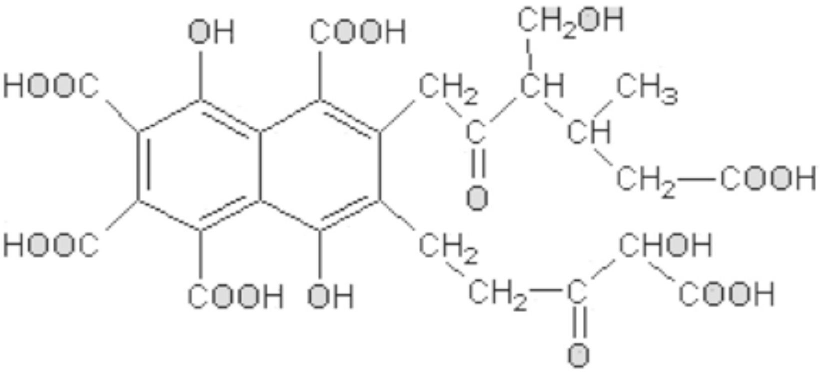


Figure 3: An example of a model structure of fulvic acid (Zularisam *et al*., 2006)

Both humic and fulvic acids (i) have high molecular weight (typically 200 to 250 000 Da); (ii) display both aliphatic and aromatic character; (iii) contain functional groups such as carboxyl and phenolic groups; and (iv) behave as polyelectrolytes in solution and, therefore, the degree of dissociation of the functional groups is dependent on pH. In most cases, humic acid is closely associated with hydrophobic acids and fulvic acid with hydrophilic acids (Wang *et al.*, 2005). Thus, humic and fulvic acids differ primarily in their hydrophobicity and solubility characteristics. In general, fulvic acid is reported to be more soluble, smaller in molecular weight, less aromatic, higher in carboxylic acid, lower in phenolic content and more highly charged than humic acids. A representation of the chemical properties of humic substances as explained by Fearing (2004) is shown in Figure 4.

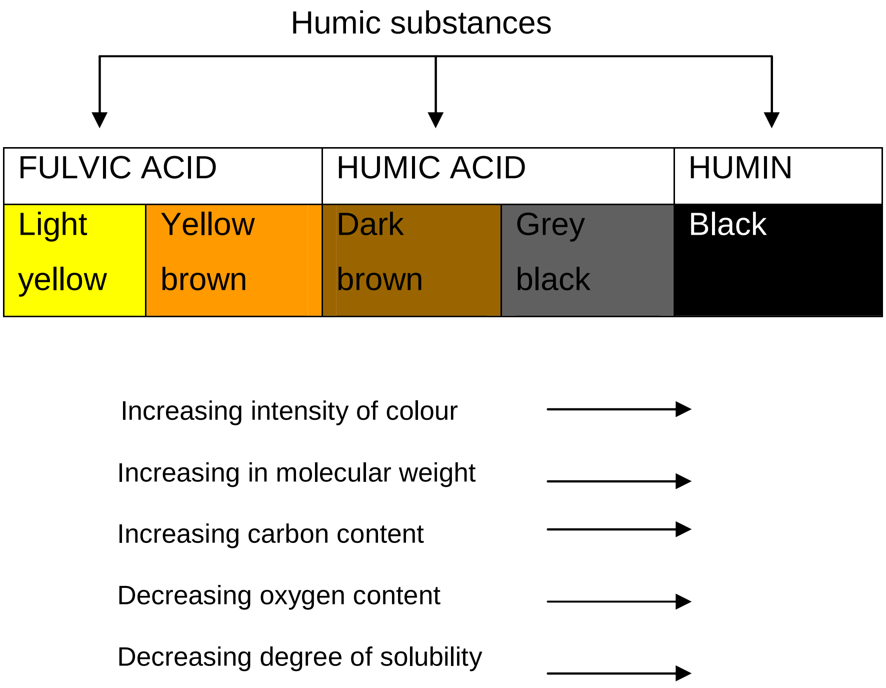


Figure 4: Chemical properties of humic substances

In his study of process options for the treatment of humic-rich waters, Fearing (2004) states that the colour of humic substances changes with increasing molecular weight and carbon content and also changes with decreasing oxygen content. This is shown in Figure 4. The degree of solubility also changes, with fulvic acid soluble at any pH, humic acid soluble at pH >2, and humin is insoluble at any pH (Gericke, 2000).

2.4 EFFECT OF SOLUTION CHEMISTRY ON NOM FOULING

Cho *et al.*, (2000) found that the extent of NOM fouling could be determined by ionic strength, pH and the presence of divalent cations. Hydraulic resistance (resistance to flow of water through the membrane) of humic acid foulants on a membrane surface was found to increase at low pH, at high ionic strength and in the presence of calcium ions. Calcium ions reduce the humic acid solubility. Other authors have also reported similar findings. For example, Schäfer *et al.*, (2000), Hilal *et al.*, (2004) and Zularisam *et al.* (2006) reported that the presence of calcium and humic substances or NOM in surface waters can cause severe UF and NF membrane fouling. It was found that at high calcium concentrations that NOM fouling was detrimental, with the mechanism depending on solution chemistry. The effect of solution chemistry on NOM fouling is shown in Figure 5.

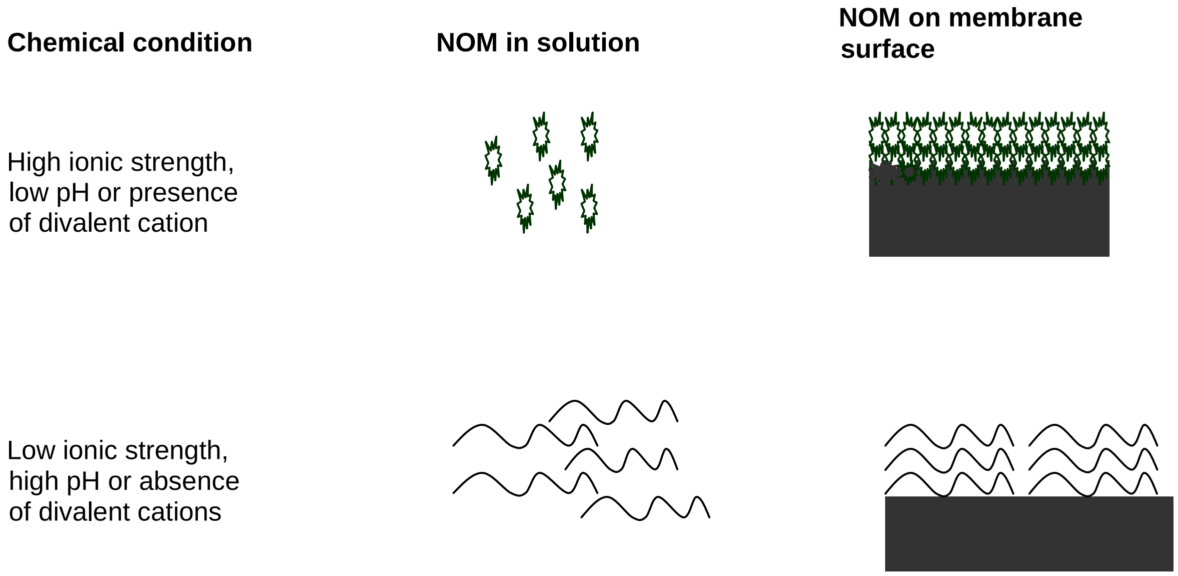


Figure 5: Effect of solution chemistry on NOM fouling (Hilal *et al.*, 2004).

The interactions between the solution chemistry and NOM in solution have implications for NOM fouling as shown in Figure 5. Calcium-humate complexes caused the highest flux decline due to their highly compactable floc-like structure, compared to calcite precipitates. NOM particles are seen to agglomerate more at low pH and in the presence of highly multivalent cation concentrations. On the contrary, the particles stretch to more linear chains in low concentration, at neutral pH and at low ionic strengths (Figure 5). The researchers (Schäfer *et al.*, 2000; Hilal *et al.*, 2004; Zularisam *et al.*, 2006) state that the deposition of calcium and organics increased with increasing pH, due to the precipitation of calcite and adsorption of organics on the calcite surface.

Incremental increase in surface water pH was found to affect the behaviour of NOM in solution. An increase in pH attributed to the ionisation of the carboxyl groups increased intramolecular repulsion and solubility and led to an increase of UV_{254} absorbance. A decrease of UV_{254} absorbance was observed at low pH conditions and this was attributed to NOM carboxyl groups that were protonated and formed large complexes and, consequently, were less soluble and precipitated resulting in the removal of most of the humic acid fraction and left only soluble fulvic acid in the solution (Zularisam *et al.*, 2006). Lee *et al.*, *(*2006) conducted an investigation into the fouling of RO membranes by hydrophilic organic fractions and the findings were similar to those of other authors. They found that solution chemistry greatly influences the fouling of the RO membrane and that the fouling of RO membranes is accelerated at lower pH, at higher ionic strength and in the presence of calcium ions. Of these chemical factors, calcium ions had a much greater effect on fouling.

2.5 TYPES AND MECHANISMS OF MEMBRANE FOULING

2.5.1 Types of Fouling

Generally, membrane fouling includes colloidal fouling, biofouling, organic fouling and inorganic fouling (which is actually referred to as scaling).

2.5.1.1 Colloidal Fouling

The presence of colloids, including silicates such as iron and aluminium silicates, result in colloidal fouling of the membrane. Colloidal fouling is attributed to the following mechanisms (Mousa, 2007):

- Accumulation of the particles present in the feedwater at the membrane surface, forming a cake layer;
- Adsorption of the smaller particles at the inner pores of the membrane;
- Restriction of the passages or blockage of the pore mouth.

For salt rejecting membranes (NF and RO) although colloidal fouling is caused by the accumulation of colloidal particles on the membrane surface and formation of a cake layer, it has been recently demonstrated that the permeate flux decline for RO and NF membranes is not caused by the colloidal cake layer resistance, but rather due to cake-enhanced osmotic pressure (Lee *et al*., 2005). It is the back diffusion of salt ions within the colloid deposit layer that results in elevated salt concentration and, thus, an enhanced osmotic pressure near the membrane surface. The increase in the osmotic pressure leads to a substantial decrease in the net driving pressure across the membrane and, consequently, to a severe flux decline (Lee *et al*., 2005).

The potential of the water to cause colloidal fouling is given by the Silt Density Index (SDI). An SDI <5 is required for RO feedwaters in order minimise colloidal fouling of RO membranes. Pre-treatment is required to reduce the colloidal and particulate matter in the feedwater when the SDI > 5.

2.5.1.2 Biofouling

Biofouling refers to the accumulation of attached micro-organisms to the surface of membrane systems. Biofouling affects the performance of a membrane system negatively, causing pressure drops and resulting in higher pumping and cleaning requirements that lead to early membrane replacement (Dreeszen, 2003). The development and growth of these micro-organisms on membrane surfaces result in a biofilm.

Biofilms

Biofilms are a collection of micro-organisms that are surrounded by slime that is excreted by the organisms and which are attached to a surface, such as a pipe surface or a membrane surface. These biofilms exist wherever surfaces are in contact with water (Dreeszen, 2003). The steps of biofilm formation are shown in Figure 6 (from Dreeszen (2003) and Saad (1992)).

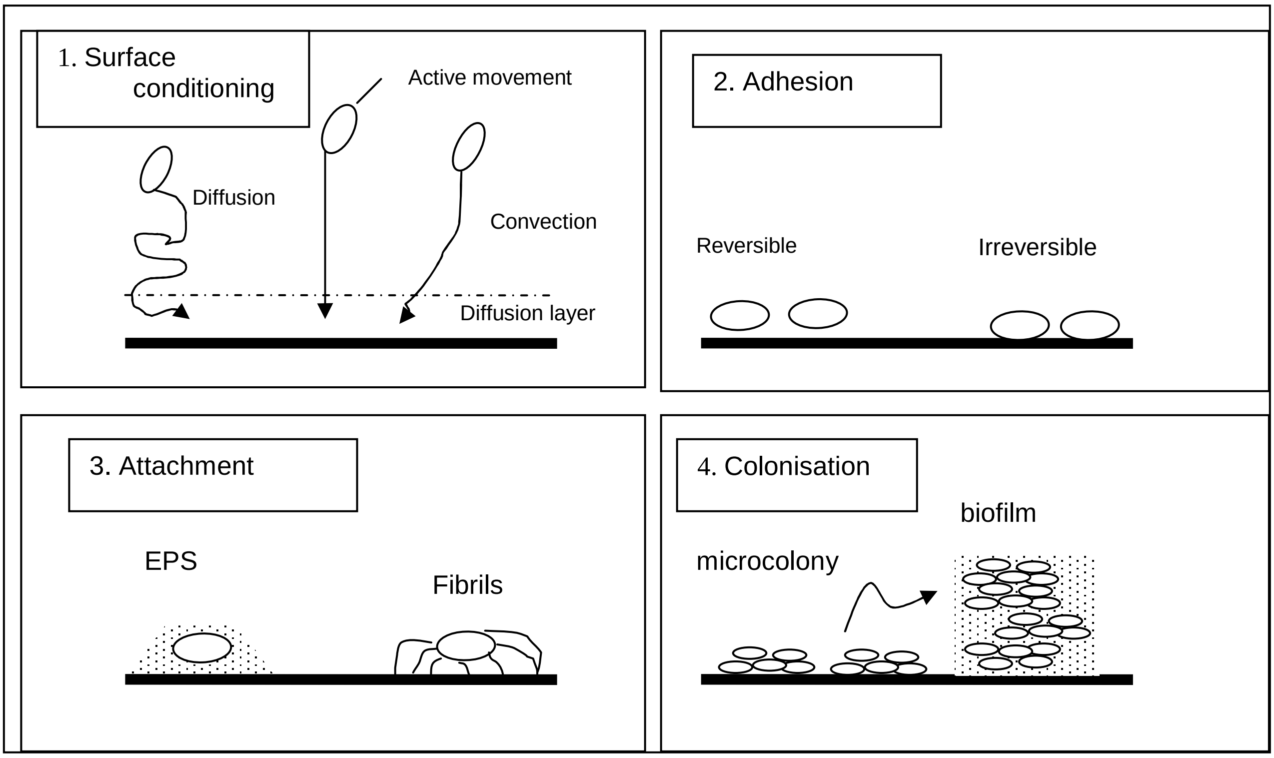


Figure 6: The steps of biofilm formation

Surface conditioning: At this stage an organic layer is deposited on the water/solid (surface) interface and forms a conditioning layer that neutralises excessive surface charge and surface free energy, which may prevent a bacteria cell from approaching near enough to initiate attachment. The adsorbed organic molecules also serve as nutrients for the bacteria.

Adhesion of bacteria: In the pipe with flowing water, some of the planktonic (free floating) bacteria will approach the pipe wall and become entrained within the boundary layer (when the flow velocity in the pipe falls to zero). Some of the bacterial cells adsorb on the surface for some time and then desorb (reversible adsorption). Some of the bacteria form structures by which they adsorb on the surface irreversibly, however at this stage the infant bioflim can be easily removed (Dreeszen, 2003; Fallon, 2004).

Glycocalyx/slime formation (attachment): The bacteria excrete extracellular polymeric substances (EPS) that hold the biofilm together and cement it to the pipe wall. EPS consist of charged and neutral polysaccharide groups that facilitate attachment; act as an ion-exchange system for trapping and concentrating trace nutrients from the overlying water as well as a protective coating for the attached cells, which mitigates the effects of biocides (Dreeszen, 2003).

Colonisation: As nutrients accumulate, the cells reproduce – thus increasing the size of the colony. When the biofilm has achieved a thickness of 10 to 25 µm, the conditions at the bottom (where the bacterial growth is in contact with the substrate) are anaerobic. Further, the attached biofilm shields sulphate reducing anaerobic bacteria (SRB) underneath its layers which leads to H_2S induced corrosion problems on metal surfaces (Saad, 1992; Fallon, 2004).

When the biofilm has reached this critical thickness, the biofilm is approaching maturity (Reynolds & Benadie, 2007). In a mature biofilm, more of the volume of the biofilm is occupied by the loosely organised glycocalyx matrix (75 to 95%) than by bacterial cells (5 to 25%) (Dreeszen, 2003). Because the glycocalyx matrix holds a lot of water, a biofilm-covered surface is gelatinous and slippery.

Similarly other researchers (El-Aleem *et al*., 1998; Kokare *et al*., 2009; Winters and Isquith, 1979) have described the formation of a biofilm on RO membranes as a stepwise process involving formation of a conditioning layer as a result of adsorption of organics on the wetted surface of the membrane, transport of the microbial cells to the conditioned surface followed by bacterial adhesion, bacterial growth (colonization) and biofilm expansion.

According to Dreeszen (2003), a mature biofilm may take from several hours to several weeks to develop. *Pseudomonas* cells were found to have adhered to stainless steel and even electroplated surfaces within 30 seconds of exposure. Flemming (1997) reported that biological activity can be found even in high-purity waters with a conductivity resistance of 18 MOhm and their mode of survival in these nutrient-low waters is by adhesion to the pipe and membrane surfaces. Bereschenko *et al*. (2010) found that a biofilm had developed on the surface of an RO membrane (ESPA2, Hydranautics) after 4 days of operation, with SEM clearly showing the distinct layer of microcolonies imbedded in an EPS matrix on the membrane surface as well as the surface of the feed side of the spacer.

2.5.1.3 Organic Fouling

The classification of organic fouling overlaps with that of colloidal fouling and biofouling. This is because organic foulants can include organic colloids, while biofouling is also induced by the presence of organics and subsequent interaction with microbiological matter. Organic fouling associated with bulk organic matter is a major challenge in water treatment processes that use membranes. This is shown in Figure 7, where common foulants in RO plants are presented.

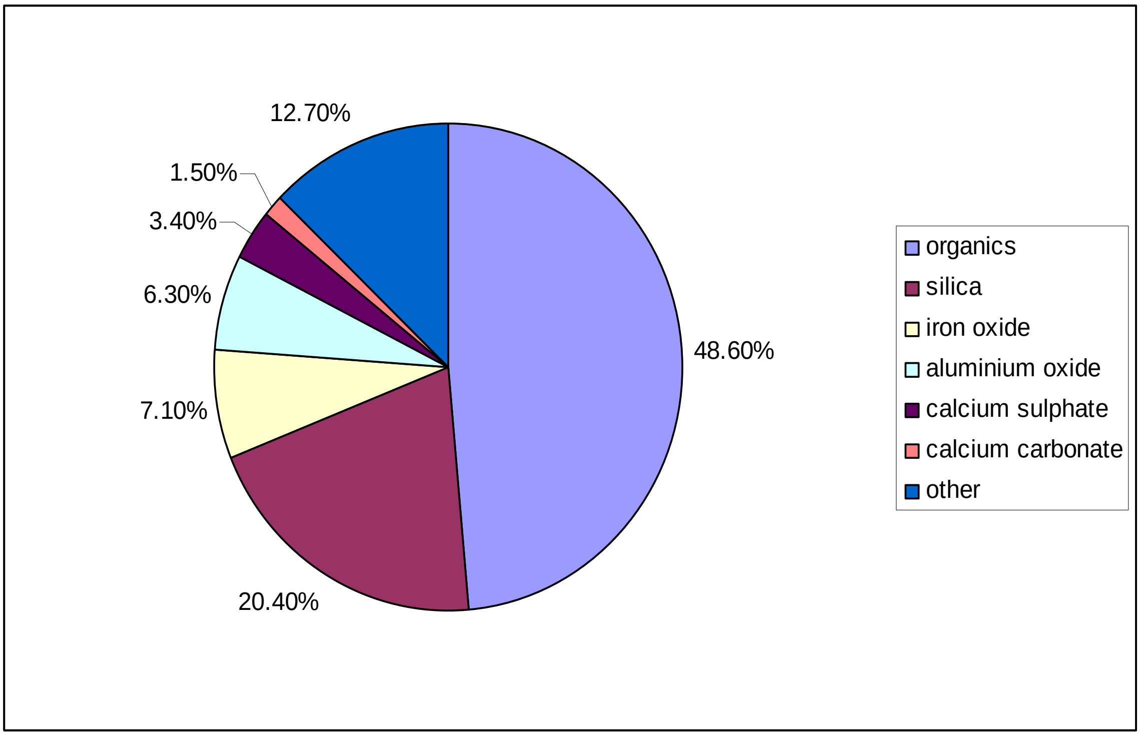


Figure 7: Common foulants/scales affecting the performance of RO membranes (Kucera, 2007)

Organic fouling is the main cause of deterioration in performance of RO membrane systems and could account for close to 50% of membrane failures (Figure 7).

Potential sources of organics include NOM, such as humic and fulvic acids, and water treatment chemicals added to feed water. Cationic polymer coagulants used in clarifiers or as filter aids can cause RO membrane fouling when overdosing occurs, due to electrostatic interaction of cationic polymers with the anionic membrane surfaces.

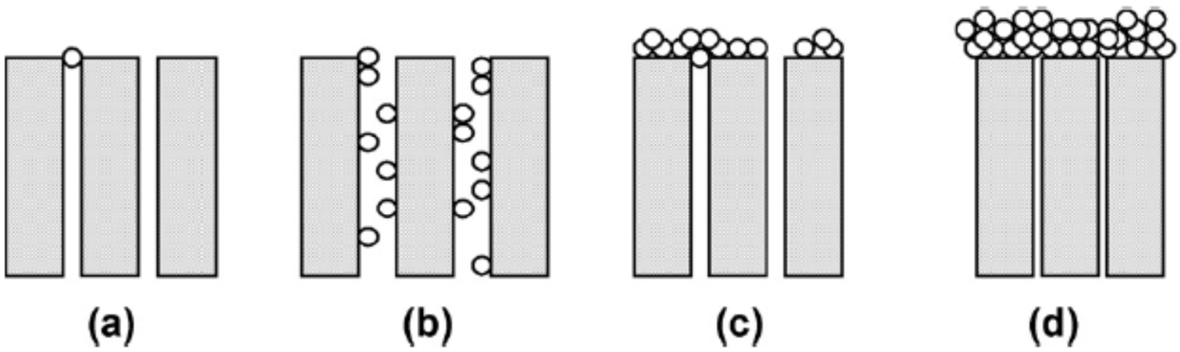
Polymeric RO anti-scalants, if overdosing occurs, can deposit on membrane surfaces and cause fouling. Acid addition, while sometimes used for scale control, has been reported to increase the propensity for NOM to cause membrane fouling (Flynn, 2009:8.19).

2.5.2 Mechanisms of Fouling

Fouling of membranes can be due a number of mechanisms such as pore blocking by solutes, cake deposition and precipitation of inorganic and organic particles at the membrane surface (Zularisam *et al.*, 2006). Mousa (2007) reported that fouling due to pore blockages is much more severe than cake layer build-up and it can be significantly reduced by increasing the particle size. The severity of the fouling appears to differ depending on membrane type. In RO membranes, cake layer build-up is generally considered to be the dominating mechanism, while in MF (MF) and UF membranes, pore blockage, and cake build-up both contribute to membrane fouling. In addition, internal pore adsorption reduced the internal pore diameter and appeared to enhance rejection of foulants in UF while in NF the fouling mechanism was mostly governed by cake deposition and concentration polarisation (Zularisam *et al.*, 2006; Mousa, 2007) .Zularisam *et al.* (2006) explained the consecutive steps of membrane blocking in flux decline during MF as follows:

- The smallest pores are blocked by all particles arriving at the membrane.
- The inner surfaces of bigger pores are covered.
- Some particles arriving at the membrane cover other particles that are already on the membrane surface, while others directly block some of the pores.
- Cake layer starts to build up.

The different fouling mechanisms are shown in Figure 8.



(a) Complete blocking (b) standard blocking (c) intermediate blocking (d) cake filtration

Figure 8: Illustration of fouling mechanisms (Mousa, 2007)

Cake layer build up (surface fouling) in RO involves the initial deposition of organics on the membrane surface and the subsequent growth of a fouling layer (Lee and Elimelech, 2006). This initial deposition of the foulant on the membrane surface is controlled by the interaction between the membrane surface and the bulk foulants while the evolution of the fouling layer is controlled by the intermolecular adhesion between the bulk foulants and the foulants that are deposited on the membrane surface. However, because monolayer coverage on the membrane surface as a result of foulant-membrane interactions is attained in a short time, for practical applications, the rate of fouling is therefore determined by foulant-foulant interactions. These are also responsible for the foulant layer structure which determines the hydraulic resistance of the fouling layer, and hence, the rate of flux decline during fouling (Li and Elimelech, 2004; Hong and Elimelech, 1997). Various researchers have found that the presence of divalent cations can enhance organic fouling of NF membranes as a result of complexation of the cations such as calcium with the NOM (humics) molecules as described in section 2.4. A study conducted by Lee and Elimech (2006) shows similar findings for organic fouling of RO membranes where atomic force microscopy (AFM) was used to quantify the intermolecular adhesion forces between the bulk foulants and the foulant deposited on the membrane surface. The results showed that solution pH, charge characteristics of the organic foulants and the presence of divalent cations enhanced organic fouling of the RO membrane (LFC1, Hydranautics) as evidenced by the flux decline.

Biofouling of RO membranes is accompanied by a decline in the permeate flux and an increase in salt passage. During a laboratory scale study conducted by Hertzberg *et al*.

(2007) it was observed that there was a sharp increase in the salt passage just prior to the onset of permeate flux decline. This indicated that there was an increase in the salt concentration (consequently an increase in the osmotic pressure) near the membrane surface as a result of deposited bacterial cells on the membrane, leading to the decline in permeate flux (Hertzberg *et al*., 2007, Hoek and Elimelech, 2003). This effect is markedly less than that resulting from the biofilm growth. Thus bacterial cells in the EPS matrix of a biofilm enhance concentration polarization near the membrane surface resulting in what is termed “biofilm enhanced osmotic pressure”. The increased concentration polarization and the associated increase in osmotic pressure near the membrane surface are responsible for the decline in permeate flux. The additional flux declined experienced in the presence of a biofilm is attributed to the hydraulic resistance of the EPS matrix.

2.6 REMOVAL OF NOM BY COAGULATION AND FLOCCULATION

Coagulation is a unit process that can be used to remove colloids and other suspended matter from water. Colloids are particles within the range of 0.001 µm to 1 µm (Crittenden *et al*, 2005:647). These particles do not settle when standing and cannot be removed by physical processes such as strainers and filters. These colloids can be either hydrophobic or hydrophilic (such as proteins, humic acids, etc.). Hydrophobic colloids have a well defined interface between the water and solid phases and have little or no affinity for water while the hydrophilic colloids have a marked affinity for water. In addition, hydrophobic particles are thermodynamically unstable and will aggregate irreversibly over time on solid surfaces. Hydrophilic colloids bind with water forming stable colloids that are readily dispersed in water. As a result the colloids cannot settle out of water.

The stability of the colloid is due to repulsive electrostatic forces and, in the case of a hydrophilic colloid, it is also due to the bound water that retards flocculation (Eckenfelder *et al.*, 2009:139). Because of the repulsive electrostatic forces, it is necessary to reduce or neutralise this charge, in order to destabilise the electric charge of particles with respect to the bulk suspending phase or media (zeta potential) so that precipitation can occur. The zeta potential can be lowered by the following means (Eckenfelder *et al.*, 2009:141):

- Change in the concentration of the potential determining ions;
- Addition of ions of opposite charge;
- Contraction/compaction of the diffuse part of the double layer by an increase in the ion concentration in solution.

Since most colloids in industrial water treatment are negatively charged, the zeta potential is lowered by adding high valence cations such as Na^+ and Al^{3+}. In this way, the colloid is destabilised and coagulation is induced. Coagulation is said to occur by two mechanisms primarily (Corbitt, 2004:5.72):

- Adsorption of the soluble hydrolysis species on the colloid, as well as destabilisation;
- Sweep coagulation where the colloid is entrapped within the precipitating hydroxide.

The reactions in the adsorption-destabilisation step occur extremely fast (within micro seconds without formation of polymers and within one second with formation of polymers). Sweep coagulation is slower and occurs between one and seven seconds (Corbitt, 2004:5.72). The concept of colloidal removal through adsorption-destabilisation and sweep coagulation in a water stream with aluminium sulphate (alum) is shown in Figure 9.

Chemical Stream **Fast (10^{-4} to 1s)** **Fast – slow (1 to 7s)**

$Al_2(SO_4)_3{\cdot}14.3H_2O \rightarrow Al(H_2O)_6^{\,3+} + SO_4^{\,2-} + H_2O \rightarrow Al(OH)_2^{\,+} + Al_7(OH)_{17}^{\,4+} \rightarrow Al(OH)_{3\,(s)}$

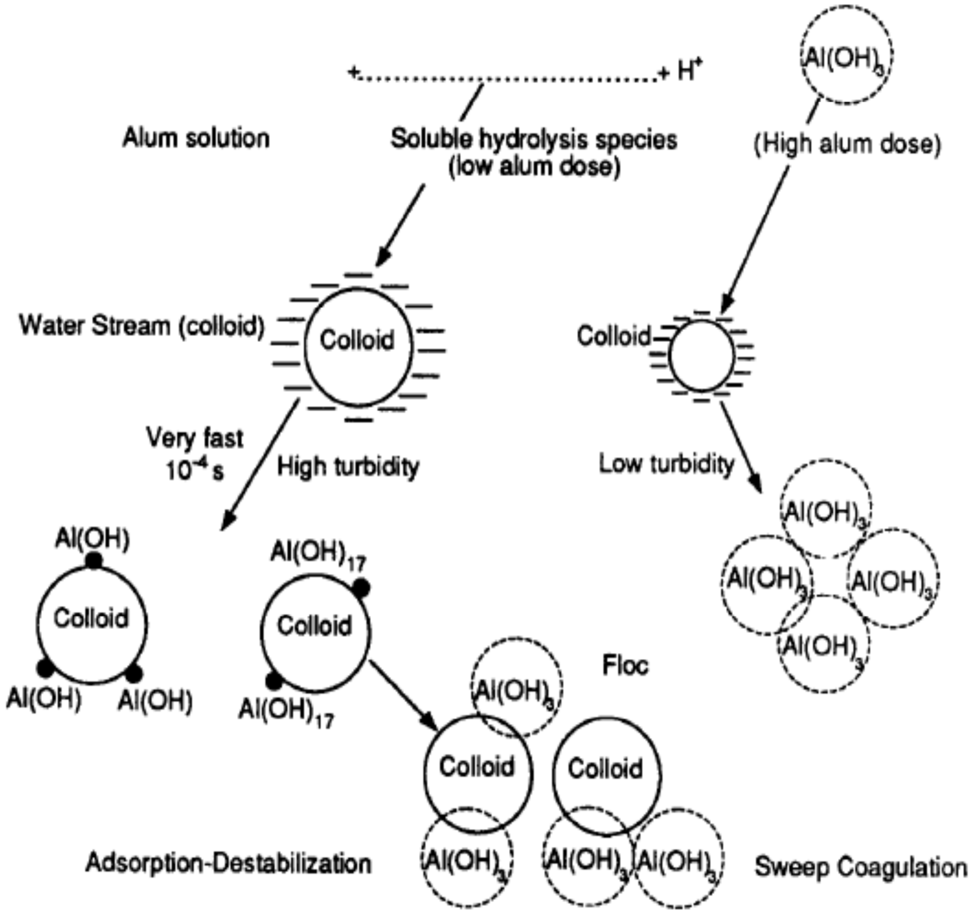


Figure 9: Illustration of adsorption-destabilisation and sweep coagulation in a water stream with alum as the coagulant (Corbitt, 2004)

A soluble hydrolysis species ($Al(OH)_2^{+}$) is formed in the stream (Figure 9): the positively charged species is attracted to the negatively charged colloid, adsorbs on the colloid and destabilises the charge (fast reaction). In sweep coagulation, it can be seen that a hydroxide is formed when the reaction reaches equilibrium. This aluminium hydroxide precipitate traps the colloid within it and carries it away, out of solution.

It is common in the water treatment industry to practise enhanced coagulation to facilitate organic removal. Enhanced coagulation involves adding excess coagulant (over and above the amount required for optimum clarification) in order to reduce NOM in the water and this process also involves destabilising the colloid, accompanied by charge neutralisation as well as precipitation and co-precipitation. Particulate NOM is said to be removed by colloid destabilisation, while the DOC portion of NOM is removed by precipitation or co-precipitation. The disadvantage, however, in increasing TOC removal by increasing the coagulant dose is that inevitably there is an accompanying increase in the amount of sludge generated and associated increases in costs and operators' time (Murray & Parsons, 2004).

2.6.1 Properties of Popular Coagulants in Water Treatment

The most popular coagulant in water treatment application is aluminium sulphate, or alum [Al_2 $(SO4)_3$ .18 H_2O], which can be obtained in either solid or liquid form (Eckenfelder *et al.*, 2009). When alum is added to water in the presence of alkalinity, the products of the reaction are calcium sulphate and aluminium hydroxide (Al_2O_3 xH_2O).The most effective coagulation for aluminium-based coagulants is at a pH that provides minimum solubility for the coagulant (for example pH 6.3 for the aluminium hydroxide precipitate at 25 ℃ and pH 6.8 at 4 ℃, Taylor & W iesner, 1999). It is at this pH that the production of aluminium residuals in the water is reduced and the maximum number of flocs for NOM adsorption is achieved: and hence, better NOM removal (Pernitsky, 2003).

Alum is not a preferred coagulant for use at Eskom power plants. Research conducted by Gericke & Aspden (2009) showed that alum can negatively affect the demineralisation process, as it introduces additional sulphate residuals, which present increased loading of sulphates that must be removed in the desalination process.

Ferric salts are also commonly used as coagulants but have the disadvantage of being corrosive and more difficult to handle. An insoluble hydrous ferric hydroxide [$Fe(OH)_3$] is produced over the pH range of 3.0 to 13.0. The solubility of the ferric ion is lower than the solubility of Al (III). This means that ferric salts can be used as a coagulant over a much wider range (pH range 3.0 to 13.0) than alum without worrying about Fe residuals in the product water (Pernitsky, 2003). Fe (III) is least soluble near pH 8.8. However, ferric salts cannot be used optimally at their point of minimum solubility due to the weakly ionised $Fe(OH)_2^+$ species, which does not have much positive charge to destabilise the colloids. Therefore, pH correction is required to lower the pH (down to pH 5.5) for optimal coagulation with ferric salts. The coagulation is more effective at this lower pH due to the more positively charged species ($FeOH_2^+$) as well as the negative charge of the colloid and NOM being lower at this low pH.

Kabsch-Korbutowicz (2006) found that pre-coagulation of feed water with aluminium based coagulants improved the reduction of NOM substances in a UF process under acidic conditions, thus resulting in a decrease in membrane fouling. The lowest flux decline was achieved when alum and polyaluminium chloride (PAC) was used. The removal of NOM achieved was 62.3%. Optimal pH values for NOM removal were found to range from pH 5 – 6. This is because at acidic conditions the positively charged hydrolysis products of the coagulant neutralize the negatively charged NOM particles.

UF of raw water without coagulation treatment results in a rapid flux decline when there is no pretreatment of the water (that is no coagulation). With coagulation used as pretreatment, the flux of the coagulated water and that of the supernatant is greatly improved compared to that of the un-coagulated water (Chen *et al*., 2006). When raw water was filtered without coagulation, the membrane removed only 7.5% of the DOC, This shows that the UF alone was not effective in NOM removal. In the filtration of coagulated water (with alum as the coagulant), NOM removal by membrane was 21%. In the filtration of the supernatant, NOM removal achieved by the membrane was 28.5% of the NOM. This result suggested that filtration of a supernatant might remove more NOM than filtration of a coagulated feed water.

Although the literature (Kabsch-Korbutowicz, 2006; Choi and Dempsey, 2004; Chen *et al.*, 2006) shows that aluminium sulphate and ferric chloride coagulants achieve effective NOM removal, Gabelich *et al.* (2002) established that residuals of alum and ferric chloride can have an adverse effect on polyamide RO membranes. The researchers found that there was a decline in flux of the RO membranes tested during the study which was attributed to aluminium hydroxides and/or aluminium silicate fouling (indicated by scanning electron microscopy (SEM) and Energy-Dispersive X-Ray Spectroscopy (EDX) data) when alum was used as the coagulant. Using ferric chloride as the coagulant resulted in a foulant comprising of aluminium, iron and silica on the membrane surface as detected by SEM and EDX. This suggests that monitoring of coagulant residuals should form part of an effective plant monitoring programme of any RO plant and that selection and control of dose rates of coagulants is very important in ensuring proper process control and minimizing the risk to the membranes.

2.7 REMOVAL OF NOM USING MEMBRANE TECHNOLOGY

Membranes present a viable technology for the removal of NOM, especially since they have become economically competitive with conventional water treatment processes. These processes, namely, NF, MF, UF and RO, differ significantly. The different processes overlap in their removal potential for various compounds (Flynn, 2009). UF works in the macromolecular to the molecular ranges (500 000 to 1 000 g/mol) while NF works in the molecular to the ionic ranges (10 000 to 100 g/mol). RO works in the ionic range. UF can remove colloids, proteins, viruses, oil emulsions and some sugars; while NF can remove proteins, soluble salts, endotoxins and some sugars; RO can remove metal and soluble salts. The separation technique involved differs for the different membrane processes.

MF is a pressure-driven membrane-separation technique used for separating particles, including bacteria, from liquid streams. The MF membrane has pores through which water (or the carrier fluid) and some solids pass. Separation is based on size exclusion, where particles larger than the pore size are retained by the membrane and particles smaller than the pore size pass through (Flynn, 2009:8.1).

UF is also a pressure-driven membrane-separation technique used for separating particles such as colloidal silica and higher molecular weight soluble species from solution. UF relies on size exclusion where species larger than the pore size of the molecular weight cut-off (MWCO) of the membrane are retained by the membrane and smaller species pass through. The MWCO of a membrane is a measure of the pore size of the membrane which can range from 0.1 to 0.001 µm for UF (Flynn, 2009:8.9).

RO and NF are also pressure-driven membrane-separation techniques used to separate dissolved solids from solution. RO relies on solution/diffusion permeation to achieve separation and NF relies on a combination of size exclusion and solution/diffusion permeation to achieve separation. The solution/diffusion model of transport through a RO or NF membrane involves three important steps (Flynn, 2009:8.9):

- Sorption of solute into the membrane;

- Diffusion of solute through the membrane;
- Desorption of solute out of the membrane.

The ability of a solute to dissolve and diffuse through a membrane polymer is a function of several variables, including temperature, pressure, concentration of solute and the membrane polymer selected (Flynn, 2009:8.9).

2.7.1 RO and NF

RO and NF are increasingly being used for NOM removal and control due to their high level of NOM rejection. High levels of TOC rejection (95%) have been reported for RO membranes (Shan *et al.*, 2009). The removal efficiency of dissolved material in a groundwater and its fouling potential were investigated by conducting pilot plant studies. The groundwater was pre-treated with UF (Gwoon *et al.*, quoted by Thanjekwayo, 2008). The TOC removal was found to be about 80% for both NF and RO membranes.

A pilot study conducted on municipal waste water (6.2 – 9.5 mg /L TOC) in Changi, Singapore using Dow's membranes showed that TOC rejection of 99.4 – 99.5% could be consistently achieved by the membranes (Majamaa *et al*., 2009).

In another study, two membranes, namely, TMG10 – an ultra low pressure RO membrane (Toray America) and NF-90 (a DOW/Filmtec NF membrane), used for pilot testing at the West Basin Waste Water Recycling Plant in El Segundo, achieved >98% TOC rejection of secondary and tertiary treated effluent water (Drewes *et al*., 2005).

Analysis of spent solutions used in the chemical cleaning of NF and RO membranes (acid and alkaline cleaning) by SEM, to identify major surface foulants showed that there were foulants attached to the membrane surface (including calcium, silica, and iron) (Gwoon *et al.*, Thenjakwayo, 2008). The SEM analysis indicated that organic matter was forming the first fouling layer close to the membrane and that the inorganic matter was layered on top of the organic fouling layer in a tetragonal shape.

Due to the fouling of membranes by various foulants more efforts are now going into manufacturing fouling-resistant RO membranes and high-organic-rejection NF membranes. A study conducted by Norberg *et al.* (2007) to evaluate 20 RO and NF membranes (including Filmtec, Trisep, Hydranautics and Osmonics) for organic rejection efficiency showed that rejection of the organics ranged from 79.7% (lowest) to 98.9% (highest rejection obtained with a NF membrane). The highest TOC rejection obtained with a RO membrane (SaeHan BL-FR membrane) was 97.5%. Four fouling-resistant membranes (Filmtec BW30FR, Trisep X20, Hydranautics LFC1 and Osmonics SG) were tested in a pilot study and the tests proved that the fouling resistant membranes could be operated over a longer period with increasing specific flux than the standard membranes and this supported claims made by the manufacturers that fouling-resistant membranes are more suited to high-organic-containing waters.

The surface analysis of these membranes showed that each membrane exhibited one or two unique surface characteristics that are favourable for minimising membrane fouling. These characteristics were found to be neutral or highly negatively surface charge, less hydrophobic and smooth surfaces. This finding suggested that the selection of commercial fouling-resistant membranes should be done carefully by considering the foulants' characteristics as well as membrane surface properties (Norberg *et al.*, 2007).

2.7.2 Ultrafiltration

The use of a low-pressure membrane process has received considerable attention lately in the water utility industry to try to meet the stringent drinking water standards specifically for removing NOM in general and for minimising disinfection by-products in particular.

Schäfer *et al.* (2000) found that organic rejection was very poor for MF and for large pore size UF membranes treating surface water. Organics rejection improved with decreasing UF membrane pore size and is higher for NF (85% DOC removal and 95% UV_{254} rejection). The UF membrane pore size of about 6 nm demonstrated a 10% decrease in the rejection of organics when compared to that of the NF membrane.

Maartens *et al.* (1999) used capillary UF membranes, with a molecular weight cut-off of 40 kDa to remove up to 98% of NOM from natural brown water. The capillary UF membranes were pre-coated with hydrophilic agents (Triton X100 and Pluronic F108) prior to filtration. Although high levels of organics removal were achieved this was accompanied by a decline in permeate flow during the filtration of the water, indicating that the membrane was fouled.

It was reported in another study that 85% removal of NOM could be obtained from Mon Villa water sources in the Western Cape of South Africa. The rejection rate appeared to be slightly lower compared to the Suurbraak water sources, but still relatively high, which is typical of UF (Thanjekwayo, 2008).

Thanjekwayo (2008) stated that the control of colloidal and biological fouling during membrane filtration is a challenge due to the enormous range and properties of the foulants that may be present in any given aqueous system, and that the use of UF for the removal of NOM is dependent on the molecular weight cut-off and the nature of the NOM.

2.7.3 Challenges Associated with Use of Membrane Technology for NOM Removal

One of the factors that limits the use of membrane technologies in water treatment plants is membrane fouling by NOM. Various factors contribute to fouling, including membrane properties, solute concentration and the hydrodynamic conditions of the system. Wang *et al.* (2004) has evaluated a range of NF, UF and MF membrane modules in order to determine the influence of NOM on the fouling of the membranes. Results of fouling experiments with humic acids demonstrated that hydrodynamic conditions and the chemical composition of the feed solutions could significantly influence the tendency of fouling. Divalent cations were found to be more readily adsorbed on NF membrane surfaces than divalent anions and monovalent anions, because the divalent cations, such as Ca^{2+} and Mg^{2+}, can act as a bridge between the membrane surface and the negatively charged humic acid molecules, as well as between the negatively charged carboxyl groups of the humic acid that are not in contact with the membrane. This results in a highly compacted fouling layer and severe flux decline. Wang *et al.* (2004) also found that the extent of fouling of humic substances for NF membranes could be reduced by increasing the crossflow velocity because a higher crossflow velocity helps to decrease the accumulation of Ca^{2+} at the membrane surface, hence, decreasing membrane fouling. Yuan and Zidney (1999) also found during MF of a 2 mg/L humic solution, that complexation of the humic acid with calcium was responsible for the flux decline.

It is apparent that the effect of NOM on membrane fouling is a complex matter involving many variables and interactions that are not clearly understood and still requires further investigation.

2.8 OTHER TECHNOLOGIES FOR THE REMOVAL OF NOM

Adsorption on activated carbon and ion-exchange are other processes that have been used for the removal of NOM from water and are popular materials for organics removal in water treatment.

2.8.1 Activated Carbon

Kim *et al*. (2008) reported that the adsorbent used in most studies of hybrid adsorption/membrane processes has been powdered activated carbon (PAC), which has typically been reported to adsorb a significant fraction of the organic carbon in feed waters and to reduce membrane fouling. This is contradicted by findings by Lin *et al.* (1999) which indicated that when PAC was used to reduce UF membrane fouling, that PAC worsened the fouling. The PAC could not remove apparent molecular weight fractions (AMW) that were >17 kDa and those that were <0.3 kDa where the largest AMW fractions of 6.5 to 22.6 kDa of humic acid (Aldrich) were responsible for UF membrane flux decline. In addition, PAC-treated humic acid solutions fouled UF membranes more than the untreated solutions did (PAC treated humic acids exhibited more flux decline). Although PAC prepared from different source materials (e.g. coconut, coal or wood) adsorbed different amounts of NOM, it was found that all the PACs investigated adsorbed non-foulant molecules preferentially over foulant molecules (Lee *et al.* (2006), quoted by Kim *et al.* (2008). According to Kim *et al.* (2008), it is the addition of the PAC to the feed water which results in greater fouling effects on the membrane as the PAC can deposit on the membrane surface and form a foulant layer that can have a negative impact on the membrane performance, even though a fraction of the DOC has been successfully removed by the PAC upstream of the membrane.

Another variation of adsorption using activated carbon is the biological activated carbon process.

2.8.2 Biological Activated Carbon

Bancroft *et al.* (1983) reported that the occurrence of large numbers of bacteria on granular activated carbon filters is well documented and, consequently, bacterial TOC removal has been studied as a mechanism by which more adsorption sites on GAC could be made available for the removal of non-biodegradable compounds: thus extending the bed life of GAC columns and reducing operating costs.

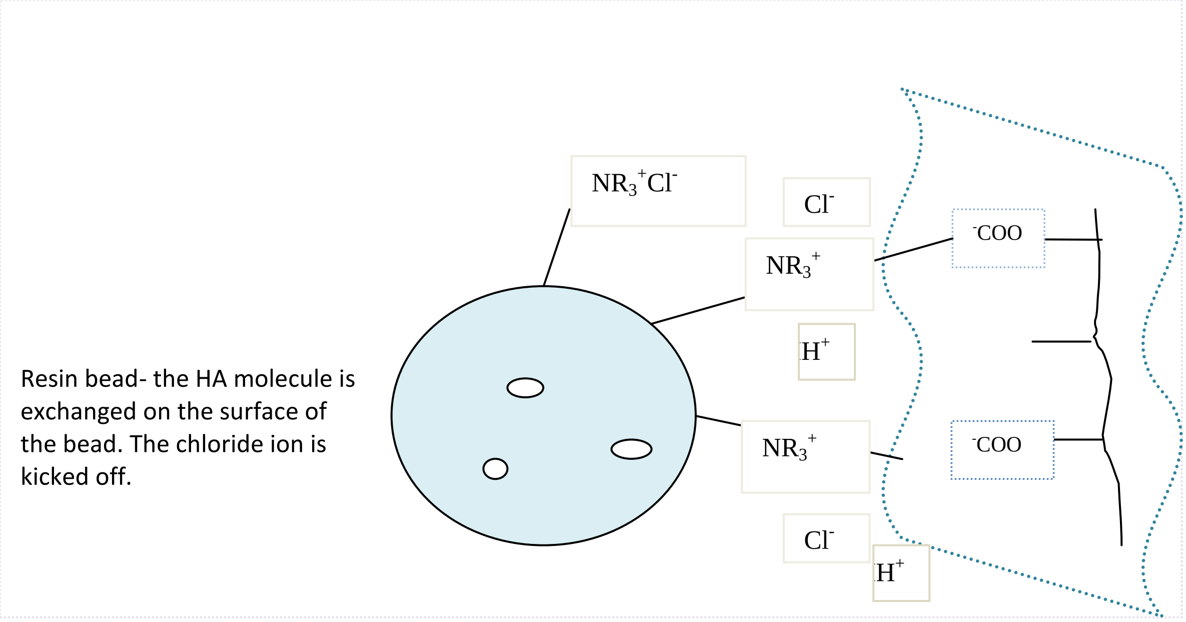
Bancroft *et al.* (1983) further state that it has been proposed that in the biological activated carbon process, carbon bed life can be prolonged by converting a portion of the recalcitrant organics to biodegradable organics by pre-ozonation. The micro-organisms attached to the GAC and then convert the biodegradable portion to biomass, carbon dioxide, and waste products, before that material can occupy adsorption sites on the GAC. It has also been hypothesised that increased biodegradation in GAC columns occurs due to the larger surface area provided by the carbon particles for attachment and growth of bacteria. According to existing data covered in the investigation by Bancroft *et al.* (1983), the data do not conclusively show that extended service time of GAC beds is a result of enhanced biodegradation. Slow adsorption kinetics could also explain long-term TOC removal. Thus, it could not be proven conclusively whether GAC provides a more favourable environment for bacterial growth and attachment than other media.

2.8.3 Ion-Exchange

Bolto *et al.* (2004) established that of a total of 160 water treatment plants designed for NOM removal in Norway, only 12 use ion-exchange. The main concerns at these plants are pre-treatment requirements, loss of capacity with time as a result of fouling and more frequent regeneration than expected. They reported that ion-exchange was claimed to be more economically effective than the use of activated carbon, carbonaceous resins or metal oxides for NOM adsorption in cases where onsite regeneration was performed.

2.8.3.1 Fundamentals of Working Principles of Organic Scavengers

The anion resin removes organics by exchanging the anionic portion of an organic molecule for an inorganic ion on the resin. Naturally occurring aromatic multivalent organics such as humics are only partially soluble and not always fully ionised and, therefore, are not as easily exchanged as highly soluble and fully ionisable compounds (Meyers, 1995). With naturally occurring organics such as humics, the carboxylic group is attracted to the cationic sites of the anion resin. The exchange occurs at the surface of the resin bead where a Cl or other inorganic anion is released and the organic functional group takes its place (this concept is illustrated in Figure 10). The organic molecule then slowly exchanges further into the inside of the bead, thus making the exchange site available for further organic removal (Meyers, 1995; DeSilva, 1997).



Resin bead- the HA molecule is exchanged on the surface of the bead. The chloride ion is kicked off.

Figure 10 Schematic of Humic Acid (HA) uptake on an ion-exchange resin (source: derived from Martin, 2009)

During regeneration, the organic inside the bead must first be exchanged with an inorganic anion and diffuse out onto the surface, where it can escape the bead and go into the regenerant solution.

Other variations of ion-exchange processes, for example, Magnetic Ion-exchange (MIEX) resins for the removal of organics in water are also available.

2.8.4 Magnetic Ion-exchange Resin

Magnetic Ion-exchange (MIEX) resin has been specifically designed for the removal of DOC from drinking water supplies and thus has a high selectivity for DOC (Bourke *et al.*, 2005; Sallanko *et al.*, 2005). The negatively charged DOC is removed from water by exchanging with a chloride ion on active sites on the resin surface. The MIEX resin is a micro-size, macroporous, strong base, magnetic ion-exchange resin. The resin has a very small particle size with a mean particle diameter of only 180 µm. While the specific surface area is comparable to other conventional macroporous resins, MIEX resin has more external bead surface area. This is said to benefit the DOC exchange kinetics (less controlled by particle diffusion) and the resistance to fouling (less DOC exchanged into the particles due to shorter diffusion paths within the smaller beads). The resin incorporates a magnetic component in its polymeric structure, which makes individual resin beads behave like small magnets that are capable of forming large, heavy agglomerates that facilitate recovery and re-use of the resin through several regeneration cycles. Another property of the resin that is significant to water treatment is the high degree of physical stability that allows pumping of the recycled resin with minimal damage to the resin.

MIEX resin preferentially removes the low to medium molecular weight fraction that is not removed by inorganic coagulants, even at very high coagulant doses, and this allows the resin to be used effectively with coagulants, since the coagulants can then easily remove the high molecular weight fraction of TOC (Bourke *et al.*, 2005; Sallanko *et al.*, 2005). MIEX resin has been successfully used for TOC reduction in groundwater (Wanneroo, Australia). The DOC in the groundwater was reduced from 9.5 mg/L to 1.8 mg/L (81.05% removal) while alum at a dosage of 90 mg/L reduced TOC from the same ground water from 9.5 mg/L to 3.5 mg/L (63% removal) (Martin, 2009). Jar tests performed on secondary tertiary effluent from the EIS Treatment Plant showed that MIEX reduced DOC from 12.8 mg/L in the feed to 7.3 mg/L (42.97% removal) (Martin, 2009). The TOC in raw water from the John Preston Water Treatment Plant in Miami was reduced using MIEX at a dosage of 8 mL/L from 14.5 mg/L to 2.5 mg/L – 82.76% removal; MIEX at a dosage of 10 mL/L reduced TOC from 14.5 mg/L to 1.6 mg/L, a removal of 88.97% and MIEX at a dosage of 12 mL/L reduced TOC from 18 mg/L to 1.8 mg/L, which was a TOC removal of 90% (Bourke, 2001).

2.9 OPERATIONAL INFLUENCES ON THE NOM FOULING OF MEMBRANES

Flux, recovery, pre-treatment, chemical cleaning and hydraulic backwashing of the membranes are operational parameters that can be influenced by NOM fouling. Researchers have found that higher flux and/or recovery can exacerbate NOM fouling. This led to the premise that NOM fouling can be minimised by conservative plant operation in terms of flux and recovery.

2.9.1 Effect of Pre-treatment on UF Membrane Fouling

Kabsch-Korbutowicz (2006) found that pre-coagulation of feedwater with aluminium-based coagulants improved the elimination of NOM substances in an UF process under acidic conditions. Optimal pH values for NOM removal were found to range from pH 5 to 6. This is because at acidic conditions the positively charged products of the coagulant hydrolysis neutralise the negatively charged NOM particles. The most effective coagulants were found to be alum and polyaluminium chloride as they provided the lowest flux decline and the highest removal of NOM (62.3%).

Aluminium sulphate and ferric chloride coagulants are widely used, as previously indicated, and they achieve effective NOM removal. Gabelich *et al.*, (2002) however, has conducted an investigation into the effect of residuals of alum and ferric chloride on polyamide RO membrane fouling and found that there was a decline in flux of the RO membranes, which was attributed to aluminium hydroxides and/or aluminium silicate fouling on the membrane. Testing with ferric chloride resulted in a foulant comprising aluminium, iron and silica on the membrane surface. This suggests that monitoring coagulant residuals should form part of an effective plant monitoring programme of any RO plant and that selection and control of dose rates of coagulants are very important in ensuring proper process control and minimising the risk to the membranes.

Dong *et al.*, (2006) studied the effect of pH on membrane fouling using coagulation as pre-treatment for river water. The findings indicated that lowering the pH (to pH of 4) could remove and decrease the organics in the feed effectively; however, this was accompanied by significant fouling of the membrane. This was attributed to the fact that a reduction in pH resulted in a decrease in the repulsion forces between NOM and the membrane and this enhanced the adsorption of NOM onto the membrane.

Maartens *et al.* (1999) proposed that in order to prevent or minimise foulant adsorption on UF membranes, solubility effects of NOM should be taken into consideration. Solubility influences the adsorptive properties of NOM during UF. The solubility of NOM is influenced by the pH of the water (humin is insoluble in water at any pH; humic acid is not soluble under acidic conditions and precipitates at pH values below 2; fulvic acid is soluble under all pH conditions). Thus, the feedwater pH can be adjusted for operation at a pH that provides maximum solubility of NOM, resulting in minimum membrane fouling. This point was found to be at a feedwater pH of 7 (confirmed by best results in terms of permeate flux, colour rejection and foulant adsorption).

Choi and Dempsey (2004) conducted a study on in-line coagulation with low-pressure membrane filtration. The experiment was conducted using alum and aluminium chlorohydrate under different conditions, including underdosing at a slightly alkaline pH of 7.47; underdosing at a more alkaline pH of 8.73; underdosing at an acidic pH of 4.81; charge neutralisation conditions at a pH of 5.27; and sweep floc conditions at a pH of 6.78 as well as at pH of 7.18 (the last three conditions are considered to be conventional coagulation conditions).

The researchers found that the highest turbidity removals (between 96 and 99%) were achieved under the so-called “conventional” coagulation conditions. The removal achieved under the slightly alkaline pH of 7.47 was 88%, while an increased (more alkaline) pH resulted in a poor turbidity removal of 10% for the same dose rate. The difference is attributed to the formation of much smaller floc, which pass through the UF membrane, resulting in the higher turbidity after UF.
Alkaline underdosing at pH 7.47 achieved 34% removal of UV_{254} absorbing species, while a more alkaline dosing condition achieved poorer removal of 9%. Higher removal of UV_{254} absorbing species (88%) was achieved at acidic (pH 4.81) underdosing conditions,

while charge neutralisation and sweep floc achieved over 90% removal of UV_{254} absorbing species. The researchers reported that the findings were consistent with those from the previous work of other researchers who reported that typical NOM concentrations of NOM in raw water (1 to 10 mg/L) were removed during coagulation by adsorption onto aluminium or ferric hydroxides.

Chen *et al*., (2007) established that UF of the raw river water used in the study resulted in a rapid flux decline when there was no pre-treatment of the water (that is, no coagulation). When coagulation was used as pre-treatment, the flux of the coagulated water and that of the supernatant was greatly improved compared to that of the uncoagulated water. The flux of the coagulated water however was higher than the flux of the supernatant (Chen *et al.,* 2007 and Dong *et al.,* 2006). The researchers proposed that this result indicated that the floc deposited on the cake layer on the membrane surface played a significant role in preventing membrane fouling. Since the foulant is deposited on the cake layer rather than on the membrane itself, it could be easily removed by backwashing. A model for the effect of coagulation on membrane fouling is shown in Figure 11 (Chen *et al.,* 2007 and Dong *et al.,* 2006).

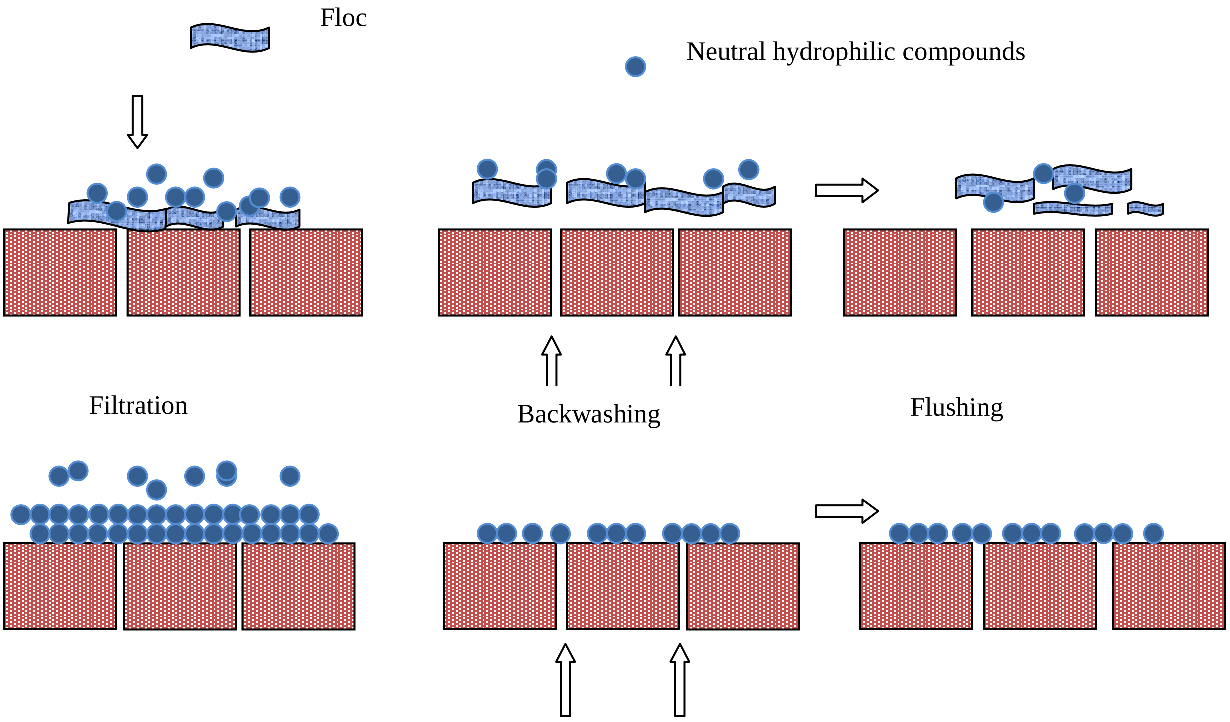


Figure 11: Prevention of membrane fouling by coagulation treatment (As proposed by Chen *et al.*, 2007; Dong *et al.*, 2006)

The model in Figure 11 shows that “as the coagulated water was filtered, the flocs deposited on the surface of the membrane and formed the cake that absorbed the neutral fraction of NOM. The cake could be easily removed by backwashing and flushing, because it was not closely adhered to the surface” (Dong *et al.*, 2006).

2.10 CHEMICAL CLEANING OF NOM FOULED MEMBRANES

Zondervan & Roffel (2007) conducted a review of the research conducted in the field of membrane chemical cleaning in relation to membrane fouling and findings of this review are presented in Table 2. The plus and minus sign in the fouling columns of the table indicate if a certain type of fouling was present (+) or absent (−). The symbols in the chemical cleaning agents columns mean: (+) tests were performed with this type of cleaning agent, (−) no tests were performed with this cleaning agent, (0) this type of cleaning agent was found to be most effective and (*) no information available.

Table 2 Membrane fouling and cleaning of fouled membranes – research conducted (source: Zondervan & Roffel, 2007).

Researcher	Application	Membrane type	Organic fouling	Inorganic fouling	Microbial fouling	Acidic cleaning	Caustic cleaning	Detergent cleaning	Sequestering cleaning	Oxidative cleaning	Enzymic cleaning	Blend cleaning
Li et al.	Gluth.Ferm. Broth	UF/PS	+	-	+	+	0	0	-	+	+	+
Madaeni and Mansourpanah	Boiler water	RO/PA	-	+	-	+	0	0	0	+	-	+
Maartens et al.	Paper effluent	UF/PES	+	-	-	-	+	+	-	+	+	0
Mohammadi et al.	Waste water	RO/PA	-	-	-	-	0	0	0	-	-	-
Lee et al.	Surface water	UF/PES	+	-	-	+	0	+	-	-	-	-
Liikanen et al.	Surface water	NF/PPA	+	+	+	+	0	+	0	+	-	+
Mohammadi	Waste water	UF/PS	+	+	-		0	0	+	+	-	-
Madaeni et al.	WPC	RO/PA	+	+	+	0	-	+	+	+	+	+
Zhu and Nystrom	BSA/LYS	UF/PES	+	-	-	+	0	+	-	+	-	-
Munoz et al.	BSA/WPC	UF/PS	+	-	-	-	0	+	-	-	0	-
Bartlett	WPC	MF/CER	+	-	-	0	-	-	-	-	-	+
Pavlova	Waste water	UF/PAN	+	+	+	-	0	-	-	0	-	-
Sungpet et al.	Textile effluent	NF/*	-	+	-	0	+	-	-	-	-	-
Mo and Huanga	Raw water	MF/PE	+	+	+	0	0	-	-	-	-	-
Veza and Rodriguez-Gonzalez	Waste water	RO/*	-	+	-	-	0	-	-	0	-	-
Chen at al.	Waste water	UF/PES	-	-	-	+	+	-	-	-	-	0
Gwoon et al.	Ground water	UF/PA	+	+	-	0	0	-	-	-	-	-
Weis	Sulphite liquor	UF/PES	+	+	-	-	0	-	-	-	-	+

Lim and Bai	Waste water	MF/PVDF	+	+	+	0	0	-	-	0	-	-

(+) present/tested; (-) not present/tested; (*) no information available

Table 2 gives an overview of what kind of foulants were found on membrane surfaces and which cleaning agents were accordingly found to be most effective in removing the foulant. It shows that for surface water and wastewater that the feed is a complex mixture of the three basic fouling categories. The review of the research also showed that organic fouling has been reported in many publications and has been found to be most effectively removed with caustic cleaning procedures (Zondervan & Roffel, 2007).

Proprietary chemicals, RoClean L211 and RoClean L403 (Avista) were used to clean organically fouled Dow Filmtec RO membranes at a plant treating river water for the production of boiler feed water. The cleaning chemicals successfully removed organic fouling and restored plant performance with notable reduction in feed pressure of the first pass from 25 bar prior to cleaning to 13 bar and improved permeate water quality (conductivity of 4.75 µS (prior to cleaning) to 3.71 µS (after cleaning) (Jeffries & Venter, 2009).

Hong & Elimelech (1997) found that EDTA (a sequestering/chelating agent) achieved a more effective cleaning for NOM (humic acid) fouled NF membranes compared to cleaning with a basic solution at pH 10. EDTA cleaning completely recovered the flux of the NOM fouled membrane. The feed water used in the study was a model water based on three types of humic acid (Suwannee River, humic acid, peat humic acid and Aldrich humic acid).

Botes *et al*. (1998) conducted a long term operation UF pilot study (from 1995 to 1997) for the treatment of irrigation water (from the Theewaterskloof/Helderberg Irrigation Scheme) for the production of potable water. The study established that the use of NaOH (pH 12)/Sodium Lauryl Sulphate (SLS) and ETDA solution in the chemical cleaning of the UF membrane was successful in removing the organic foulants and restoring permeate flow to values between 85 – 95% of the original values over the test period.

Kimura *et al.* (2004) investigated various chemical solutions to determine the most effective cleaning regime that gives the best flux restoration after cleaning UF membranes (used in the treatment of river water) that were fouled with NOM (polysaccharides). When cleaning with a single reagent (reagents tested were NaCl – 0,1 M, EDTA – 20 mM, oxalic acid, HCl – pH 2, NaOH – pH 12 and NaOCl), the most effective reagent that gave the highest flux restoration was NaOCl at 500 mg/L together with NaCl. The order of flux restoration by the reagents in increasing order was:

NaCl < EDTA < HCl < oxalic acid < NaOH < NaOCl
→

Kimura *et al.* (2004) also tested cleaning regimes with a combination of reagents and found that the most effective combination was oxalic acid at pH 2 (first cleaning agent) and NaOCl at 500 mg/L (second cleaning agent). The flux was restored up to 95% (compared with that of an unused membrane) with an oxalic acid/NaOCl combination. The order of flux restoration by the combinations tested in ascending order was:

HCl/NaOH < oxalic acid/NaOH < NaOH/NaOCl < HCl/NaOCl < oxalic acid/NaOCl
→

Zondervan & Roffel (2007) found that caustic and oxidising agents were most effective in cleaning organic foulants from UF membranes which had been used in the treatment of surface water (Twente canal water, TOC = 9.4 mg/L).

Field *et al.* (2008) conducted an investigation into the chemical cleaning of a protein-fouled UF membrane using 0,1% NaOH, NaOCl and Ultrasil 53 (a commercial cleaning agent), and found that NaOCl was found to be more effective as a cleaning agent than the other solutions tested. All the cleaning agents failed to achieve 100% flux recovery or even the normally accepted standard of 95% recovery.

The literature findings (Kimura *et al.*, 2004; Field *et al.*, 2008; Zondervan & Roffel, 2007) confirm that NaOCl (an oxidising agent) and NaOH (a base) are effective in the removal of organic foulants, with NaOH being somewhat less effective.

The case studies cited above illustrate that various cleaning solutions give varying degrees of success when employed in the cleaning of fouled NF/RO membranes. This is because various factors affect the efficiency of cleaning fouled membranes and these include chemical factors (such as salt concentration, salt type and organic foulant composition) and physical factors (such as cleaning contact time, cross flow shear rate, temperature of the cleaning solution and permeation rate during cleaning) (Lee and Elimelech, 2007; Al-Amoundi and Lovitt, 2007). Researchers (Li and Elimelech, 2004; Al-

Amoundi and Lovitt, 2007) have shown that the cleaning efficiency of EDTA is strongly dependant on solution pH – cleaning at pH 4.8 resulted in flux recovery that was 35% lower than that achieved when cleaning at pH 11; while the efficiency of cleaning with sodium dodecyl sulphate (SDS) is affected more by the concentration of the SDS solution than by the pH of the solution. The efficiency of cleaning with SDS increases with an increase in the concentration of the cleaning solution (Li and Elimelech, 2004; Al-Moundi and Lovitt, 2007). It has been found that where there is a favourable chemical reaction between the foulants and the cleaning solution, physical factors such as cross flow velocity and contact time can enhance the efficiency of the clean. This is achieved by increasing the cross flow velocity or increasing the contact time. Increased temperature also increases the cleaning efficiency while an increase in pressure (or permeation during cleaning)- although not a common practice- have a positive effect on flux recovered after salt cleaning of organic fouled RO membranes (Lee and Elimelech, 2007).

It is clear from the literature and the case studies cited, that to achieve the best cleaning results possible, it is critical that the foulant be known as foulant-cleaning solution chemistry (solution pH and concentration) plays a major role, and it is equally important that physical conditions such as contact time and temperature be increased to ensure effective chemical cleaning of the fouled membranes.

2.11 CHARACTERISATION AND MEASUREMENT OF NOM

Measurements and characterisation methods that can be applied for the characterisation of bulk samples of feed water include DOC (represents the amount of NOM in the sample), UV absorbance at 254 nm (represents aromatic character of the NOM) and specific ultraviolet absorbance (SUVA) (reveals the relative amounts of humic organic matter in the sample). SUVA is a ratio of UV at the wavelength of 254 nm and DOC. Teixeira & Rosa (2002) reported that SUVA can be used to classify the DOC nature of water. High SUVA values of between 4 and 5 L/(mg*m) correspond to DOC of the water consisting of mostly humic substances and the DOC being relatively hydrophobic, aromatic and of high molecular weight. For SUVA values of less than 3 L/(mg*m), the

DOC is composed largely of non-humic substances, and is relatively hydrophilic, less aromatic and of lower molecular weight compared to waters with high SUVA.

2.11.1 Spectroscopic Techniques

Spectroscopic techniques such as ultraviolet/visible (UV/Vis.), FTIR, nuclear magnetic resonance (NMR) and fluorescence have been used in the past for the quantitative and qualitative characterisation of NOM. FTIR and NMR analysers have been useful in estimating humic properties (Maurice *et al.*, 2002). Some of these spectroscopic techniques are presented in this section.

2.11.1.1 UV/Vis., Dissolved Organic Carbon and Specific Ultraviolet Absorbance

In general, NOM in water absorbs light in the UV (200 to 400 nm) and visible (400 to 700 nm) regions (Wang & Hsieh, 2001). According to research conducted by Korshin *et al.*, (1996/1997), inorganic chemicals that are typically present in natural fresh waters do not absorb light significantly at wavelengths greater than approximately 230 nm. As a result, the absorbance of light by natural waters is a semi-quantitative indicator of the concentration of NOM in the water. When light passes through the body of water containing humic substances the light is absorbed by the organic fractions in the water, thus resulting in a reduction of the strength of the light. The amount of this reduction, referred to as absorbance, depends on the concentration of the specific organic compound and can be used to measure the concentration of the compounds in the water using Beer's Law, in which the absorbance of a compound at a specific wavelength is proportional to the concentration of the compound (Wang & Hsieh, 2001).

Light absorbance of humic substances in water increases with the degree of aromatic rings in the humic substances, the ratio of carbon in the aromatic nuclei to carbon in the aliphatic or alicyclic side chains, the total carbon contents in the water and the molecular weights of the humic acids (Wang & Hsieh, 2001). The presence of unsaturated compounds produces a distinct colour and can therefore be detected by UV/Vis. The difference in the UV_{254} of the feed water and permeate gives the amount of humic substances rejected by the membrane.

2.11.1.2 Chromatographic Techniques

Organic carbon detection is coupled with chromatography in liquid chromatography-organic carbon detection (LC-OCD) (Batsch *et al.*, 2005). LC-OCD gives quantitative information of NOM and gives qualitative information regarding molecular size distribution. Quantification is realised by determining the carbon mass using an organic carbon detector. Qualitative information is obtained using size exclusion chromatography (SEC) (Batsch *et al.*, 2005). SEC columns divide the organic carbon into several fractions as a function of size and hydrophobic characteristics (Schäfer *et al.*, 2000). The fractions are divided into hydrophobic and hydrophilic fractions. The hydrophilic fraction consists of biopolymers (these are polysaccharides, proteins and amino sugars), humic substances, building blocks as well as low molecular weight neutrals (neutrals include mono-oligosaccharides, alcohols, aldehydes, ketones and amino sugars). In addition, UV detection and determination of the spectral absorption coefficient (SAC) at 254 nm are used to provide information regarding the aromaticity of the fractions. The hydrophobic organic carbon (HOC) is NOM consisting of lipids, but these are not further characterised in the LC-OCD technique.

A summary of techniques that are available for measuring and characterising NOM is presented in Table 3. This table is derived from various references, listed in the reference column of the table.

Table 3: Summary of techniques for the measurement and characterisation of NOM

Technique	Working Principles/Information Obtained	Advantages	Disadvantages	References
Atomic force microscopy	Describes the NOM foulant layer morphology in terms of membrane surface topography and pore distribution. It gives topographic images by scanning a sharp tip over a surface. The technique may be used in different modes such as the contact mode (the tip responds to a very short range of repulsive forces) and the non-contact mode (the tip responds to attractive van-der-Waals interactions with the sample).	Quantitative information on pore diameter, pore density and pore size distribution of membranes.		Bowen et al, 1996, quoted by Hilal *et al.*, 2004
Infrared spectrophotometry (FTIR)	Reveals foulant layer composition in terms of organic functional groups, including amines (proteins), carbohydrates (polysaccharides) and carboxylic acids (humic substances).	Identification of foulant (functional groups). Fast, repetitive scanning. Simultaneous measurement of all wavelengths.		Maurice *et al.*, 2002 Patnaik, 2004:7.11-7.12
SEM	Visually reveals foulant layer morphology using a focused beam of high-energy electrons to generate a variety of signals at the surface of solid specimens.	Reveals information about the sample, including external morphology (surface topography); the bulk microstructure (chemical composition); quantitative studies of the minerals; crystalline structure and orientation of materials making up the sample obtained by varying the incident-beam angle at a given spot.	Quantitative studies are less accurate than the electron probe micro analyzer (EPMA).	Patnaik, 2004:23.4-23.5
High-performance size exclusion chromatography (HPSEC)	In HPSEC analysis, organic molecules are separated primarily on the basis of differing molecular size so that the largest molecules are eluted first in the column, while the smallest	Determines the molecular weight distribution of aquatic humic substances.	Does not always produce reliable results at NOM concentrations <3 to 5 mg/L C. Hence, cannot be applied to waters with low DOC	Potschka 1993, quoted by Matilainen *et al.*, (2002); Maurice *et al.*, 2002

	molecules are eluted last.		concentrations without pre-concentration.	Patnaik, 2004:5.86-5.87
Fluorescence correlation spectroscopy and related spectroscopy	Organic matter fluorescence occurs when a loosely held electron in an atom or a molecule is excited to a higher energy level by the absorption of energy, for example, a photon, and fluorescence occurs when energy is lost as light as the electron returns to its original energy level (ground state). The fluorescence is characteristic for each different compound. Other available techniques include fluorescence emission spectrometry, in which emission is scanned over a range of wavelengths for a fixed excitation wavelength. This increases the range of fluorophores that might be found. Excitation emission matrix fluorescence spectroscopy (EEMS) is the state-of-the-art technique used. The principal of EEMS is that excitation, emission and fluorescence intensity can be scanned over a range of wavelengths synchronously and plotted on a single chart, developing a “map” of optical space: an excitation emission matrix.	EEMS is rapid (1 min per sample). The production of a 3D plot of fluorescence excitation wavelength, emission wavelength and intensity allows the visualisation of a range of fluorophores in a given sample, in their relative positions in optical space. Relative concentrations can be determined based upon calibration of fluorescence intensity against TOC or standards with detection limits at ppb or ppm levels	The fluorescence technique is relatively slow and inflexible, particularly if a range of excitation and emission wavelength scans is required for the study of more than one fluorophore. Output is restricted to a linear scan, in which the choice of excitation wavelength determines the molecules that may be identified.	Hudson *et al.*, 2007

		depending on the fluorophore. The technique is non-destructive and requires little or no sample preparation.		
UV/Vis.: UV_{254}	NOM absorbs light over a wide range of wavelengths; hence, light absorbance is a semi-quantitative indicator of the organics concentration in water.	UV_{254} is used in the water treatment industry for monitoring DOC once the correlation between DOC and UV absorbance has been established.	When the concentration of humic acids is low, the UV_{254} obtained may be too low and cause a relatively high random error (hence large deviations of NOM).	Korshin *et al.*, 1996/1997; Wang & Hsieh, 2001
UV/Vis.: Scanning spectrophotometry	Offers a more stable analytical result (less random error) than the widely used UV_{254} method.			Wang & Hsieh, 2001
Organic detectors	Measure NOM as DOC and TOC.	Quantification of dissolved and particulate NOM.	Long analysis time.	
SUVA	The specific UV_{254} (defined as UV_{254} per milligram of organic carbon, also known as $SUVA_{254}$) is used as another surrogate of the disinfection by-product precursors.	In drinking water treatments, $SUVA_{254}$ can be used to predict the reactivity of humic acids with coagulants and disinfectants. Test is quick and, hence, results are available quickly for troubleshooting.		Wang & Hsieh, 2001
Chromatography: LC-OCD	Quantifies NOM by determination of carbon mass using an organic carbon detector and gives qualitative information regarding molecular size distribution using SEC.	Different fractions of NOM can be quantified and the aromaticity can be determined.		Batsch *et al.*, 2005

2.12 CHARACTERISATION OF FOULED MEMBRANES

Techniques used for characterising fouled membranes are shown in Table 4. The table is derived from various references as indicated in the reference column of the table.

Table 4: Techniques for that have been used for characterising NOM fouled membranes

Technique	Information Yielded by Technique	Reference
Atomic force microscopy	Describes the NOM foulant layer morphology in terms of membrane surface topography and pore distribution.	Maurice *et al.*, 2002
Infrared spectrophotometry (FTIR)	Reveals foulant layer composition in terms of organic functional groups, including amines (proteins), carbohydrates (polysaccharides) and carboxylic acids (humic substances).	Maurice *et al.*, 2002
SEM	Reveals information about the sample, including external morphology (surface topography); the bulk microstructure (chemical composition); quantitative studies of the minerals; crystalline structure and orientation of materials making up the sample obtained by varying the incident-beam angle at a given spot.	Patnaik, 2004:23.4-23.5
EDX	Analytical technique used for elemental analysis or chemical characterisation of a sample. An energy dispersive detector is used to separate the characteristic X-rays of different elements into an energy spectrum which is analysed using software to determine elements within the spectrum.	Patnaik, 2004: 9.20
SEM/EDX	Analysis gives an ionic breakdown of the compounds on the membrane surface.	Maurice *et al.*, 2002
Electron spectroscopy for chemical analysis (ESCA) or X-ray photoelectron spectroscopy	Analysis may be used to identify organic compounds that have been bound to the membrane surface. The technique can be used to distinguish between different binding or oxidation states of the compounds. Can detect all elements except hydrogen.	Patnaik, 2004: 23.5; 9.20-9.21

2.12.1 Membrane Autopsies

The techniques involved can be broken down into two groups: destructive tests, which require the membrane casing to be opened, and non-destructive tests, which are carried out without opening the membrane. The advantage of a non-destructive autopsy is that the element may be re-used, with the disadvantage that the range of analytical techniques that can be used is much smaller (Jefferies & Champion, 2009). Destructive techniques can give much more detail on the foulants and are presented in this section.

2.12.1.2 Destructive Techniques (Jefferies & Champion, 2009)

Basic scale tests may be carried out on material that has been removed from the membrane surface using standard laboratory reagents. For example, mineral scale that fully dissolves in acid and gives off a gas (CO_2) will be calcium carbonate. Additional tests on the spent acid for specific ions, such as calcium, magnesium, sulphate and phosphate, will allow an approximate cation/anion balance to be made. Tests may also be carried out to check for the presence of iron, manganese and aluminium.

Material/foulant that has been removed from the membrane surface may also be analysed to determine elemental or chemical composition of the foulant using SEM/EDX as described in Table 4 (Section 2.12).

Foulants that are not crystalline in nature are tested using an adenosine triphosphate test. This is a two-minute test to identify if the foulant is predominantly biological or if it has an organic base such as humic or fulvic acid. A more accurate assessment of biological fouling may be made using dip slides, laid on the surface of the membrane and then incubated.

A loss-on-ignition test is a quantitative measure of the amount of moisture contained in the foulant and measures how much of the foulant is inorganic and how much is organic. First, a given area of membrane is carefully scraped into a crucible and weighed. This gives a grammes/cm^2 figure for foulant density. The foulant is then dried at 110 °C and weighed again to give the percentage of moisture that was in the sample. Finally, the foulant is incinerated at 550 °C, which destroys any organic materials in the sample (by combustion). The moisture content and the dry residue give information regarding the nature of the foulant (organic or inorganic in nature) (Majamaa *et al.*, 2009; Jefferies & Champion, 2009).

CHAPTER 3:

EXPERIMENTAL

3.1 INTRODUCTION

This chapter describes the experimental procedures and analytical techniques used in the study.

3.2 FEEDWATER QUALITY

Experimental procedure/method

Komati Power Stations' raw water supply is pumped from the Nooigtedacht and/or Vygeboom Dams. The water quality (chemical composition) of the pilot plant feed water was determined by the Eskom Komati Power Station laboratory. The TOC of the raw water feed to the power station was monitored to determine variations in the incoming raw water TOC to the stations for the period of January 2009 until October 2009.

Analytical techniques

- TOC

The TOC was determined by analysing unfiltered samples on a Sievers 900 TOC analyser.

- Chemical composition

The water quality (chemical composition) of the raw water, concentrated raw water and cooling water that was used during the pilot plants operation was determined in order to establish the quality of the feed to the membranes.

  - Alkalinity (P-Alk and M-Alk) was analysed using a Metrohm 809 Titrando autotitrator.
  - The cations and anions were analysed using an ICP PerkinElmer Optima 7000DV and a Dionex ICS-1500 IC units, respectively.

3.3 KOMATI POWER STATION DESALINATION AND PILOT PLANTS' SOURCE WATER

Experimental procedure/method

The pilot study was conducted using raw water, evaporated (concentrated) raw water and cooling water (cooling tower blowdowns). The raw water sources for Komati Power Station are the Nooigtedacht and Vygeboom Dams (see Figure 12). The Nooigtedacht dam is the upper of the two dams on the Komati River and the nearest town is Carolina while the Vygeboom Dam is situated 53 km downstream of the Nooigtedacht Dam on the Komati River and the nearest town to the dam is Badplaas. The Usutu-Komati Scheme is shown in Figure 12 (https://hyperwave.eskom.co.za/Eskom/Generation/Generation%20Disciplines/Chemistry/Chemistry%20Training_1/General%20Knowledge%20Chemistry/Misc/Water_Sources/).

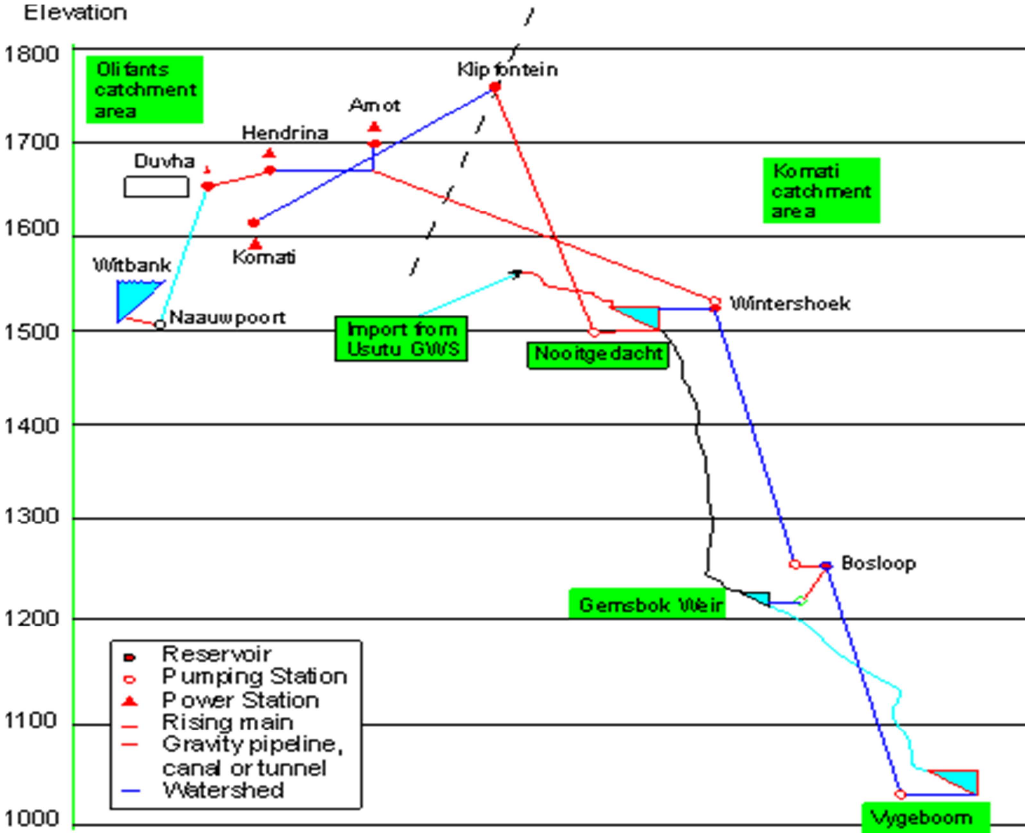


Figure 12 Usutu-Komati Scheme

Raw water from the Nooigtedach and/or Vygeboom Dams and Cooling Water (CW) from the West forebay of the Komati Power Station CW system were used as feed water to the UF pilot plant, while concentrated (evaporated) raw water and raw water were used as feed water to the RO pilot plant. The raw water was gravity fed to the UF pilot plant feed tank from a tie-in on the CW raw water make-up line, while the CW feed water to the UF pilot plant feed tank was pumped to the tank using a submersible feed pump installed in the CW West forebay.

The decision to use concentrated (evaporated) raw water for operation of the RO pilot plant was due to oil contamination of the cooling water. Raw water from the Nooigtedacht and/or Vygeboom Dams was evaporated on a hot water bath and hot plates in the laboratory to achieve “synthetic” cooling water/concentrated raw water. The temperature of the water bath was maintained between 58℃ and 61℃. As the raw water was evaporated, it became more concentrated in salts. In the cooling water circuit evaporation is achieved through the cooling towers.
The raw water was evaporated to over five times the concentration of the normal raw water TOC (6.3 mg/L on average). Under normal operating conditions, the cooling water circuit is expected to cycle up 6.8 times from the raw water concentration.

3.4 FEEDWATER CHARACTERISTICS

3.4.1 DOC Composition

The feed water to the pilot plant was characterised in terms of its DOC composition, using liquid chromatography coupled with organic carbon detection (LC-OCD).

Size Exclusion Chromatography (SEC) was used for separation of the organic fractions followed by multi-detection with organic carbon, UV-absorbance at 254 nm and organic bound nitrogen (Huber, 2008). All concentration values refer to the mass of organic (bound) carbon (OC). As a “rule of thumb”, compound mass is about twice (for acids it is threefold) the value of OC. Chromatograms are processed on the basis of area integration using the program ChromCALC. In many samples, the acid fraction contains low molecular mass humic acids which are subtracted by ChromRES on the basis of SAC/OC ratio for humics (Huber, 2008).

3.4.1.1 Parameters Determined by the LC-OCD technique

Parameters that are determined by the LC-OCD technique are explained in Table 5.

Table 5: Parameters determined by the LC-OCD technique (Huber, 2008)

Parameter	Sub-fractions	Determination of the parameter
DOC		Determined after in-line 0.45 μm filtration of the sample.
Hydrophobic Organic Carbon (OC)		DOC minus Chromatographic OC, that is, all OC retained on column.
Chromatographic DOC		OC value obtained by area integration of the total chromatogram.
	Refractory organic matter: Humics	In LC-OCD measurements, humics are defined based on retention time, peak shape and SAC.
	Refractory Organic Matter: Building blocks	These are sub-units of humics with molecular weights of 300 to 450 g/mol.
	Biogenic organic matter: Biopolymers	This fraction is very high in molecular weight (100 000 to 2 Mil. g/mol) and is hydrophilic, not UV-absorbing. Biopolymers are typically polysaccharides but may also contain protein matter (this is quantified on the basis of organic bound nitrogen).
	Low molecular weight (LMW) organic acids	In this fraction, compounds are all aliphatic, LMW organic acids which co-elute due to an ion chromatographic effect. A small amount of humic substances (HS) may fall into this fraction and is subtracted on the basis of SAC/OC ratios.
	LMW neutrals	LMW (weakly charged or uncharged) hydrophilic or slightly hydrophobic compounds appear in this fraction. This includes alcohols, aldehydes, ketones and amino acids. The hydrophobic character increases with retention time, e.g. pentanol

		appears at 120 minutes and octanol at 240 minutes.
	SUVA (SAC:DOC)	Additional parameter derived from the ratio of SAC to DOC.

3.4.2 UV Absorbance and DOC

Experimental procedure/method

The feed water (raw and concentrated raw water) samples were filtered using a 0.22 µm filter.

A calibration curve was drawn up with a KHP concentration expressed as C on the X axis and UV absorbance at 254nm on the Y-axis. A stock solution of 1 000 mg/L KHP as carbon (C) was prepared and used to prepare working solutions for the calibration curve. The calibration curve was used as reference to determine the DOC during times of TOC analyser unavailability. The detailed results of the calibration curve are presented in Appendix 1.

Analytical techniques

- DOC

The DOC of the filtered raw and concentrated raw water and the KHP solutions were measured using a Sievers 900 TOC analyser.

- UV_{254}

UV absorbance at 254 nm of the filtered raw and concentrated raw water and the KHP solutions were measured using a HACH DR5000 spectrophotometer. A 1 cm quartz cell was used in the determination of UV_{254}.

3.5 TURBIDITY AND ORGANICS REMOVAL (IN-LINE COAGULATION)

Experimental procedure/method

Jar tests were conducted using a six paddle Stuart flocculator. Jar tests were performed on 500 mL samples of raw and cooling water. Five coagulants, namely, MembraQuest 3000a, polyaluminium chloride (PAC), aluminium chlorohydrate (ACH – U3000), ferric chloride and aluminium sulphate (alum) were tested at varying concentrations (from 5 mg/L to 30 mg/L). For each coagulant tested at the specific concentration, the procedure involved a fast stir step at 257 rpm for one minute, followed by filtration on a 0.22 µm filter. There was no slow mixing and no settling done, as the intention was to imitate process conditions of the main plant which utilizes in-line coagulation with only a flocculation vessel that provides 30 seconds of retention time.

Analytical techniques

Success of the jar tests ("in-line" coagulation) was measured by removal of turbidity, DOC and UV_{254} absorbing species, after filtration of the coagulated water on a 0.22 µm filter.

- TOC

TOC of the coagulated (unfiltered feed water sample) was measured using a Sievers 900 TOC analyser.

- DOC

DOC was obtained by filtering the samples on a 0.22 µm nylon syringe filter and analysing the filtered sample on the Sievers TOC analyser.

- UV absorbance at 254 nm

UV absorbance at 254 nm was measured using a 1 cm quartz cell on a Hach DR 5000 spectrophotometer analyser.

- Turbidity

Turbidity was measured on the Hach HN2100 turbidity analyser.

- pH

pH measurement was done with a MeterLab PHM210 standard pH meter.

3.6 ULTRAFILTRATION TREATMENT

The UF pilot study was conducted in two steps:

- No coagulation (raw water and cooling water);
- Coagulation with U3000 (raw water and cooling water).

3.6.1 UF Pilot Plant

The pilot plant consisted of a single UF hollow fibre membrane module (Norit), as depicted in Figure 13. The UF membrane was made of PES. The membrane filtration area was 40 m^2 and the nominal pore size of the membrane was 0.01 µm. The membrane was operated in the dead-end mode.

A schematic of the UF pilot plant that was used in the experimental set-up is shown in Figure 13.

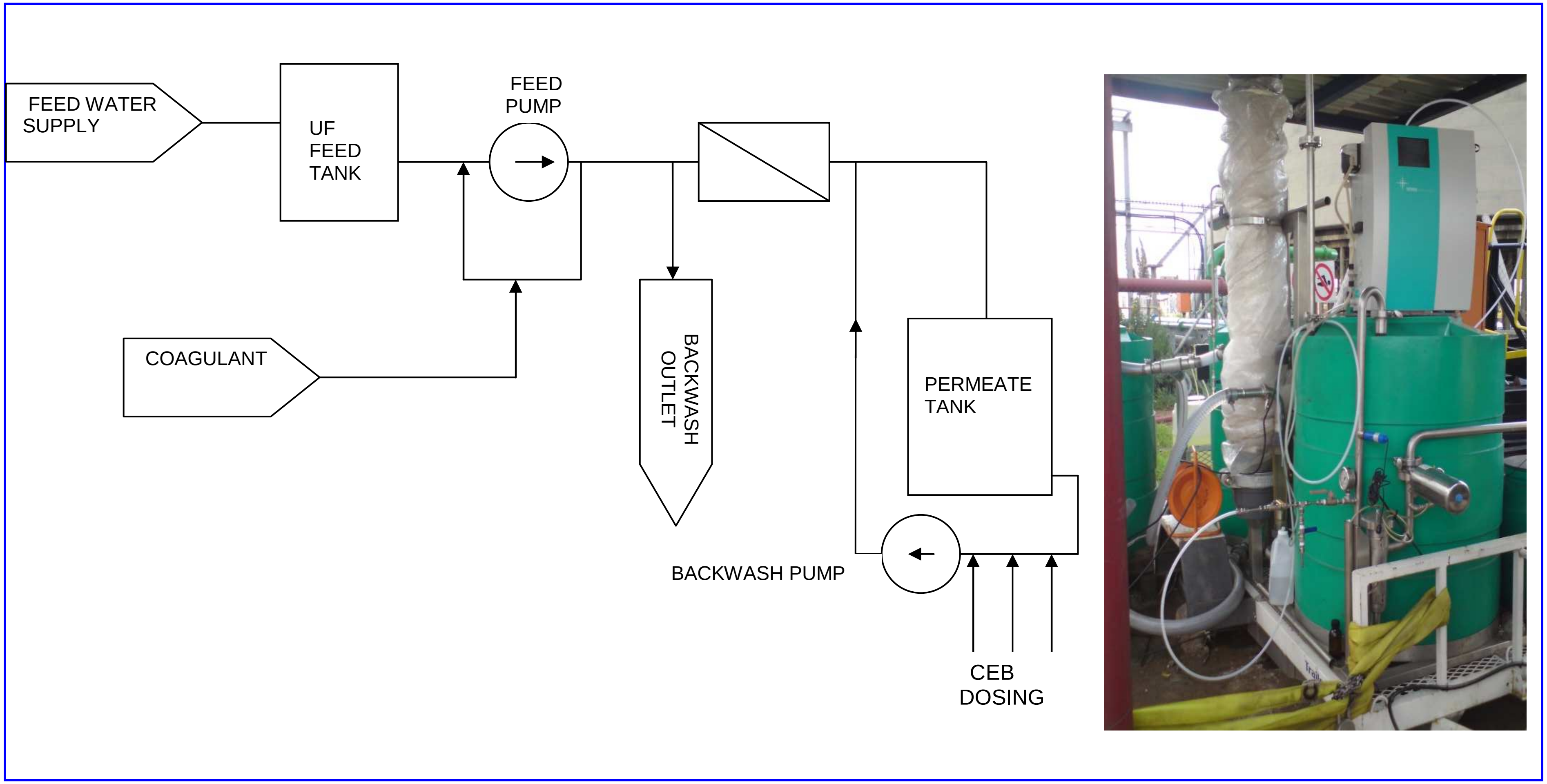


Figure 13: **Schematic and photo of the UF pilot plant**

Feed water to the pilot plant was supplied in two ways:

Raw water was gravity-fed from a tie-in on the raw water make-up line to the CW West forebay, while a submersible pump was used to pump cooling water from the forebay to the UF feed tank. The feed water flowed through a 1.5 mm mesh strainer to the feed tank to remove coarse particles from the feed water. The UF feed pump then supplied water from the feed tank to the UF membrane. A recycle line from the discharge to the suction of the feed pump was provided and coagulant was dosed into the recycle line to aid with proper mixing. From the membrane, the product was collected in a permeate tank which was fitted with an overflow back to the forebay.

The membrane backwash pump took suction from the permeate tank. On the backwash line, three dosing points were provided for acid, sodium hypochlorite and sodium hydroxide dosing. The chemical dosing was done during the chemically enhanced backwash operation, using HCl (450 mg/L), NaOCl (200 mg/L) and NaOH (525 mg/L).

The UF was operated with raw and cooling water under two different conditions (with and without coagulation). The feed pressure was adjusted using a variable speed drive on the feed pump to try and maintain the permeate flow. The feed pressure ranged from 18 kPa to 38 kPa during the period of operation of the pilot plant. The lowest temperature recorded was 19 ℃ and the maximum temperature was 28 ℃. The UF cycle was for a period of 20 minutes. During the operation with in-line coagulation, the coagulant (U3000) was dosed at a concentration of 20 mg/L during the UF cycle. The feed flow rate ranged from 1 360 to 2 000 L/h.

The selection of the coagulant of choice to be used on the pilot plant was achieved by conducting jar tests with various coagulants under the same process conditions as expected in the main plant. The decision made on which coagulant to operate with was based on the most effective TOC and turbidity reduction that could be achieved under the test conditions described in Section 3.5.

3.6.1.1 UF Backwash

After a UF cycle of 20 minutes, a hydraulic backwash was initiated on the membrane. Each backwash was done for 60 seconds at a flow of 10 m^3/h.

3.6.1.2 Chemically Enhanced Backwash

Initially, a CEB was done after every 12 hydraulic backwashes. This was later adjusted to six hydraulic backwashes. The CEB procedure is detailed in Appendix 2 - cleaning procedures (2.2 – UF CEB).

The membrane backwash pump took suction from the permeate tank. On the backwash line, three dosing points were provided for acid, sodium hypochlorite and sodium hydroxide dosing. The chemical dosing was done during the chemically enhanced backwash operation, using HCl (450 mg/L), NaOCl (200 mg/L) and NaOH (525 mg/L).

<u>Analytical techniques and parameters measured during UF, backwash and CEB procedures</u>

- Flow, m^3/h (and L/h)

Flow was measured using the rotameters installed on the permeate side of the UF membrane on the pilot plant. Normalised flux was calculated using temperature correction factors for Filmtec membranes from the DOW Technical Manual – Appendix 3.

- Pressure , kPa

Feed pressure to the UF membrane was monitored using the pressure gauge installed at the discharge of the feed pump.

- SDI_{15}

SDI_{15} was measured using an SDI assembly that consisted of a pressurised vessel, pressure regulator, 47 mm filter holder, 0.45 µm membrane filter and a stopwatch. The sample was pressurised in the pressure vessel by using control air. The pressurised sample was passed through the membrane filter contained within the filter holder at a pressure of 200 kPa. The time (initial time – t_i) taken to collect 500 ml of sample was taken and a further 500 ml sample was taken after 15 minutes and the final time (t_f) was noted. The SDI_{15} was calculated using the equation $((1- t_i/t_f)*100)/15$.

- TOC

TOC of the coagulated unfiltered feed water sample was measured using a Sievers 900 TOC analyser.

- DOC

DOC was obtained by filtering the samples on a 0.22 µm nylon syringe filter and analysing the filtered sample on the Sievers TOC analyser.

- UV absorbance at 254 nm

UV absorbance at 254 nm was measured using a 1 cm quartz cell on a Hach DR 5000 spectrophotometer analyser.

- Turbidity

Turbidity was measured on the Hach HN2100 turbidity analyser.

3.7 RO TREATMENT

Experimental procedure/method

The RO run was conducted under the following conditions:

- No pre-treatment (no pre-filtration and no biocide and anti-scalant dosing initially) using raw water and concentrated raw water ;
- Pre-treated water (UF filtered cooling water, with UF operating with in-line coagulation with U3000 as coagulant), and dosing of anti-scalant and biocide.

The pilot plant was operated in the batch and the feed and bleed modes, with feed water being topped up periodically to make up for losses due to sampling, leaks, etc. A schematic of the RO pilot plant that was used is shown in Figure 14.

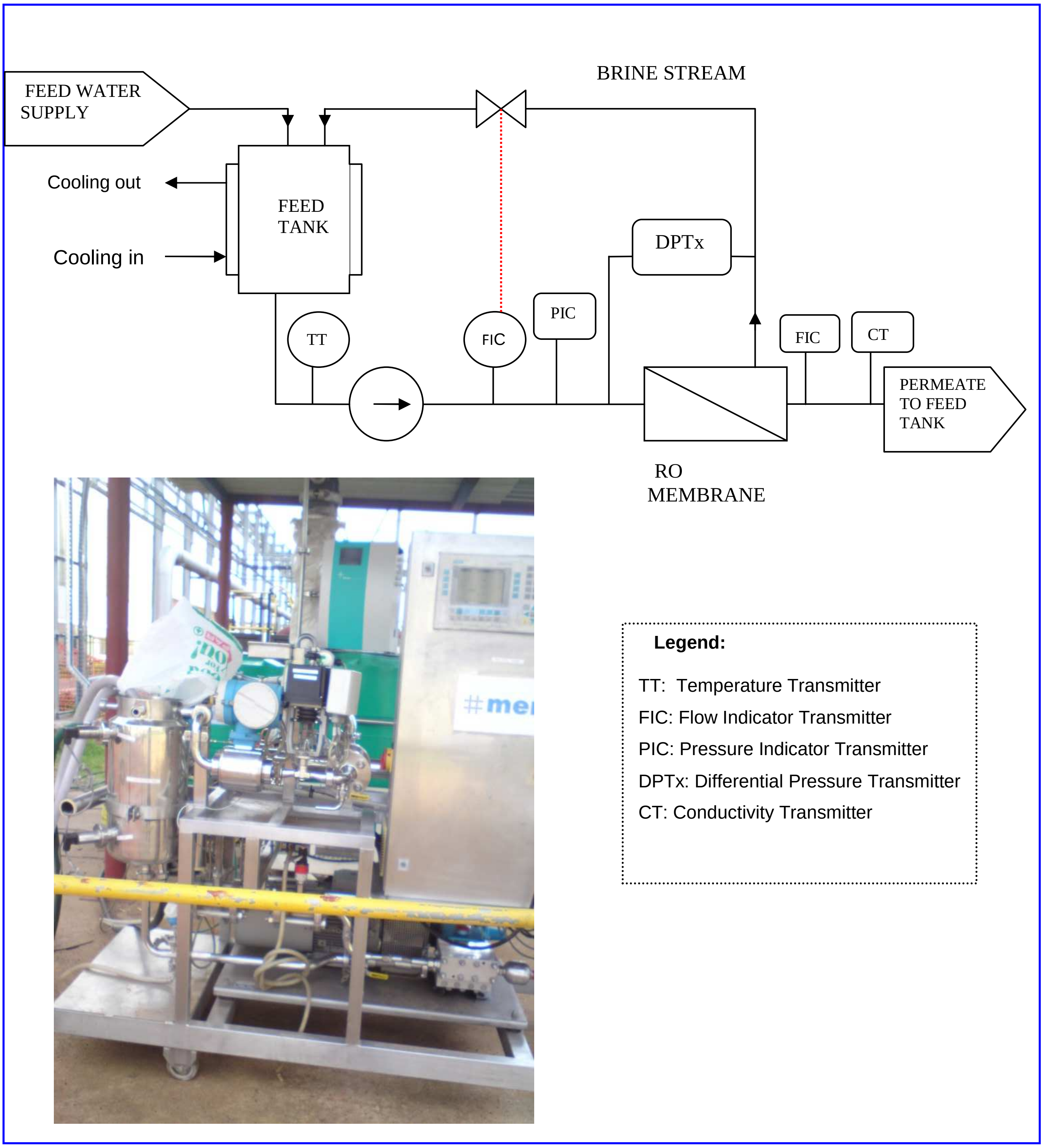


Figure 14: Schematic and photo of the RO pilot plant

A 2.5" spiral-wound RO membrane was housed in a stainless steel pressure vessel through which feed water was circulated by a plunger type feed pump from a stainless steel jacketed 20 L buffer tank. Raw water was supplied to the buffer tank as cooling water to control temperature during circulation. Both the brine and the permeate streams were returned to the feed tank during operation. The feed pressure was controlled at 10 bar. The feed flow was 900 L/h on average (maximum design feed flow is 1.4 m^3/h), thus the cross flow velocity, u, (given by Q/A_x, where Q is the feed flow rate, and A_x is the cross sectional area) was calculated to be 0.0088 m/s and 0.0072 m/s for the BW30-2540 and the BW30XFR-2540 membrane, respectively. The associated shear rate (given by 6u/H, where u is the cross flow velocity and H is the height) was 8.64 s^{-1} and 7.12 s^{-1} for the BW30-2540 and BW30XFR-2540 membranes, respectively. The equations used for the calculations are obtained from a publication by Ng & Elimelech (2004). During operation parameters such as the differential pressure, feed pressure, permeate conductivity, permeate flow, concentrate flow and feed temperature were monitored to evaluate the performance of the membrane.

Two membranes were tested during the pilot testing, namely, BW30-2540 and BW30XFR-2540 membranes. Both membranes were Filmtec membranes supplied by DOW (the membrane characteristics are given in Appendix 4). At the start of the pilot test, the standard membrane BW30-2540 was flushed with demineralised water to remove preservation chemicals and, thereafter, conditioned with a 1 000 mg/L NaCl solution before concentrated (evaporated) raw water was supplied to the membrane for the test work. A solution of 2 000 mg/L NaCl was added to the test solution after two days of operation in order to increase the conductivity of the solution.

Initially, there was no dosing regime in place for the RO pilot plant. During the first two weeks of operation, biofouling (slimy feel of piping and feed tank) and precipitation (white precipitate on the membrane) were experienced with the biofouling leading to elevated organics in the system as evidenced by an increase in TOC from 35 mg/L to levels in excess of 60 mg/L. Due to this, a dosing programme was put in place that consisted of anti-scalant dosing (to ensure no interference in terms of scaling and, hence, any reduction in performance would most probably be due mainly to organic foulants) as well as biocide to control bacterial activity.

ROSA 7.0.1 software was used to model the feed water quality. Scaling calculations (Appendix 5) from ROSA indicated the possibility of scaling (LSI >0 and Stiff & Davis Index >0) and, consequently, an anti-scalant (Genesys LF) was used to control the scaling tendencies of the water. The anti-scalant was dosed at 2.86 mg/L during RO operation with concentrated raw water feed and the dosage was 5.84 mg/L when CW UF filtered water was used as feed to the RO pilot plant. The biocide that was used was 1.5% isothiazolone at a dosage of 6 mg/L.

Similarly, at the start of the pilot test with the fouling resistant membrane BWXFR30-2540, the membrane was flushed with demineralised water to remove preservation chemicals and, thereafter, conditioned with a 2 000 mg/L NaCl solution before concentrated raw water was supplied to the membrane for the test work.

Analytical techniques and parameters measured during RO treatment

The performance of the RO membrane was also monitored in terms of salt rejection, TOC/DOC rejection and rejection of species that absorb at 254 nm. The DOW normalisation spreadsheet FTNORM was used to obtain normalised data for monitoring the performance of the RO membranes. The detailed results of the normalization are given in Appendix 6.

- TOC

TOC of the coagulated (unfiltered feed water sample) was measured using a Sievers 900 TOC analyser.

- DOC

DOC was obtained by filtering the samples on a 0.22 µm nylon syringe filter and analysing the filtered sample on the Sievers TOC analyser.

- UV absorbance at 254 nm

UV absorbance at 254 nm was measured using a 1 cm quartz cell on a Hach DR 5000 spectrophotometer analyser.

- Conductivity

Conductivity was measured in the laboratory using a MeterLab CDM210 conductivity meter.

- pH

pH measurement was undertaken on a MeterLab PHM210 standard pH meter.

- Flow

The brine and permeate flows were measured by an in-line flow meter.

- Conductivity

The permeate stream was equipped with a conductivity sensor.

- Pressure

The pressure drop, dP, across the feed and brine streams was monitored using a dP transmitter.

- Temperature

The temperature transmitter at the suction of the feed pumps was used to measure temperature for the purpose of normalisation of data.

3.7.1 Cleaning Regime

Cleaning of the RO membranes initially consisted of the use of NaOH (at pH 12) to address biofouling.

Later, the membrane cleaning was done in two steps:

- NaOH (pH 12) – first step;
- 0.1% HCl (pH 2) – second step.

Cleaning solutions used for organic fouling and biofouling were as per the recommendations in the DOW Technical Manual and are listed in Appendix 2 –2.1 Table 2B.

The rate of fouling of the RO membranes was monitored in terms of changes in the normalised permeate flow and the differential pressure that was measured using online instrumentation on the pilot plant.

3.8 MEMBRANE AUTOPSY

Membrane autopsies of the two RO membranes were conducted at an external facility and tests performed included visual (macro) examination, stereo light microscopic examination of the membrane surface, inorganic analysis of a deposit using EDX and organic analysis of a deposit (extraction with dichloromethane and analysis with FTIR spectrophotometric analysis).

3.8.1 Examination of the membrane elements

- A complete examination was conducted of the membrane elements exterior, focusing on damage or defects in o-rings and brine seals, and the condition of the feed and concentrate ends and of the outer fibreglass casing.
- The membrane elements were dissected and glue lines examined. Any colour and odours emanating from the membrane leaves were noted including the evidence of possible foulants.
- After the removal of samples for bacterial enumeration, the deposits were collected from the surface for further analysis.
- The membrane sections, corresponding to feed end, middle, and brine exit end were examined using a Research Light Microscope to determine the structural form of foulants/scalents.
- Macroscopic photographs were taken of the membrane elements as received and after cutting to see if there were any deposits on the membranes (Tigeli & Nonti, 2010).

3.8.2 Chemical analyses

- Spot tests for carbonates and sulphides were performed on the deposit removed from the BW30XFR-2540 membrane element.
- The deposit was dried and ashed at 600 ℃ and 900 ℃ to determine the percentage organic and inorganic material, respectively.
- The deposit was sent to a consulting laboratory to be analyzed for elemental compositions using Energy Dispersive X-Ray spectrometry (EDAX).
- The deposit was analyzed using Scanning Electron Microscope (SEM) to determine the structural form.
- Sections of both membrane elements from the feed end, middle, and brine exit end were tested for the presence of halogenated organics using the Fujiwara test and macroscopic photographs were taken thereafter for illustration purposes (Tigeli & Nonti, 2010).
- ATR spectrum of the white deposit from the BW30XFR-2540 membrane element was obtained to identify its composition (Tigeli & Nonti, 2010).

3.8.3 Microbiological analyses

- Quantitative microbiological swabs were conducted from three sections of each of the membrane elements (i.e. feed end, middle, and brine end) to determine the level of contamination by general aerobic bacteria, yeasts, slime formers and Pseudomonas.
- Lactophenol cotton blue staining tests were performed on the deposit collected from the BW30XFR-2540 membrane element to determine the presence of organic and inorganic material. Microscopic photographs were taken (Tigeli & Nonti, 2010).

3.9 OPTIMISATION OF PRE-TREATMENT FOR DOC REMOVAL

Activated carbon (F300 and F400) obtained from Cwenga Technologies as well as an organic scavenger (IRA958 Cl) from Rohm and Haas were used to perform preliminary adsorption tests at Komati Power Station for organics removal.

Experimental procedure/method

The pilot plant that was used is shown in Figure 15.

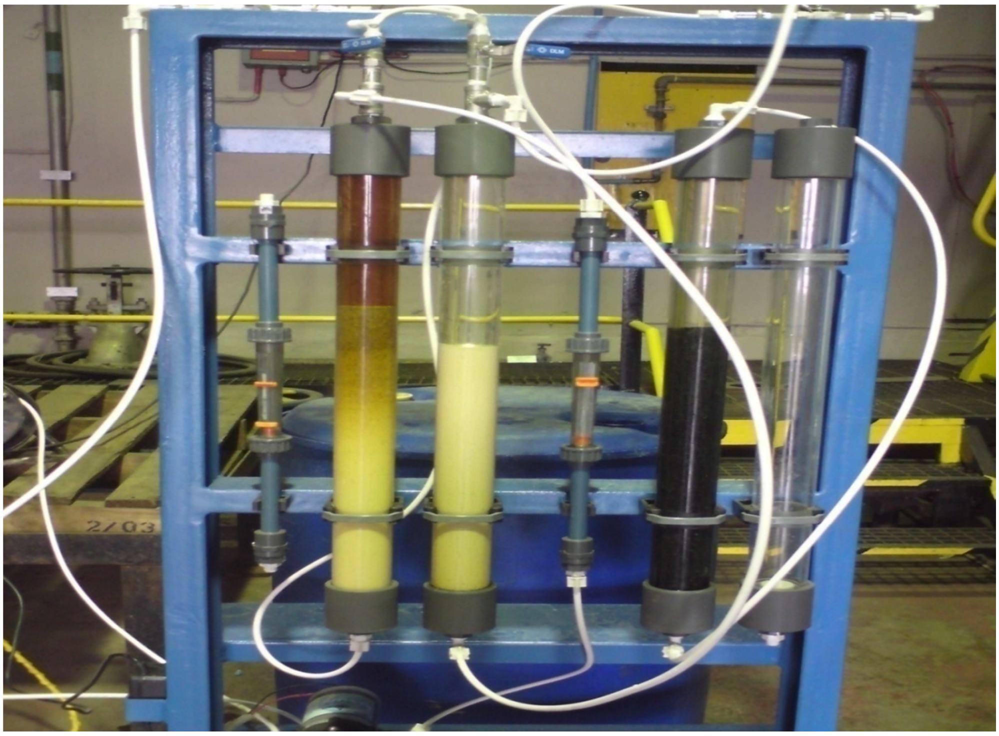


Figure 15: Activated carbon and organic scavenger test columns

The pilot plant consisted of a pump and four columns. Two columns were loaded with the organic scavenger resin while another one was loaded with activated carbon. Only one product was tested at a time; the column(s) not in use was left soaking in demineralised water.

- Granular Activated Carbon (GAC)

The carbon was washed with demineralised water to remove carbon fines prior to use. The test parameters that were used for the activated carbons tests are shown in Table 6.

Table 6 Granular Activated Carbon Test Parameters

Parameter	Value	Comment
Volume of carbon	500 mL	1 BV = 500 mL
Column diameter	43 mm	
Column height	426 mm	
Cross-sectional area of column	0.00145 m^2	
Flow rate	3.75 L/hr	7.5 BV

Empty Bed Contact Time	10 minutes	EBCTs can range from 5 to 60 minutes (Crittenden et al., 2005; Taylor & Wiesner, 1999).

The properties of the two types of granular activated carbon (Filtrasorb F300 and F400) that were used are shown in Table 7.

Table 7 Properties of the GAC (Filtrasorb F300 and Filtrasorb F400)

Property	**F300 8x30**	**F400 12x40**
Backwashed and drained bed density (kg/m^3)	460	425
Hardness number	95	95
Surface area (m^2/g)	950	1 050
Mean particle diameter (mm)	1.6	1.0
Uniformity coefficient	1.9	1.7

Source: Chemviron Carbon Product Bulletin, 2003.

Procedure for breakthrough curves and measurements

- The CW was pumped through the activated carbon column at a flow rate of 3.75 L/hr.
- Samples were taken at the outlet of the column at regular intervals.
- Samples were analysed for DOC and UV absorbance was measured to determine breakthrough of organics as measured by DOC.
- The breakthrough point was taken as the point when the outlet DOC was the same as the influent DOC (48.2 mg/L for F300 and 52 mg/L for F400).

o Organic Scavenger

An acrylic, porous, anionic resin – Amberlite IRA958 Cl – was used to conduct the organic scavenger test run. An empty bed contact time of 10 minutes was used for the test and 500 mL of resin was used and two columns (1000 mL of resin) were operated in series. The samples for DOC analysis were taken from the outlet of the second column. The regeneration of the resin was achieved by using a solution of 10% NaCl plus 2% NaOH solution, with a regenerant contact time of at least 30 minutes to ensure desorption of the organics from the resin. The resin had a total exchange capacity of ≥ 0.8 q/L (Cl form), a moisture holding capacity of 66 to 72% (Cl form) and specific gravity of 1.05 to 1.08 (Cl form) (Rohm & Haas Product data sheet, 2000).

Procedure for breakthrough curves and measurements

- Two columns were operated in parallel, with the outlet of the first column serving as feed to the second column.
- The CW was pumped through the organic scavenger resin column at a flow rate of 3.75 L/hr per column.
- Samples were taken at the outlet of the column at regular intervals.
- Samples were analysed for DOC and UV absorbance was measured to determine breakthrough of organics as measured by DOC.
- The breakthrough point was taken as the point when the outlet DOC was the same as the influent DOC (62.7 mg/L).

The performance of the activated carbon and organic scavenger resin was monitored in terms of TOC/DOC rejection and rejection of substances that absorb at 254 nm (UV_{254}).

Analytical techniques

- TOC

TOC of the coagulated (unfiltered feed waer sample) was measured using a Sievers 900 TOC analyser.

- DOC

DOC was obtained by filtering the samples on a 0.22 µm nylon syringe filter and analysing the filtered sample on the Sievers TOC analyser.

- UV absorbance at 254 nm

UV absorbance at 254 nm was measured using a 1 cm quartz cell on a Hach DR 5000 spectrophotometer analyser.

3.10 REDUCTION OF UV-ABSORBING SPECIES THROUGH THE MAIN KOMATI DESALINATION PLANT

Experimental procedure/method

The efficiency of the Komati desalination plant for the removal of organics was evaluated through monitoring of species that absorb UV at the wavelength of 254nm and TOC after the different process steps, namely, UF, first pass RO and second pass RO. Samples of the UF feed, UF permeate, first pass RO feed, permeate and brine and second pass RO feed permeate and brine were taken at various sample points on the main plant at Komati Power Station. Feed, permeate and brine (for RO) samples for each process step were filtered through 0.22 µm filters prior to analysis to determine TOC, DOC and UV_{254}.

Analytical techniques

- UV scans (UV_{254})

A Varian Cary 50 Conc scanning spectrophotometer and the Hach DR5000 spectrophotometer were used to scan through the range of 220 nm to 300 nm for species that absorb UV at the wavelength 220 to 300nm.

- TOC and DOC

The TOC and DOC were measured using the Sievers 900 TOC analyser.

3.11 EXPERIMENTAL SET-UP

The experimental pilot plant setup was designed based on the Komati Power Station desalination plant processes, in order to be able to observe performance of the pilot plants with the view to optimising the main desalination plant at Komati. The process description and process flow diagram for the Komati Power Station Desalination Plant is presented in Appendix 7.

CHAPTER 4:

RESULTS AND DISCUSSION

4.1 INTRODUCTION

Feed water (raw, concentrated raw and cooling water) quality, feed water characteristics, turbidity and organics removal (using in-line coagulation of the raw water), UF and RO membrane treatment of the feedwaters, membrane autopsies and organic removal optimisation data are presented and discussed in Chapter 4.

4.2 WATER QUALITY OF THE FEED WATER TO THE UF AND RO PILOT PLANTS

4.2.1 TOC of the Komati Power Station Raw Water

The TOC concentration of the Komati raw water to the main water treatment plant (MWTP) (prior to desalination) for the period of April 2008 to February 2011 is shown in Figure 16. The detailed results are shown in Appendix 8.1.

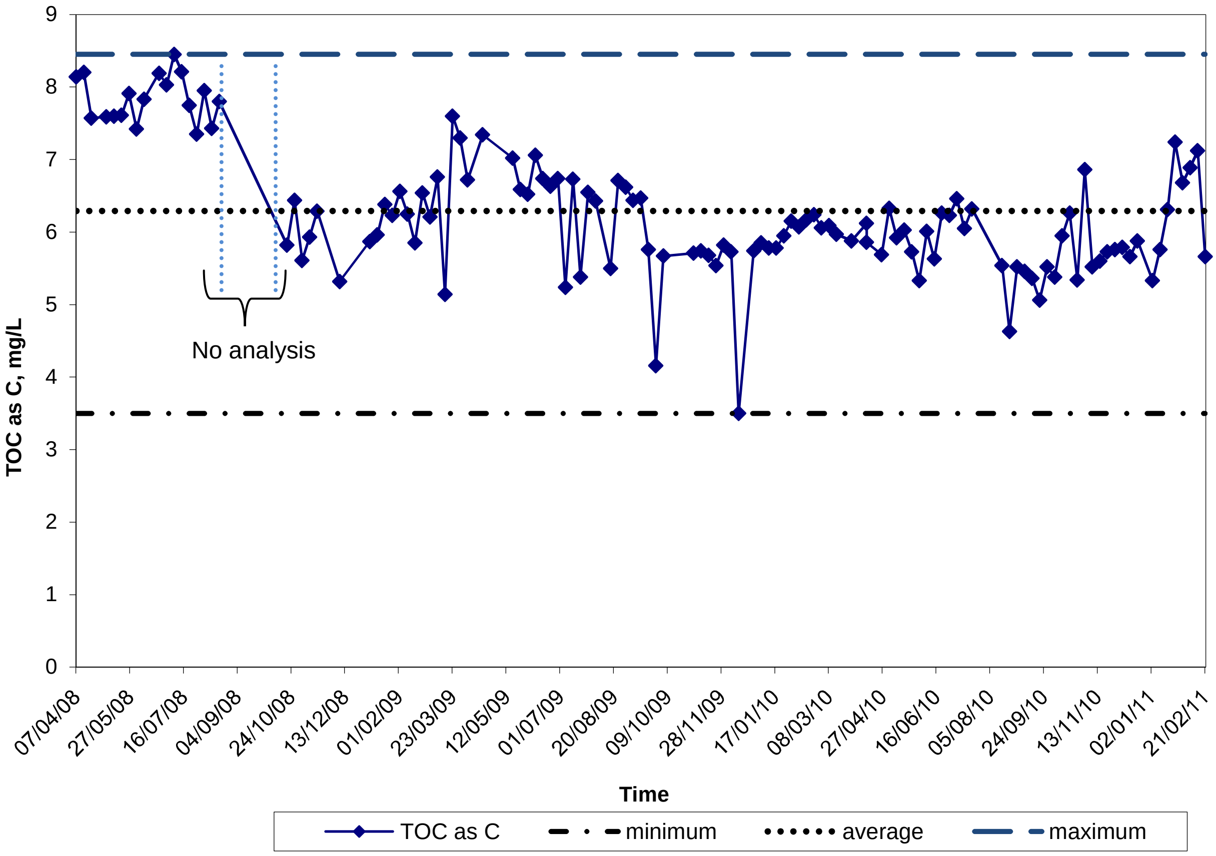


Figure 16: TOC of the raw water to the MWTP as a function of time

The Komati raw water TOC was monitored over a period of time (April 2008 to February 2011) to establish the history and seasonal variations in terms of organic matter in the raw water. The maximum raw water TOC for the time period from April 2008 to February 2011 was 8.5 mg/L, which was experienced in June 2008. The lowest (minimum) TOC concentration experienced during the time period was 3.5 mg/L during November 2009. The average TOC concentration for the time period was 6.2 mg/L.

4.2.2 TOC of the Komati Cooling Water

The TOC of the Komati cooling water (CW) from the West Forebay for the period from May 2009 to January 2011 is shown in Figure 17. The detailed results are shown in Appendix 8.2.

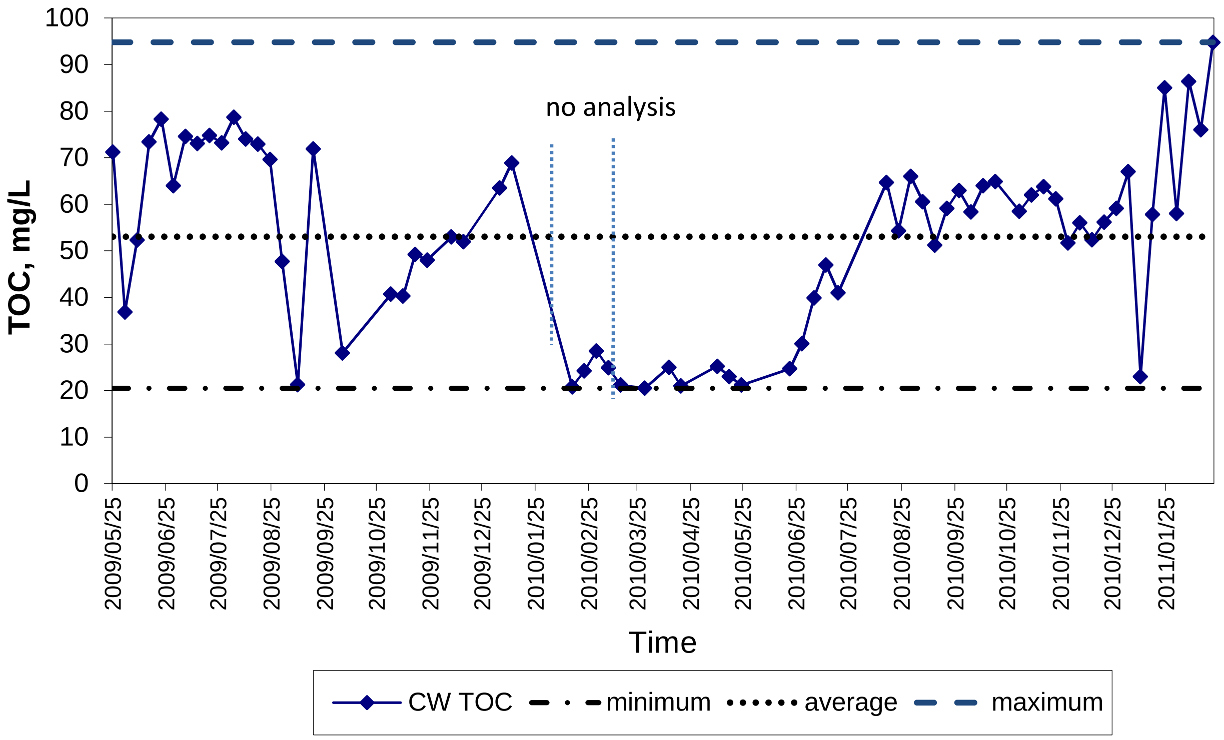


Figure 17: TOC of the cooling water as a function of time

The TOC concentration of the Komati CW was monitored over the time period from May 2009 to January 2011. During the month of June 2009, the TOC peaked at 78 mg/L and then stabilised at about 75 mg/L during the months of July and August. The TOC decreased to 28 mg/L in the beginning of October 2009 and increased steadily through the months of October to December 2009. The minimum TOC concentration experienced in the CW was 20.5 mg/L during the month of March 2010. The maximum TOC concentration (94.8 mg/L) was experienced during February of 2011 and the average TOC for the time period was 53 mg/L. Initially the high TOC concentration was due to a lack of biological control during initial start-up of the CW system when there was no implementation of the biodispersant/biocide dosing applied. During the months from February 2010 to June 2010, the TOC concentration decreased due to dilution of the CW with raw water make-up when the CW blow down was implemented. The TOC concentration later increased when the CW blow-downs were stopped due to the effluent dam being full and the concentrations of TOC in the CW system increased.

4.2.3 Feed Water Quality used in the UF and RO Pilot Plants

The raw water source for the UF pilot plant was raw water and CW, while the feed water source for the RO pilot plant was concentrated raw water (evaporated raw water). The water quality of the raw, concentrated raw and CW feed water is shown in Table 8.

Table 8 Quality of Water used during pilot plant runs

Parameter	**Raw water**	**Evaporated Raw Water**	**Cooling Water**
pH	7.61	8.57	8.47
Conductivity, µS/cm	163.8	1127	2974
Cl, mg/L	12.5	87.1	223
SO4, mg/L	19.62	126.9	844.2
M-alk, mg/L	49.2	144.6	134
Ca, mg/L	14.4	33.38	186
Mg, mg/L	8.51	22.17	126.1
Na, mg/L	9.6	93	160.8
K, mg/L	3.00	19.29	55.58
Fe, mg/L	<0.02	0.0551	0.060
Al, mg/L		<0.2	0.040
SiO2, mg/L		43	
TOC, mg/L	5.75	35	54

The concentrated (evaporated) raw water and CW TOC was concentrated by a factor of 6.1 and 9.4, respectively. The concentrated raw water pH and alkalinity were higher than that observed in the cooling water (pH of 8.57 vs. of 8.47 in the cooling water; and alkalinity of 144.6 mg/L vs. 134 mg/L in the cooling water). This is attributed to the fact that there was no pH correction during the evaporation of the raw water to concentrated raw water, while sulphuric acid was dosed into the CW to control pH and alkalinity of the water within limits (the limits for pH are controlled between 8.1 to 8.6 at 25 ℃ and M-alkalinity is controlled between 80 and 120 mg/L) specified in Eskom Cooling Water Chemistry Standard.

4.3 WATER CHARACTERISTICS OF THE UF AND RO FEED WATER

4.3.1 NOM characteristics and composition of the UF and RO pilot plants feed water

Feed water was first characterized by LC-OCD chromatographs prior to the determination of the DOC characteristics of the feed water. The LC-OCD chromatographs for the raw, concentrated (evaporated) raw and CW west forebay water are shown in Figure 18 (Huber, 2008).

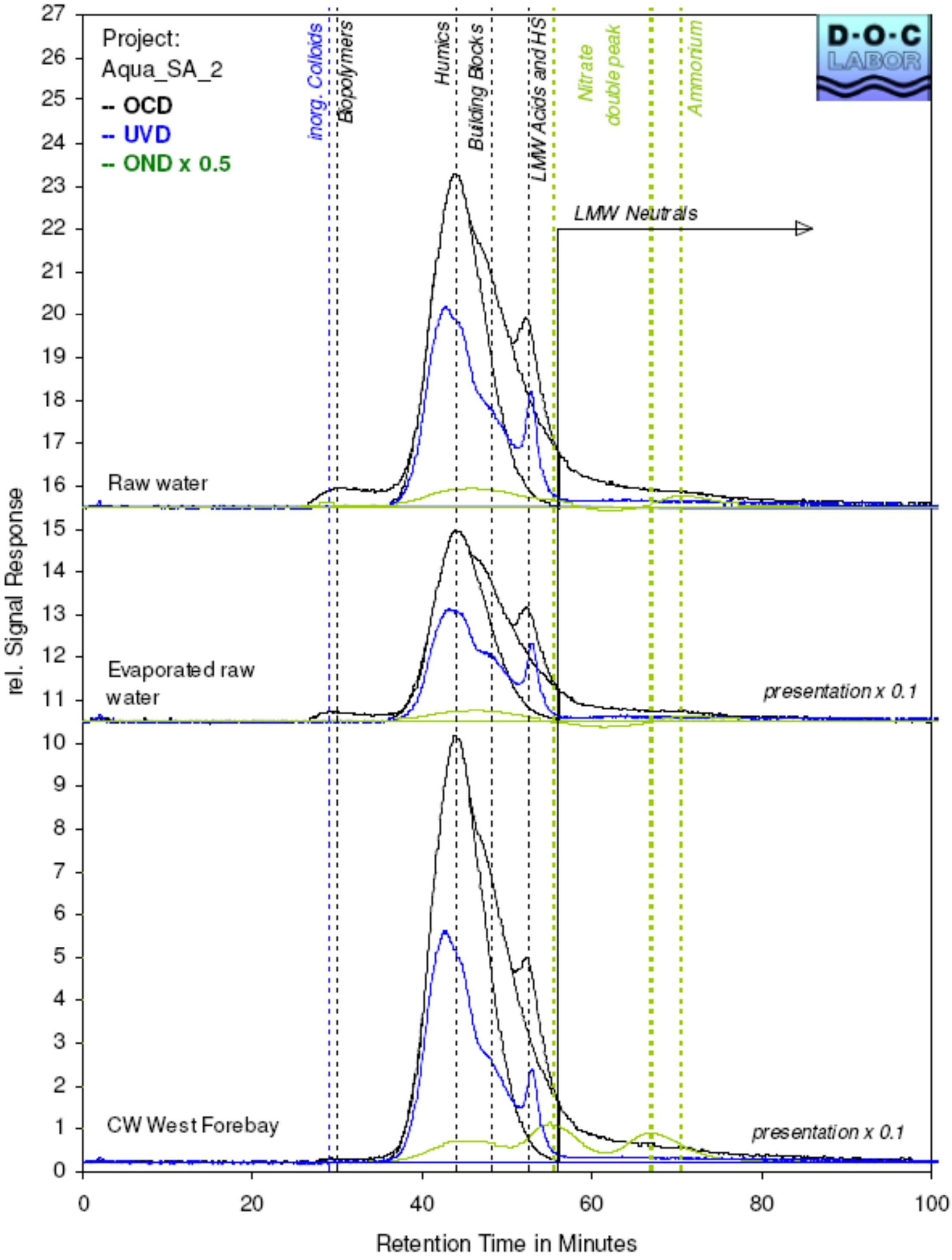


Figure 18: LC-OCD chromatographs of raw, concentrated raw and cooling water –signal response as a function of time (Huber, 2008). (OCD refers to Organic Carbon Detector, UVD refers to Ultraviolet Detector and OND refers to Organic Nitrogen Detector)

The chromatographs showed the NOM fractions that were eluted from the column. The concentration of the organics was obtained by integrating the area underneath the peaks of the chromatographs.

The NOM fractions were eluted in the following order: inorganic colloids, biopolymers, humics, building blocks, low molecular weight acids (LMWA) and humic substances (HS), nitrates and ammonia. There was a decrease in biopolymer concentration as the feed water went from raw to concentrated raw water. The peak for biopolymers that was observed for the raw and concentrated raw water disappeared in the case of the cooling water. The cooling water showed the highest peak for humics (higher than both the raw water and concentrated raw water) and it also showed traces of nitrates, while there were no nitrates in either the raw water or the concentrated raw water.

The enrichment of biopolymers in the cooling water and concentrated raw water was much less than the enrichment of humics. It was expected that the cooling water would contain high concentrations of humics, followed by building blocks and then LMWA, as could be seen from the chromatographs. This is because the cooling water is concentrated as a result of evaporative losses through the cooling towers in the cooling process. The content of biopolymers in the cooling water was very low.

The DOC composition of the raw, concentrated raw and cooling water west forebay at Komati Power Station as derived from the LC-OCD chromatographs is shown in Figures 19a, b and c, respectively (Huber, 2008).

RAW WATER

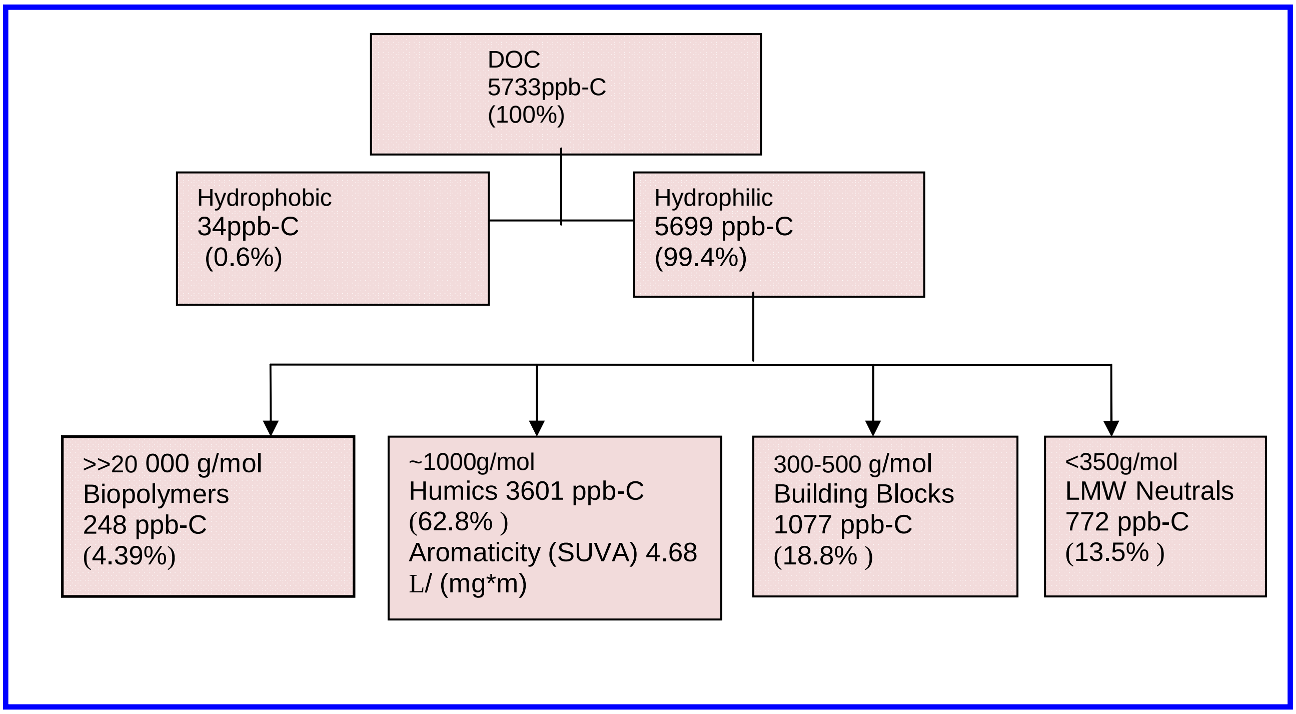


Figure 19a: Raw water DOC composition

The raw water DOC contained 0.6% of the hydrophobic fraction and 99.4% of the hydrophilic material (Figure 19a). The hydrophilic material of the raw water consisted of 4.39% biopolymers, 62.8% humic substances, 18.8% building blocks and 13.5% LMW neutrals. LMW acids are not shown as part of the DOC make-up as these were not quantifiable by the LC-OCD technique employed in the characterisation (these were less than 1 ppb-C). The humics in the raw water can be attributed to pedogenic (soil derived) fulvics. The concentration of biopolymers was found to be low in this case.

CONCENTRATED RAW WATER

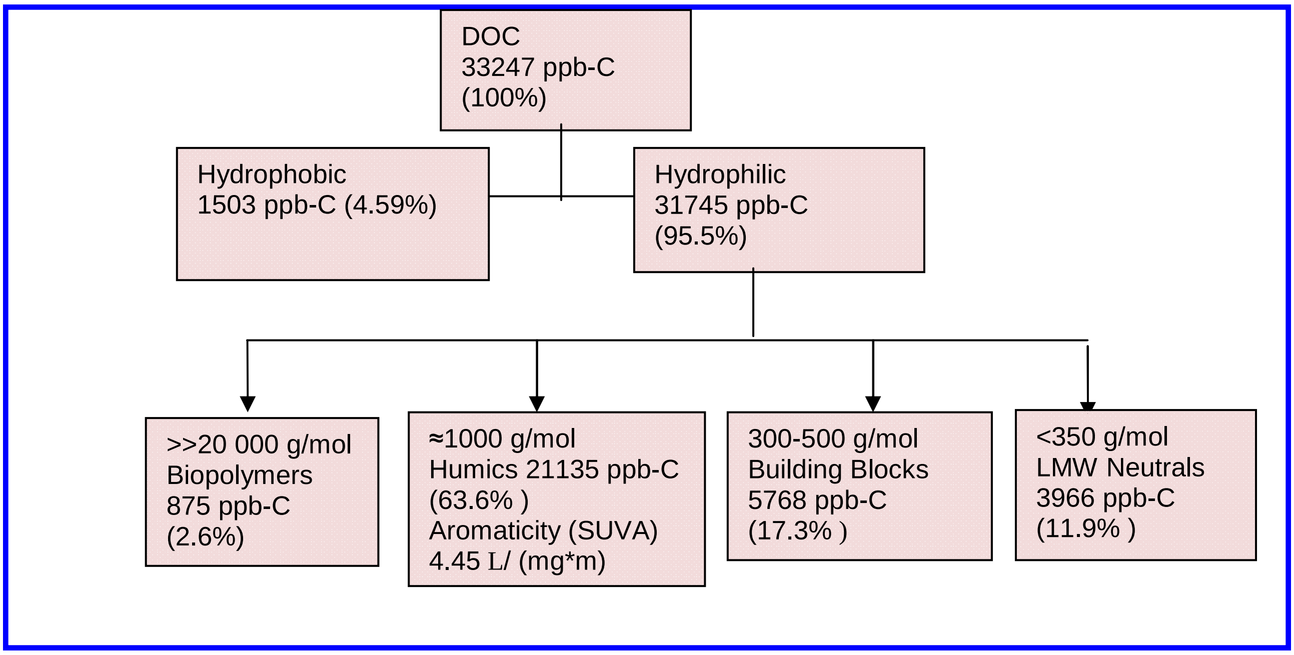


Figure 19b: Concentrated raw water DOC composition

The concentrated raw water DOC consisted of 4.59% of the hydrophobic material and 95.5% of the hydrophilic material (Figure 19b). The hydrophilic material of the raw water comprised of 2.6% biopolymers, 63.6% humic substances, 17.3% building blocks and 11.9% LMW neutrals. LMW acids are not shown as part of the DOC make-up, as these were not quantifiable by the LC-OCD technique employed in the characterisation (these were less than 1 ppb-C). The humics were concentrated by a factor of 5.9 (21135 ppb C/3601 ppb C) in the concentrated raw water and although there was a slight decrease in aromaticity, which may have been as a result of evaporative losses, the characteristics of the humics were not changed by the evaporation (the SUVA value still falls in the range given in the literature for aromatic character). Building blocks were concentrated by a factor of 5.4 (5768 ppb C/1077ppb C) in the concentrated raw water, while the biopolymers were concentrated by a factor of 3.5 (875 ppb C/248 ppb C) in the concentrated raw water.

COOLING WATER WEST

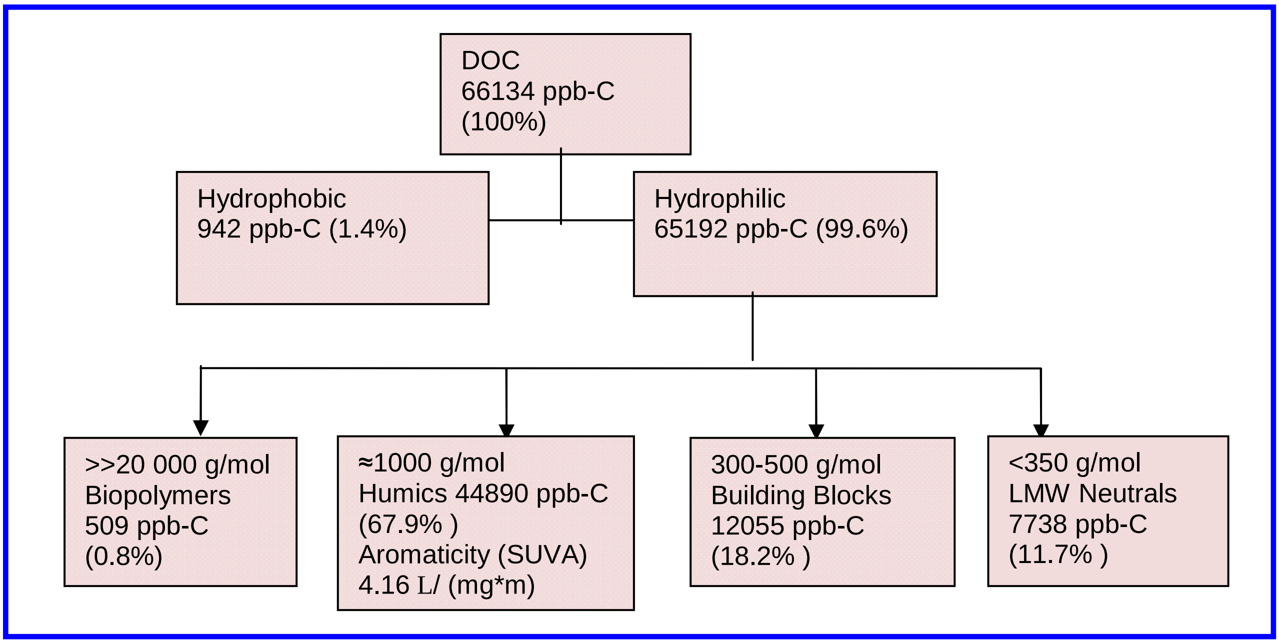


Figure 19c: DOC composition of the raw, concentrated raw and CW feed water at Komati Power Station

The CW DOC consisted of 1.4% of the hydrophobic material and 99.6% of the hydrophilic material (Figure 19c). The hydrophilic fraction of the raw water contained 0.8% biopolymers, 67.9% humic substances, 18.2% building blocks as well as 11.7% LMW neutrals. LMW acids are not shown as part of the DOC make-up, as these were not quantifiable by the LC-OCD technique employed in the characterisation (these were less than 1 ppb-C). The humics were concentrated by a factor of 12.5 (44890 ppb C/3601 ppb C) in the cooling water and although there was a slight decrease in aromaticity of the humics from 4.68 to 4.16 L/(mg*m), the character of the humics was not changed by the evaporation (the SUVA was still within the range of 4 – 5 L/(mg*m) given in the literature for aromatic character). Building blocks were concentrated by a factor of 11.2 (12055 ppb c/1077 ppb C) in the cooling water, while the biopolymers were concentrated by a factor of 2.1 (509 ppb C/248 ppb C) in the cooling water.

The NOM characteristics showed that a significant portion (>90%) of the NOM was smaller than the MWCO of 20 000 g/mol (20kDa) of the UF membrane and thus would not be retained by the UF membranes. Therefore, the UF was expected to only remove DOC material greater than 20 000 g/mol (20kDa) and this constituted of only 4.39% biopolymers (maximum in the raw water; lower in concentrated raw and cooling water as discussed above) of the DOC. The hydrophobic organic carbon (HOC) is NOM consisting of lipids, some of which may be removed by UF. The DOC is therefore expected to have a greater impact on the RO membranes as a result of the presence of a portion of humic substances, building blocks and LMW neutrals.

4.4 TURBIDITY AND DOC REMOVAL FROM THE UF PILOT PLANT FEED WATER

Jar testing was conducted to establish the coagulant of choice for turbidity and DOC removal during in-line coagulation of the feed water to the UF pilot plant. Five coagulants, namely MembraQuest 3000a (a blend of organic polymers), polyalumium chloride (PAC), U3000 (aluminium chlorohydrate- ACH, cationic), ferric chloride and alum (inorganic salts), were tested and the performance of each coagulant was evaluated in terms of turbidity and organics removal. Alum was not evaluated at a dosage of 30 mg/L due to the poor DOC removals achieved with both the 20 mg/L and the 25 mg/L dosages.

4.4.1 Turbidity Removal

The turbidity removals from the UF pilot plant feed water with different coagulants are shown in Figure 20. The detailed results are shown in Appendix 9.

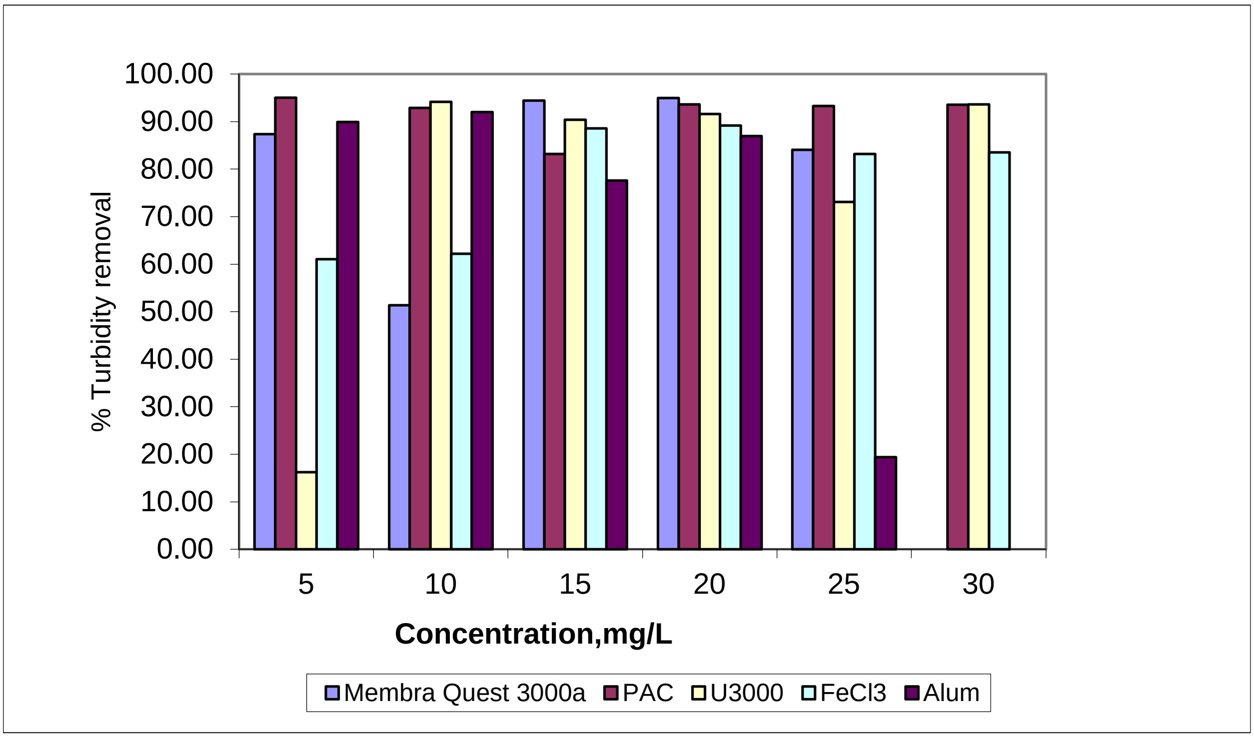


Figure 20: Turbidity removal (%) for the various coagulants as a function of coagulant dosage

The highest turbidity removals were achieved with PAC at a dosage of 5 mg/L(95.05%), followed by MembraQuest 3000a at a dosage of 20 mg/L (94.94%), MembraQueast 300a at a dosage of 15 mg/L (94.40%), U3000 at a dosage of 10 mg/L (94.15%), U3000 at a dosage of 30 mg/L (93.63%), PAC at a dosage of 30 mg/L (93.53%), PAC at a dosage of 20 mg/L (93.62%), PAC at a dosage of 25 mg/L (93.26%), PAC at a dosage of 10 mg/L (92.89%), Alum at a dosage of 10 mg/L (91.97%), U3000 at a dosage of 20 mg/L (91.62%) and U3000 at a dosage of 15 mg/L (90.38%). PAC at a dosage of 5 mg/L was the best for turbidity removal (95.05%).

4.4.2 Dissolved Organic Carbon (DOC) Removal

The DOC removal (%) from the UF pilot plant feed water for the four coagulants as a function of coagulant dosage is shown in Figure 21.

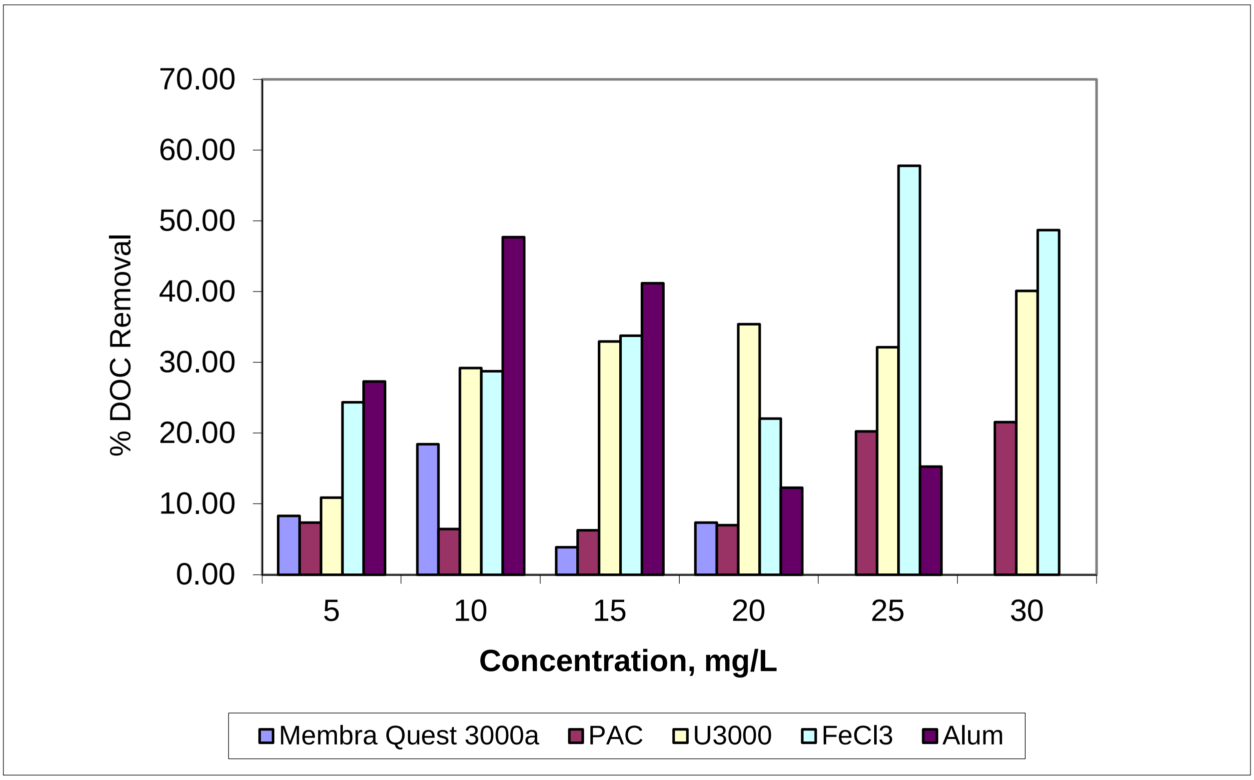


Figure 21: DOC removal (%) for the various coagulants as a function of coagulant dosage

The highest DOC removal (%) was obtained with ferric chloride ($FeCl_3$) at a dosage of 25 mg/L (57.79%) followed by $FeCl_3$ at a dosage of 30 mg/L (48.70%), alum at a dosage of 10 mg/L (47.69 mg/L), alum at at a dosage of 15 mg/L (41.20%), U3000 at a dosage of

30 mg/L (40.10%), U3000 at a dosage of 20 mg/L (35.39%), $FeCl_3$ at a dosage of 15 mg/L (33.77&) and U3000 at a dosage of 15 mg/L (32.95%). The DOC removal of 57.79% achieved with the dosage of 25 mg/L of $FeCl_3$ was disregarded as this sample had an increased retention time due to the time taken to filter and analyse the filtrate. Thus $FeCl_3$ at a dosage of 30 mg/L gave the best DOC removal (48.7%).

The DOC removal, however, must be balanced with turbidity removal for the selection of the coagulant with optimal removal for both DOC and turbidity. Alum and ferric chloride gave high removals of DOC; however, these coagulants depressed the pH and the coagulated solution had a pH of 4.9 for ferric chloride and a pH of 5.5 for alum. The operating pH of the main plant is 7.8; thus, the use of ferric chloride and alum would necessitate pH adjustment on the main plant to maintain a pH of 7.8 should these chemicals be dosed. This would entail an additional step in the process as well as additional chemical costs. In addition to this, alum and ferric chloride resulted in water that contained sulphate and chloride residuals, which would impose increased loading of chlorides and sulphates on the desalination plant and that would require removal. An increase in sulphate loading presents more costs for the desalination plant in the form of anti-scalant dosing to inhibit sulphate precipitation on the membranes.

On the other hand, dosage with U3000 had little to no effect on pH and the pH ranges at which the jar test was conducted with U3000 was close to the operating pH of the main plant, which operated at a pH of 7.8. Therefore, U3000 was selected as the coagulant of choice for testing on the pilot scale, as this coagulant provided adequate treatment in terms of both percentage turbidity and DOC removal at the plant's operating pH. However, the cost benefit justification of this option should be evaluated in future work to determine if this dose would be the most economical when compared to coagulation with ferric chloride or alum.

4.5 ULTRAFILTRATION TREATMENT OF THE RAW FEED WATER TO THE UF PILOT PLANT

4.5.1 Water Characteristics during UF Treatment

4.5.1.1 Specific Ultraviolet Absorbance

Specific Ultraviolet Absorption (SUVA) was measured on the raw water feed and permeate of the UF pilot plant. SUVA of the UF feed and permeate are shown in Figures 22 and 23, respectively. The detailed results are shown in Appendix 10.1 and 10.2 respectively.

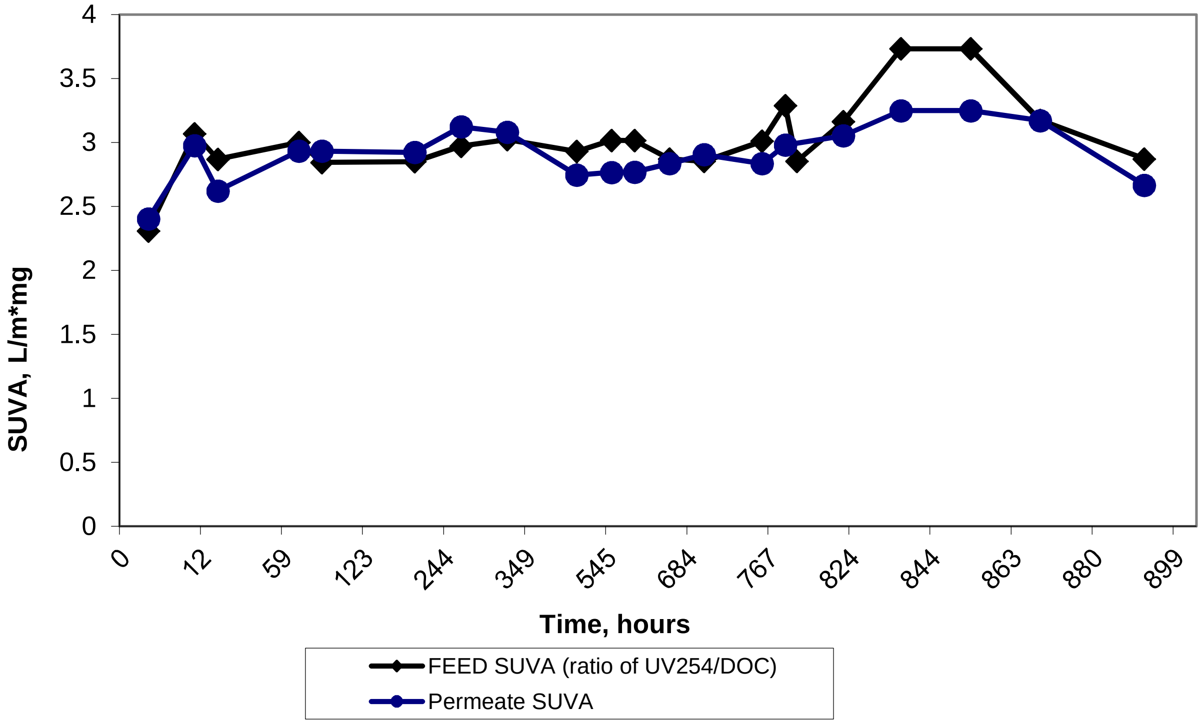


Figure 22: Comparison of feed SUVA with the permeate SUVA as a function of time for UF treatment of raw feed water without coagulation

The results (Figure 22) showed that the maximum feed SUVA obtained during the test period was 3.73 L/(mg*m) at 837 hours of operation. This is lower than the feed SUVA that was determined during characterisation (Section 4.3.1 Figure 19) of the feed water (4.65 L/(mg*m) in raw) but is still relatively high and is characteristic of surface waters (2.4–4.3 L/(mg*m) to 4.4–5.7 L/(mg*m) for natural or ground water, respectively) (Zularisam et al., 2006).

The feed SUVA is comparable to the permeate SUVA, indicating that the humics passed through the UF membrane and that there was little to no removal of humics during UF treatment of the feed water. The maximum removal of humics attained with UF without coagulation was 13.14% (when feed SUVA was 3.73 L/m*mg and permeate SUVA was 3.24 L/m*mg), as shown in Figure 22.

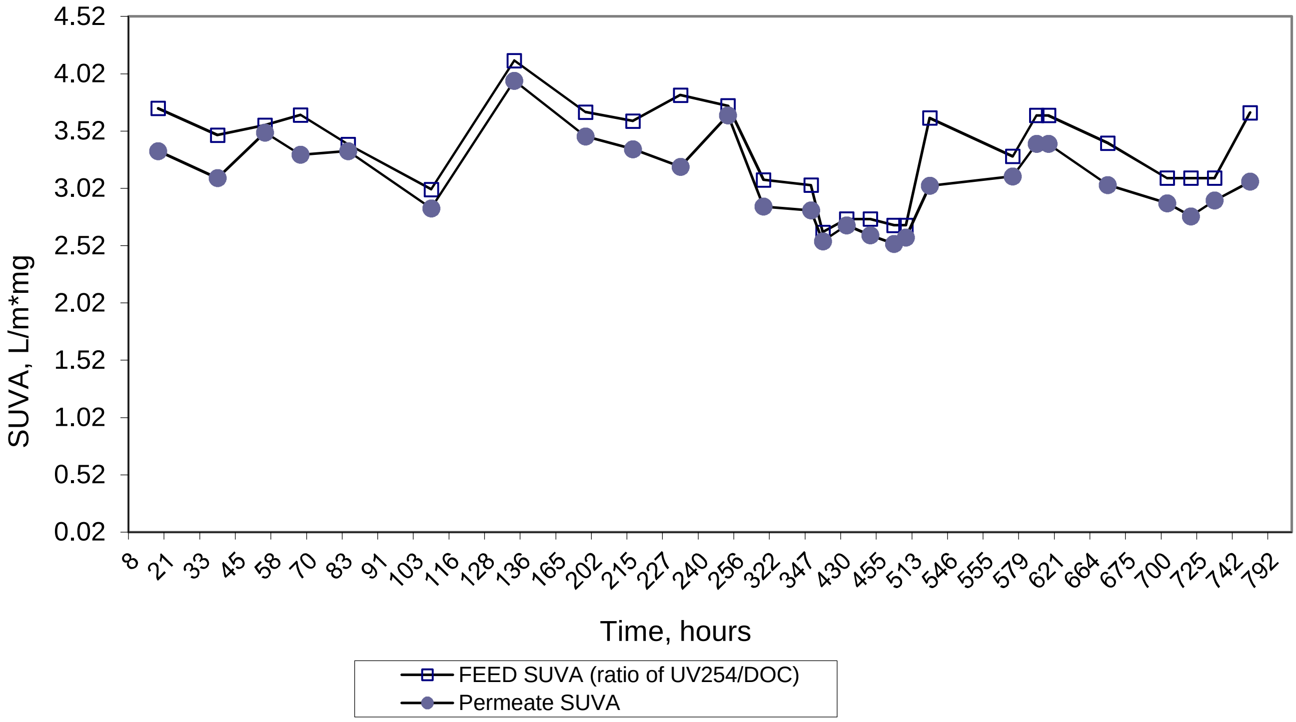


Figure 23: Comparison of feed SUVA with the permeate SUVA as a function of time for UF treatment of raw feed water with coagulation (20 mg/L dosage of U3000)

The results (Figure 23) showed that the feed SUVA was higher than the permeate SUVA when the UF membrane was operated with a coagulant. This showed that the aromaticity decreased during filtration (feed SUVA was 3.7 L/m*mg and permeate SUVA was 3.07 L/m*mg), which indicated some removal of humics with the UF membrane during operation with the coagulant. The removal of SUVA improved to a maximum of 17% (when feed SUVA was 3.7 L/m*mg and permeate SUVA was 3.07 L/m*mg) after 530 hours of operation, as could be seen in Figure 23. This showed that not all of the humic fraction was retained by the UF membrane and would therefore pass through to the RO membranes – thereby posing a fouling risk to the RO membranes.

4.5.1.2 TOC:DOC RATIO OF THE RAW FEED AND PERMEATE OF THE UF PILOT PLANT

The TOC:DOC of the UF feed and permeate is shown in Figure 24. The detailed results are shown in Appendix 10.1

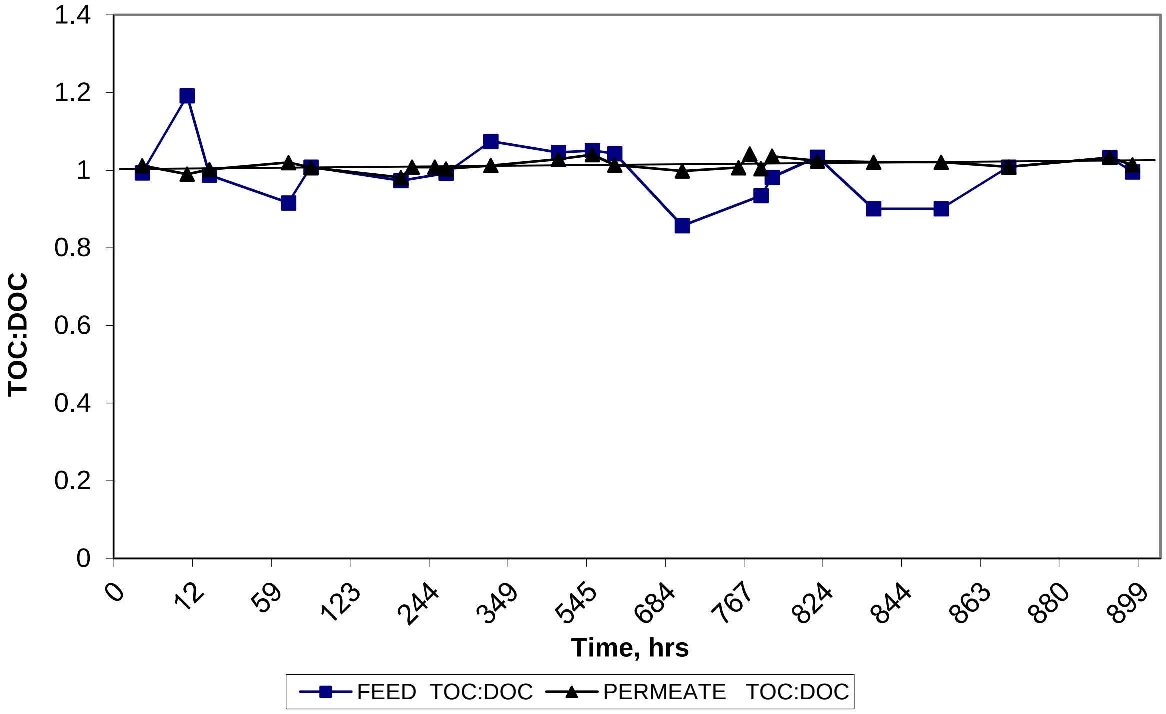


Figure 24: Comparison of the feed TOC:DOC with the permeate TOC:DOC as a function of time for UF treatment of raw feed water without coagulation

The ratio of TOC:DOC gives an indication of what makes up the bulk TOC. A TOC:DOC of ~1 shows that the TOC and DOC are approximately the same and, thus, there is little to no particulate organic carbon (POC) present. When the TOC: DOC ratio is >>1, this shows that there is a combination of DOC and POC present. The ratio of TOC:DOC of the feed and permeate streams provides an indication of the degree of DOC rejection by the membrane.

The results (Figure 24) showed that the TOC:DOC ratio of the feed stream was less than 1(minimum was 0.86, while the average was 0.99). This is close to1 which indicated that there was no or very little POC in the feed water. The permeate TOC:DOC ratio was approximately 1, which suggested that the filtered water contained mostly DOC and no or very little POC. This suggested that the DOC removal with the UF without coagulation would be very poor – increasing the organic loading on the RO membranes, which would result in a quicker decline in flux of the RO membranes, increased cleaning (implies increased loss of life of membranes) and, thus, elevated operating costs in terms of cleaning and replacement costs for the RO membranes.

The comparison of the feed TOC:DOC versus permeate TOC:DOC when the UF unit was operated with coagulation is shown in Figure 25. The detailed results are shown in Appendix 10.2.

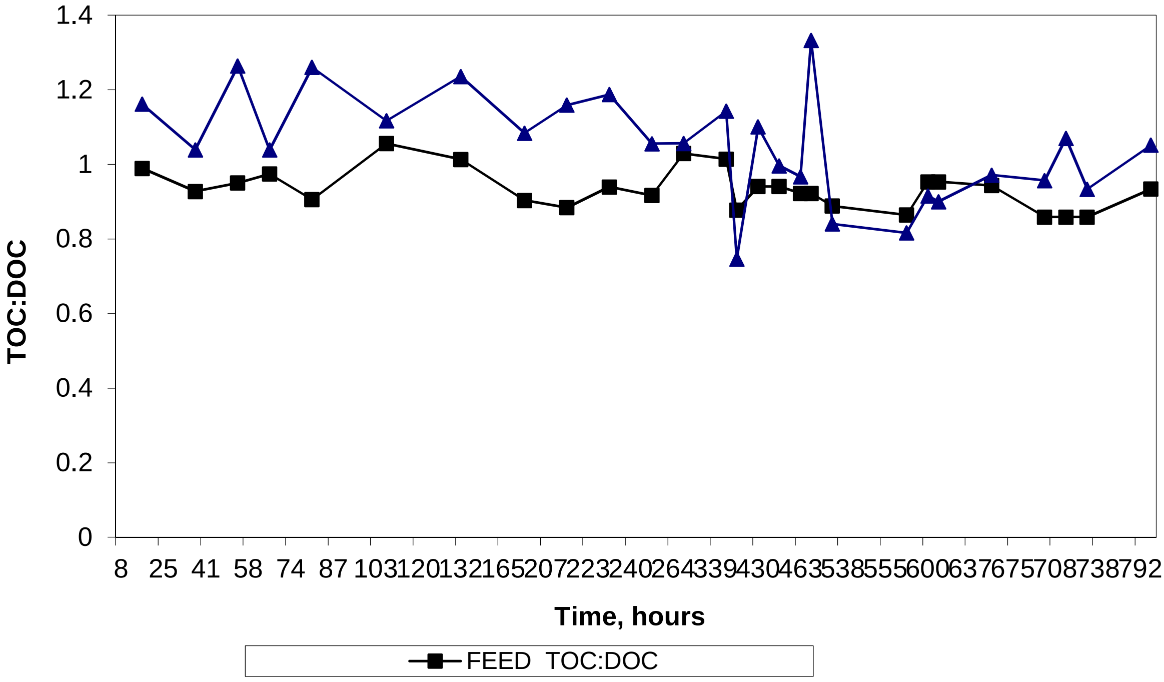


Figure 25: Comparison of the feed TOC:DOC with the permeate TOC:DOC as a function of time for UF treatment with coagulation (20 mg/L dosage of U3000)

The TOC:DOC ratio of the feed stream is close to 1- minimum of 0.84 and average of 0.93- (Figure 25). This would indicate that the TOC consisted mostly of DOC. The permeate TOC:DOC ratio was found to be greater than 1 (maximum of 1.3 and average of 1.05), which could have indicated that the water contained mostly DOC and a small fraction of POC. This suggested that operation with the coagulant, U3000 at 20 mg/L dosage, increased the particulate fraction of the permeate TOC. It was observed that the

UF permeate had a slightly milky colour. The milky colour could be ascribed to a delayed flocculation that took place causing some turbidity (POC) in the UF permeate.

4.5.2 UF treatment of the raw feed water without Coagulation

The results of the initial UF run on raw water with no coagulation are shown in Table 9. The results are for six UF cycles (before the membrane went into CEB operation) to illustrate the changes in flow, pressure and DOC through the cycles. After this initial run, readings were taken daily – where possible – instead of during each UF run.

Table 9: UF treatment of the raw feed water without coagulation of the feed – six filtration cycles

UF Cycle	Feed Pressure (kPa)	Permeate Flow (L/hr)	Permeate DOC (mg/L)	Permeate UV$_{254}$ (cm^{-1})
Feed water			**5.96**	**0.206**
1	20	3 200 to 2 700	6.43	0.172
2	20	3 200 to 2 700	No sample	No sample
3	22	3 200 to 2 700	6.19	0.179
4	22	3 200 to 2 700	6.01	0.172
5	22	3 250 to 2 700	6.05	0.171
6	22	3 200 to 2 700	6.08	0.187

The results showed that during each UF cycle the permeate flow decreased from 3 200 L/hr to 2 700 L/hr as a result of fouling. Each UF cycle was followed by a hydraulic backwash with UF permeate; thus, it was possible to determine if the membrane's performance was restored by hydraulic backwashes. This happened for UF cycles 1 and 2. From UF cycle 3, the feed pressure increased to 22 kPa, indicating that the performance of the membrane was becoming impaired as a result of membrane fouling; thus, more feed pressure was required to attain the same permeate flow through the membrane. There was no removal of DOC during each filtration cycle (feed DOC ≈ permeate DOC – see Table 9). The permeate UV absorbance at 254 nm was lower than that of the feed, indicating that there was removal of some organic species as measured by UV_{254} (maximum 16.99% removal when the feed was 0.206 and the permeate was 0.171 cm^{-1}.

4.5.2.1 Normalised Permeate Flux for UF treatment of the raw water without In-line Coagulation

The normalised permeate flux through the UF membrane that was operated without coagulation is presented in Figure 26. The detailed results are shown in Appendix 11.

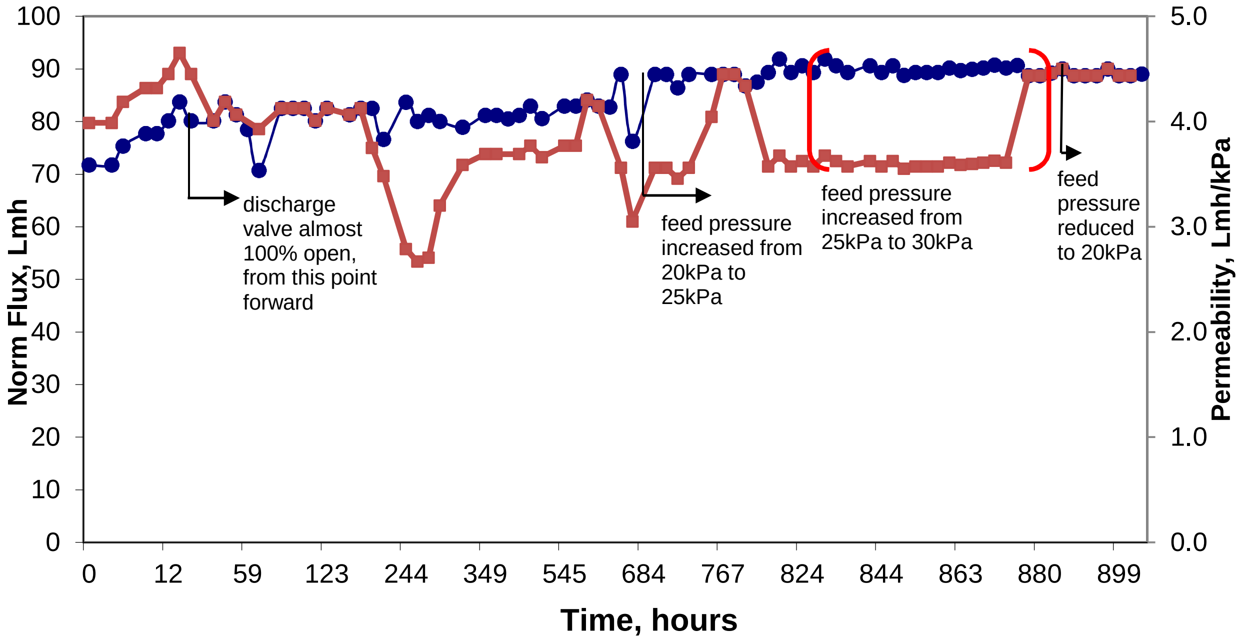


Figure 26: Normalised permeate flux as a function of time for UF treatment of the raw water feed with no coagulation

The initial permeate flux was 71.78 Lmh at the start of the operation. The flux remained at 71.78 Lmh for the first 6 hours of the run after which it increased to 75.364 Lmh (7 hours). After 8 hours of operation the flux had increased to 77.76 Lmh and remained at this value for another 2 hours and then increased to 80.15 Lmh (12 hours). After 17 hours of operation the flux had peaked at 83.7 Lmh. The increase in permeate flux was due to increased feed water flow when the feed pump discharge isolating valve was almost 100% opened. Flux declined to 80.14 Lmh after 22 hours of operation and increased to 83.74 Lmh after 34 hours of operation after which there was a steady decline in permeate flux to 70.7 Lmh (64 hours of operation). After 68 hours of operation the flux increased to 82.5 Lmh and remained stable for a number of operating hours (210 hours). After 218 hours of operation the flux had declined to 76.62 Lmh and had increased to 83.69 after 244 hours of operation. This showed that after the initial increase in flux (83.7 Lmh after 17 hours), there was a decrease in flux of 15.53% after 64 hours of operation (from 83.7 Lmh to 70.7 Lmh), followed by a further decline of 7.13% (between 210 and 218 hours of operation). The flux declined from 83.7 Lmh to 82.75 Lmh (1.12% flux decline) from 244 hours to 640 hours of operation. A slightly higher flux decline was experienced from 642 hours of operation to 810 hours of operation (88.99 Lmh to 86.79 Lmh – 2.47% decline). Thereafter, a flux decline of 3.11% (91.89 Lmh to 89.03 Lmh) was experienced from 818 hours to 900 hours of operation. The feed water pressure increased from 20 to 25 kPa, reaching a maximum of 30 kPa from 637 hours of operation until about 875 hours of operation. The increase in the feed pump pressure was required to try and maintain the permeate flux through the UF membrane which was declining due to fouling effects. The flux decline was as a result of fouling as can be confirmed by the associated decline in permeability. The flux could be recovered by performing hydraulic backwashes and CEB procedures – see Appendix 11. Permeate flux; however, could be maintained at about 90 Lmh (from 642 hours of operation).

4.5.2.2 DOC Removal during UF treatment of the raw water feed without In-line Coagulation

The DOC removal (%) as a function of time for UF treatment of the raw water feed to the UF pilot plant without in-line coagulation is shown in Figure 27. The detailed results are shown in Appendix 11.

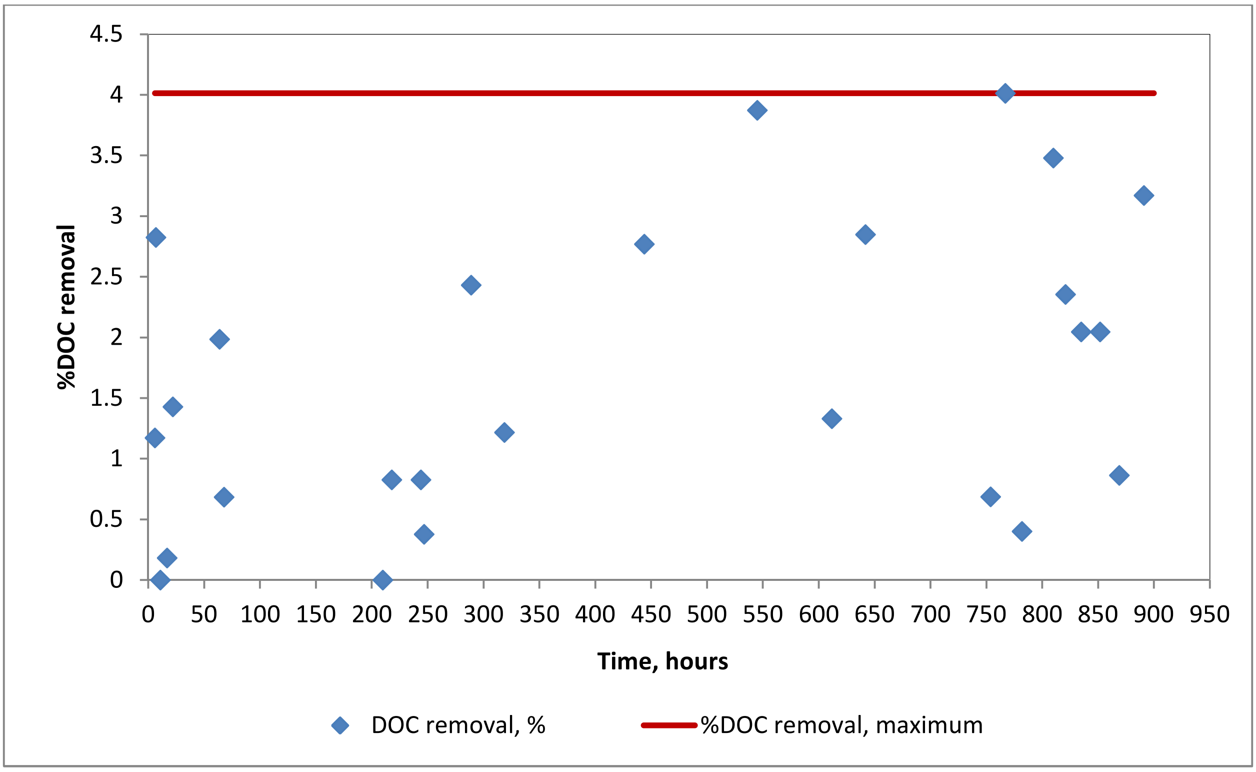


Figure 27: Percentage DOC removal as a function of time for UF treatment of the raw water feed without coagulation

The percentage DOC removal varied between 0 to 4% when the UF was operated without in-line coagulation. There was a slight reduction in permeate flow during each of the UF cycles (from 3 200 to 2 700 L/h), which showed that the membrane had fouled (Table 9). Each UF cycle was followed by a hydraulic backwash, which resulted in an increase in the permeate flow during the initial stages of each UF cycle (as illustrated in Table 9). The fouling of the membrane could be attributed to suspended solids in the feed water as well as the small fraction of biopolymers that were large enough in size to be removed with UF.

4.5.2.3 Turbidity Removal during UF treatment of the raw feed water without In-line Coagulation

The turbidity removal (%) as a function of time for UF treatment of the raw feed to the UF pilot plant without in-line coagulation is shown in Figure 28. The detailed results are shown in Appendix 11.

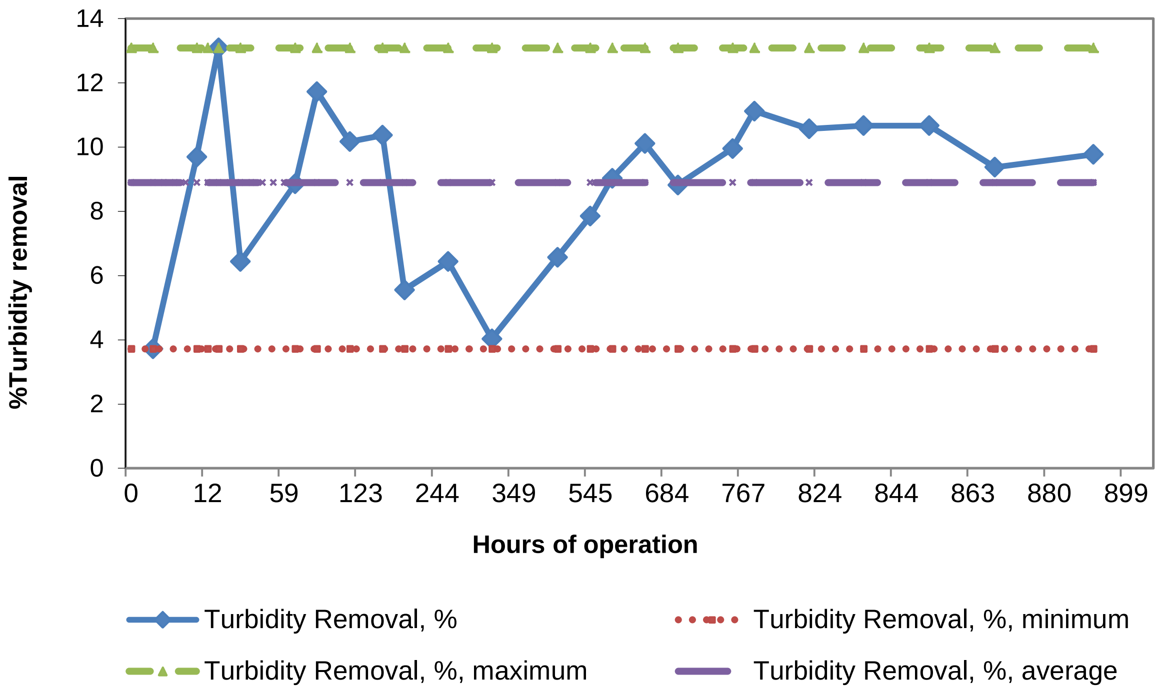


Figure 28: Percentage turbidity removal as a function of time for UF treatment of the raw feed water without coagulation

The minimum turbidity removal was 3.72% and the maximum turbidity removal obtained without coagulation was 13.09%. The average turbidity removal was 8.9%. Therefore, UF without coagulation gave poor performance in terms of turbidity removal.

4.5.3 UF treatment of the raw feed water with Coagulation

U3000 was used as a coagulant in the feed water prior to UF treatment. The chemical was dosed in-line during the UF cycle at a concentration of 20 mg/L.

4.5.3.1 Normalised Permeate Flux for UF treatment of the raw feed water with In-line Coagulation

The normalised permeate flux as a function of time for UF treatment of the raw feed water with in-line coagulation is shown in Figure 29. The detailed results are shown in Appendix 12.

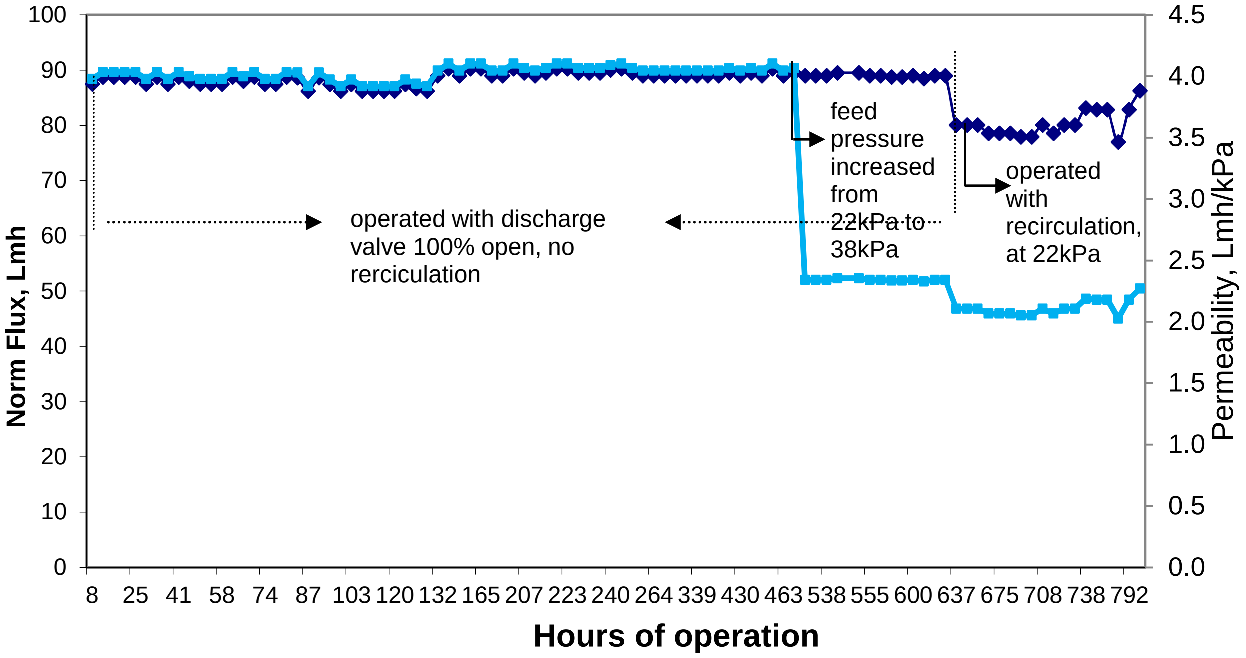


Figure 29: Normalised permeate flux as a function of time for UF treatment of the raw feed with in-line coagulation

The initial flux was 87.5 Lmh and increased to 88.7 Lmh after 12 hours of operation. The flux remained at 88.7 Lmh until 25 hours of operation after which it declined to 87.5 Lmh (1.35% flux decline). This pattern was repeated for the first 132 hours of operation with the flux varying between 88.7 Lmh and 86.7 Lmh (2.25% decline). The flux decline indicated that some membrane fouling took place. The flux then varied between 90.2 Lmh and 89.5 Lmh and remained stable at approximately 88.9 Lmh from 256 hours of operation to 621 hours of operation, however from 500 hours of operation to 621 hours of operation the feed pressure was increased from 22 kPa to 38 kPa to maintain the permeate flux through the membrane. During this period there was a decline in permeability which indicates that the membrane was fouled and the increased feed pressure was required to maintain permeate flux.

After 629 hours of operation the feed water was re-circulated from the discharge to the suction of the feed pump to achieve better mixing of the coagulant (see Figure 12 for the UF schematic). Introduction of the recirculation resulted in a reduction in permeate flow at 637 hours and, hence, reduced flux (declined from 88.9 Lmh to 80.1 Lmh– 9.9% decline). There was a further decline in flux (to 78 Lmh) after 671 hours of operation.

CEB cleaning of the membrane after this period resulted in a flux recovery (to 83 Lmh at 738 hours) which was less than the initial flux (87.5 Lmh). It was observed after 708 hours of operation that the flux further declined from 83 Lmh to 77 Lmh (7.2% decline) although this was not sustained as the flux thereafter increased to 80.08 Lmh and continued to increase with the recirculation valve 100% open . The % reduction in flux prior to cleaning was less than the 15% which is acceptable according to the industry norm of 15%. It appeared that flux could be maintained at around 90 Lmh (without circulation) which was about the same as when no coagulant was applied (see Figure 26).

4.5.3.2 DOC removal from the raw water feed to the UF pilot plant with In-line Coagulation

The DOC removal (%) as a function of time for UF treatment of the raw feed to the UF pilot plant with in-line coagulation is shown in Figure 30. Detailed results are shown in Appendix 12.

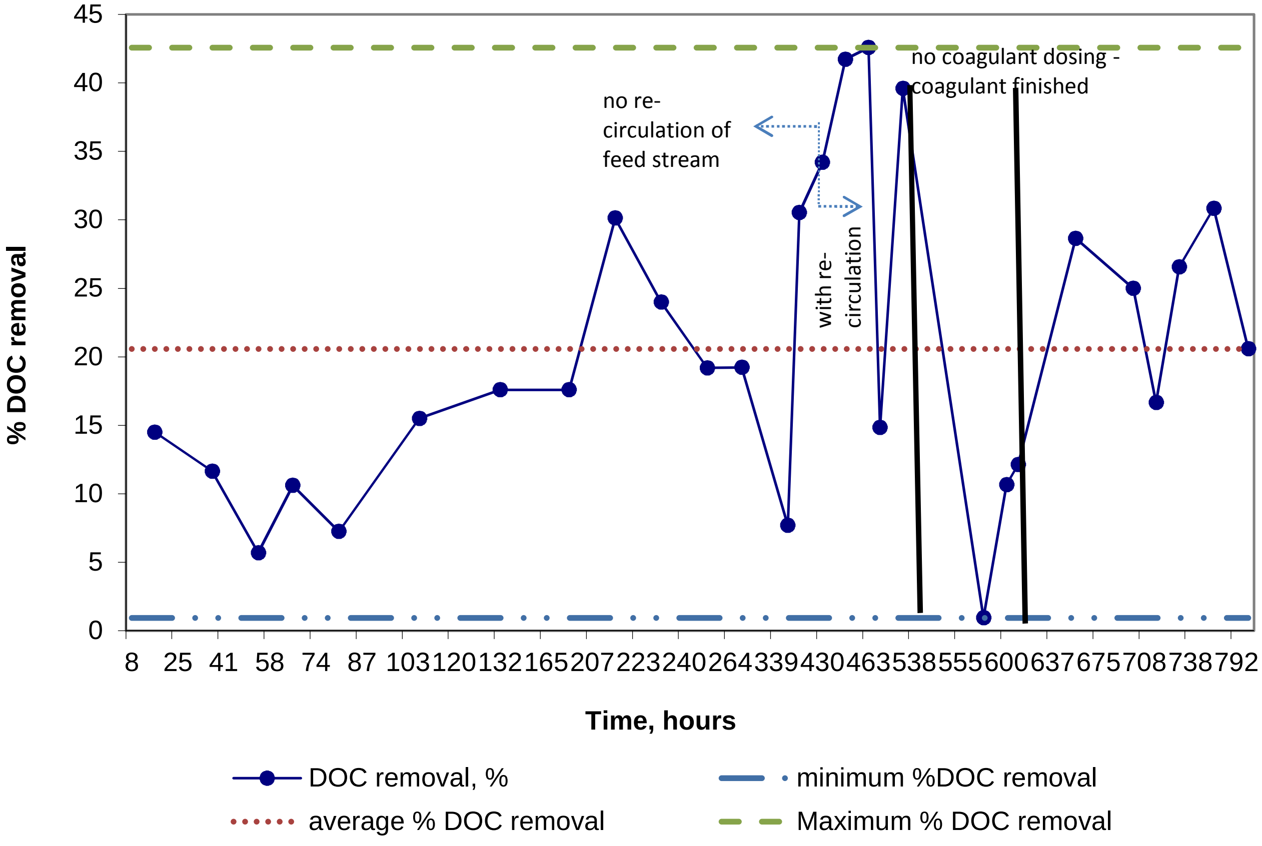


Figure 30: DOC removal (%) as a function of time for UF treatment of the raw feed with in-line coagulation (20 mg/L of U3000)

The results showed that when the UF pilot plant was initially operated with in-line coagulation that the DOC removal obtained was 14.51%. The minimum DOC removal that was obtained was 0.95%. This was obtained during the period when the coagulant dosing tank was pumped below the suction line of the dosing pump. Visual observation of the filtered water suggested that some coagulant had passed through to the permeate stream. The average DOC removal obtained was 20.58% and the maximum DOC removal achieved was 42.58%. This was achieved when the recirculation of the feed stream (recirculation flow of 60 L/hr) from the discharge to the suction side of the feed pump was introduced to achieve better mixing of the coagulant. The DOC removal then improved from 30.13% to 42.58% (from 355 hours to 465 hours of operation) when the recirculation was introduced after 355 hours of operation. There was a decrease in the DOC removal during the operating period between 571 hours and 612 hours because there was no coagulant dosing during this time.

Since U3000 only gave a DOC removal of approximately 35% at a dosage of 20 mg/L, this suggested that the in-line coagulation of the UF feed was not sufficient to reduce the DOC loading (UF permeate DOC > 55 mg/L) on the RO membranes (original equipment manufacturer requirements dictate DOC of less than 6 mg/L and ideally less than 3 mg/L) and, hence, this process would need further optimisation for DOC removal. However, the nature of the organics also did not lend itself to removal with UF, as most of the organic fractions passed through the UF membrane as could be seen by the characterisation LC-OCD chromatographs of the feed water (see Section 4.3.1). This suggested that other options for reducing the TOC loading on the RO membranes should be identified for the Komati feed water, which is high in NOM.

4.5.3.3 Organics removal as measured by UV Absorbance as a Function of Time for UF treatment of the raw feed water with In-line Coagulation

Organics removal as measured by UV absorbance removal (%) as a function of time for UF treatment of the raw feed water to the UF pilot plant with in-line coagulation is shown in Figure 31. The detailed results are shown in Appendix 12.

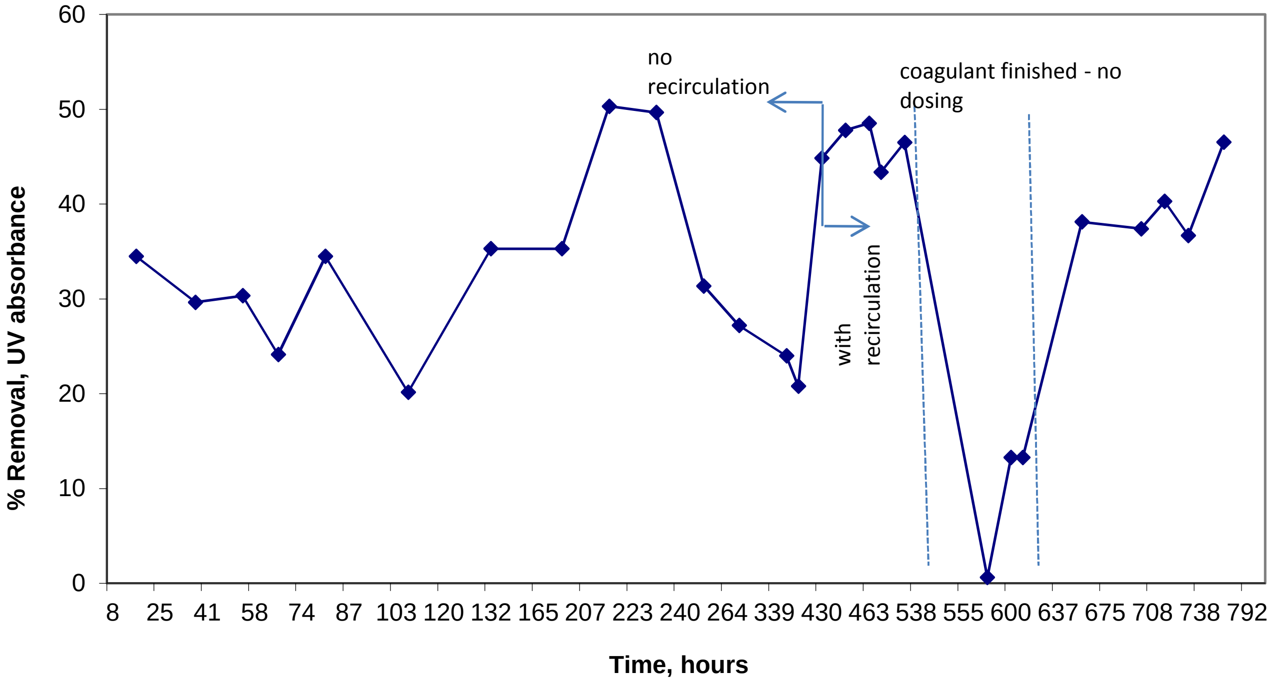


Figure 31: Organics removal as a function of time for UF treatment of the raw feed water with in-line coagulation (20 mg/L of U3000)

The initial organics removal was 34.48%. The removal declined to 20.15% after 107 hours of operation and increased from 35.3% to 50.33% (between 132 hours and 215 hours of operation). This indicated that in-line coagulation resulted in improved organics removal as opposed to operation with no coagulation.

4.6 REVERSE OSMOSIS TREATMENT

4.6.1 RO treatment of the concentrated raw water without pre-treatment

4.6.1.1 Normalised Permeate Flux without Pre-treated Water

The normalised permeate flux that was obtained with the standard BW30-2540 membrane without pre-treated water is shown in Figure 32. The detailed results are shown in Appendix 13.1.

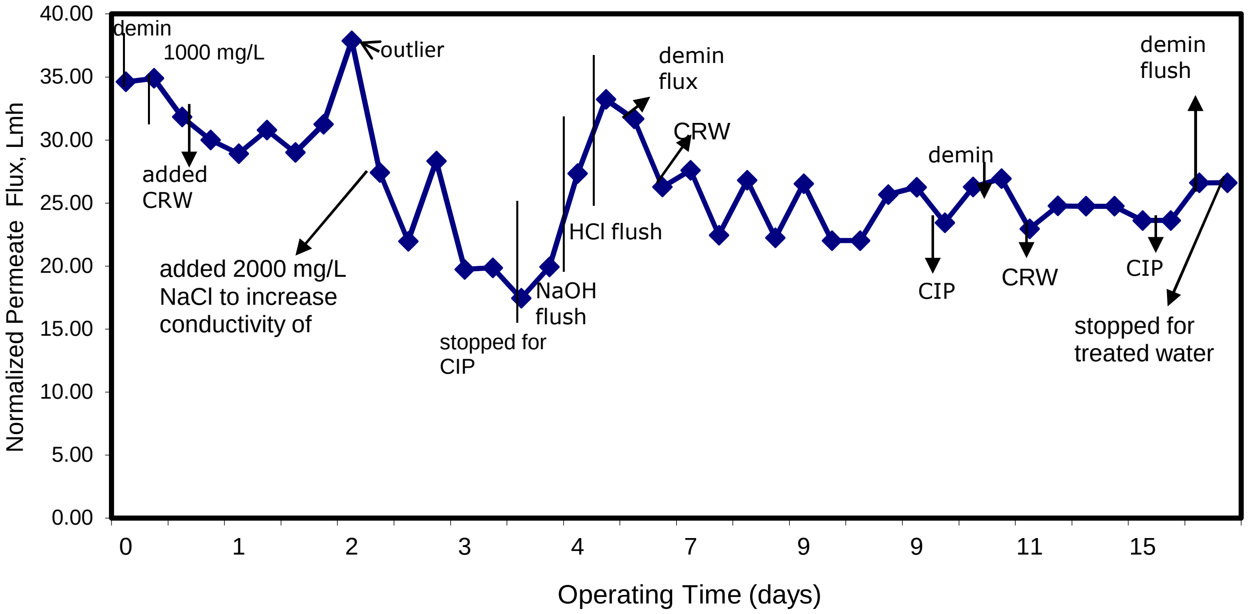


Figure 32: Normalised permeate flux as a function of time for the BW30-2540 membrane without pre-treatment

(Note: graph is not to scale. Number of data points plotted varied from day to day of operation-see Appendix 13.1).

The clean water flux (CWF) was 34.62 Lmh (demineralized water was used to determine the clean water flux of the membrane). This increased slightly to 34.90Lmh during the two hours of flushing with demineralised water. The membrane was conditioned with 1 000 mg/L of NaCl after 2 hours of operation to check the salt rejection of the virgin membrane and at this point the permeate flux declined to 31.82 Lmh (after 2 hours, first day of operation).The decline in permeate flux was as a result of the increase in the osmostic pressure of the solution after addition of 1 000 mg/L of NaCl. A flux of 30.01 Lmh was obtained when the concentrated (evaporated) raw water (CRW) feed was added. This was expected due to increased osmostic pressure of the feed water, which resulted in the decline in permeate flux. The flux then continued to decline to 28.93 Lmh, thereafter increased to 30.78 Lmh and peaked at 37.88 Lmh (second day of operation). The permeate flux of 37.88 Lmh was an outlier that was attributed to inaccuracies in the flow measurements and human error as a result in change of operating personnel. During the second day of operation 2000 mg/L of NaCl was added to the feed water to increase the conductivity of the feed to simulate the CW conductivity. The flux declined from 37.88 Lmh to 21.97 Lmh (second day of operation). The decrease in permeate flux was attributed to the increase in the osmostic pressure of the feed solution due to the increase in the salt concentration of the solution. The permeate flux continued to decline to 19.7 Lmh on day 3 of the operation (flux decline of 34.36% - 30.01 Lmh to 19.7 Lmh) and consequently, a CIP (caustic and acid) was performed on day 4. The CWF attained after the CIP was 31.68 Lmh compared to 34.62 Lmh of the virgin membrane (8.49% reduction). When CRW was added after the CIP the permeate flux attained was 26.27 Lmh (day 6). The flux declined to 22.03 Lmh (day 9), and then increased to 26.26 Lmh (day 9). When the flux declined to 23.44 Lmh (day 10) (flux decline 10.77% - from 26.27 Lmh to 23.44 Lmh), another CIP was performed (day 10). The CWF obtained after the CIP was 26.94 Lmh (22.18% reduction compared to CWF of virgin membrane – 34.62 Lmh). The flux declined to 22.96 Lmh when CRW was added (day 11), then increased to 24.77 Lmh and remained stable until day 13 after which the flux declined to 23.61 Lmh and a CIP was performed (day 15) (flux decline of 4.68% - from 24.77 Lmh to 23.61 Lmh). The CWF obtained after the CIP was 26.61 Lmh. After the CIP the unit was stopped for the operation with treated water. The overall CWF decline was 23.14% (from 34.62 Lmh to 26.61 Lmh).

It can be seen from the results that initially the flux declined to a value well above the accepted 15% decline prior to a CIP (34.36%) after which the flux declined to values within 15% decline prior to cleaning. The cleaning cycles did not restore the flux to the original CWF (restored to 26.94 Lmh and 26.61 Lmh compared to initial CWF of 34.62 Lmh).

The normalised permeate flux for the low-fouling membrane (BW30XFR-2540) test run without pre-treated water is shown in Figure 33. The detailed results are shown in Appendix 13 –13.2.

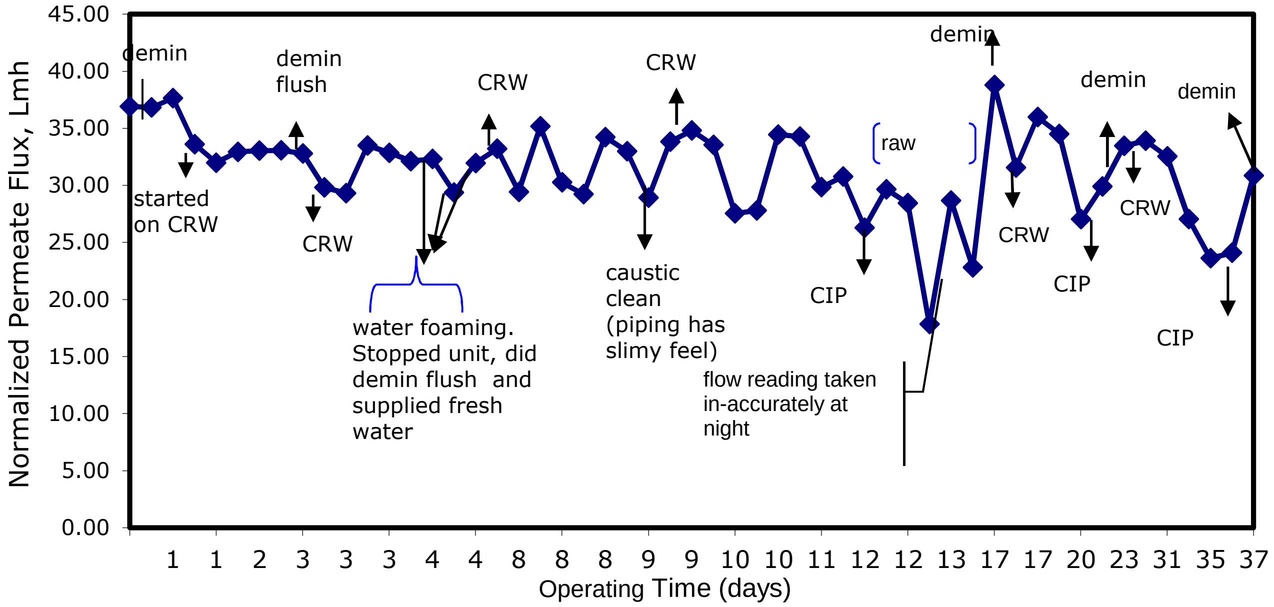


Figure 33: Normalised permeate flux as a function of time for the BW30XFR-2540 membrane without pre-treatment

(Note: graph is not to scale. Number of data points plotted varied from day to day of operation, see Appendix 13.2).

The initial CWF was 36.92 Lmh (demineralised water was used to flush preservation solution and to determine the clean water flux of the membrane). The membrane was conditioned with 2 000 mg/l of NaCl after 2 hours of operation to check the rejection of the virgin membrane and at this point the permeate flux increased to 37.64 Lmh. It was expected that the permeate flux would decline after addition of the salt solution as a result of the increase in the osmotic pressure. Since the feed pressure was not changed (increased), the increase in permeate flux could only be attributed to measurement errors when the permeate flow was measured. A permeate flux of 33.61 Lmh was obtained when the concentrated raw water (CRW) feed was added (day 1). This decrease in the permeate flux was due to the higher osmotic pressure of the feed solution. The flux continued to decline to 31.99 Lmh (day 1), thereafter increased to 32.94 Lmh (second day of operation) and remained stable at approximately 33 Lmh. On day 3 of operation the membrane was flushed with demin water (flux 32.80 Lmh). The flux declined to 29.82 Lmh when CRW was added (day 3) and thereafter increased to 33.50 Lmh (day 3). On day 4 of operation the flux declined to 29.39 Lmh at which point the feed tank was decanted due to foaming of the water. A new batch of feed water was evaporated at the laboratory during which time demineralised water was circulated through the membrane (until day 8). When fresh feed water was added to the tank, the permeate flux attained was 33.20 Lmh (day 8 of operation). The flux did not stabilize but declined steadily. On day 9 of the operation the flux had declined to 28.91 Lmh (12.92% flux decline- from 33.20 Lmh to 28.91 Lmh) at which point a high pH CIP was performed with caustic as biological fouling was suspected (due to slimy feel of the tank and piping). The permeate flux obtained when CRW was added after the CIP was 34.80 Lmh (day 9). The flux declined to 27.54 Lmh on day 10 of operation and later increased to 34.45 Lmh. On day 12 of operation the flux had declined to 26.29 Lmh (flux decline 24.45% - from 34.80 Lmh to 26.29 Lmh) and a CIP (caustic and acid) was performed. After the CIP the unit was operated on raw water due to shortage of CRW (from day 12). Thereafter the unit was flushed with demineralised water on day 17 of operation (CWF was 38.80 Lmh – this was higher than the CWF of the virgin membrane (36.92 Lmh)) and the unit was started again on CRW. The permeate flux attained when CRW was added was 31.56 Lmh (day 17). The flux increased to 35.98 Lmh (day 17) after which it declined steadily until it reached 27.04 Lmh when a CIP was done on day 20 (flux decline of 14.32% - from 31.56 Lmh to 27.04 Lmh). The CWF obtained after the CIP was 29.92 Lmh (day 22), this was lower compared to the CWF of 38.80 Lmh previously obtained, which suggested that the membrane did not completely regain the flux through cleaning. After the CIP the feed tank was made up with CRW (day 23). The permeate flux attained was 33.46 Lmh. The flux increased slightly to 33.90 Lmh (day 29). The unit operated on CRW with the flux declining steadily until it reached 23.61 Lmh (day 35) and another CIP was performed on day 35 of operation (flux decline of 29.44% - from 33.46 Lmh to 23.61 Lmh). After the CIP the unit was flushed with demineralised water (day 37) in preparation for operation with pre-treated water and the CWF obtained was 30.87 Lmh.

The cleaning cycles did not always restore the flux to original values. Initially, the CWF obtained after CIP operations was comparable to that of the virgin membrane (38.80 Lmh vs. 36.92 Lmh of the virgin membrane, 4.85 % decline) which suggested that the CIP operations had been successful. Subsequent CIP operations, however, were not as as successful, and the CWF obtained was lower than that of the virgin membrane (29.92 Lmh and 30.87 Lmh vs. 36.92 Lmh). The permeate flux could not be maintained through the operating cycles and it declined steadily through a series of cleans. The flux declined to values well above the accepted 15% decline prior to CIP on two occasions. The overall flux decline was 29.75% (from 33.61 Lmh initial permeate flux to 23.61 Lmh).

The low fouling membrane achieved better CWF recovery after the cleaning cycles (81.26% - approximately 30 Lmh vs. 36.92 Lmh of the virgin membrane on the occasions when there was flux loss) compared to the standard membrane (restored to 26.94 Lmh and 26.61 Lmh- average CWF of 26.78 Lmh, compared to initial CWF of 34.62 Lmh, which represented a recovery of 77.35%). The flux decline was quicker for the standard membrane (membrane had to be cleaned after 2 – 3 days of operation compared to the low fouling membrane which operated for longer periods – ≥4 days prior to cleaning). Therefore the low fouling membrane had a slight advantage over the standard membrane in terms of fouling resistance and flux regained after the cleaning cycles.

4.6.1.2 TOC Rejection during RO Treatment of the Concentrated Raw Water Feed Without Pre-treated Water

The TOC (%) rejection of the standard BW30-2540 and the low fouling BW30XFR-2540 RO membranes during treatment of the concentrated raw water feed without pre-treated water is shown in Figure 34. The detailed results are shown in Appendix 13.1 and 13.2, respectively.

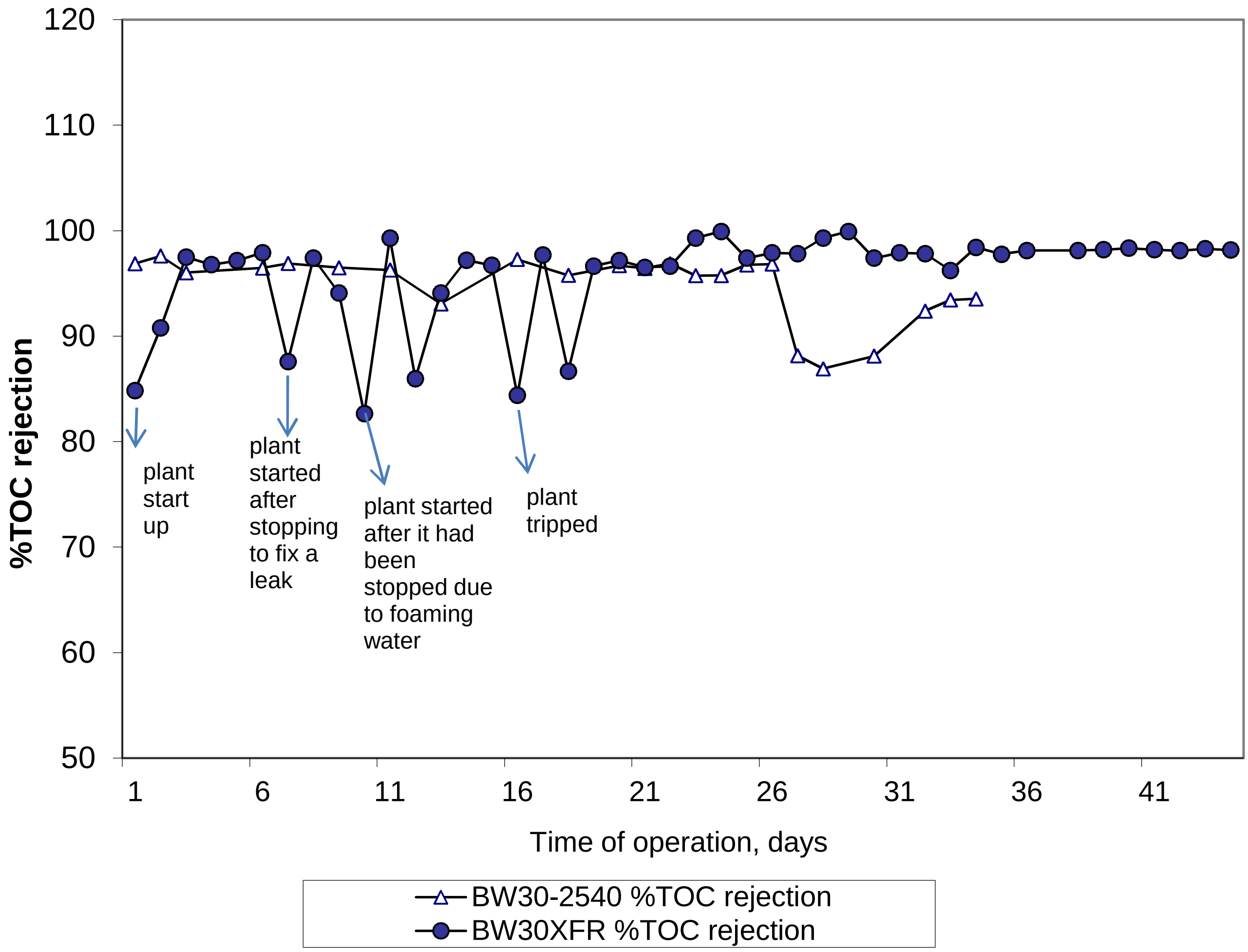


Figure 34: TOC rejection (%) of the BW30-2540 and the BW30XFR-2540 membranes during treatment of the raw concentrated raw water as a function of time

The TOC rejection varied between 86.9% and 97.61% and 82% and 99% for the BW30-2540 and the BW30XFR-2540 membranes, respectively. It was observed that lower rejections (84%, 87%, 82% and 84%, respectively) were obtained with the BW30XFR-2540 membrane during times of plant start/stops than with the standard BW30-2540 membrane as indicated in Figure 34. The maximum TOC rejection obtained when the plant operated with the BW30-2540 membrane was 97.61% (BW30XFR-2540 membrane's maximum TOC rejection was 99%).

4.6.1.3 The Effect of Biological Activity in the feed water on the RO membranes during treatment of the feed water

The bacterial analysis of the concentrated (evaporated) raw water and RO feed is shown in Table 10.

Table 10: Bacterial analysis of the concentrated raw water and the RO feed in the RO feed tank

Sampling Points	Total Coliform Bacteria CFU/100mL	Faecal Coliform CFU/100mL	Escherischia Coli CFU/100 mL	Heterotrophic Plate CFU/ mL
Concentrated raw water (no biocide addition)	Nil	Nil	–	90
RO feed (after caustic CIP and biocide dosing)	Nil	Nil	–	14

The heterotrophic plate count of the CRW was 90 counts per mL and that of the RO feed in the RO feed tank was 14 counts per mL. This indicated that the biocide dosing was partly successful in suppressing the biological activity. Although there were indications of increased biological activity during operation as indicated by an increase in TOC of the RO feed water (from an average of 35 mg/L to values between 60 and 90 mg/L) and the slimy feel of the piping and feed tank surfaces, the plate counts yielded by the analysis were not excessive. Therefore, a better measurement method for biological activity should be looked for. A Robbins device or a similar method that could give information regarding possible colonies forming on the surfaces should be investigated.

Although there was no evidence in the study to confirm the presence of slime on the pipe surfaces, literature (Flemming, 1997; Dreeszen, 2003), indicated that it would be possible to have biological activity in ultrapure water (18 MOhm) and that cells could adhere to surfaces within 30 seconds of exposure and biofilms could start to develop within two hours or less, depending on the conditions prevalent at the time.

4.6.2 RO treatment of the cooling water feed with UF Pre-treatment

Cooling water was filtered through the UF membranes and treated with the low-fouling RO membrane (BW30XFR-2540). The water was dosed with an anti-scalant (Genesys LF at a dosage of 5.84 mg/L) and a biocide (1.5% isothiazalone at a dosage of 6 mg/L) to control scaling and biofouling effects. The RO permeate was collected and treated further with the standard RO membrane (BW30-2540). The objective of the test was to determine the levels of TOC rejection that could be attained with a second pass RO configuration (similar to the configuration of the main plant) and, therefore, to establish what TOC levels could be expected in the feed to the CEDI, which is the final polisher on the main plant. The inlet requirement for the CEDI is a TOC of less than 500 µg/kg.

4.6.2.1 Normalised Permeate Flux for RO treatment of the cooling water with UF Pre-treatment

The normalised permeate flux for the RO treatment of CW with UF pre-treatment using a low-fouling membrane BW30XFR-2540 is shown in Figure 35. The operation with pre-treated water commenced on day 37 after the operation without pre-treatment had ended. The detailed results are shown in Appendix 14 – 14.1.

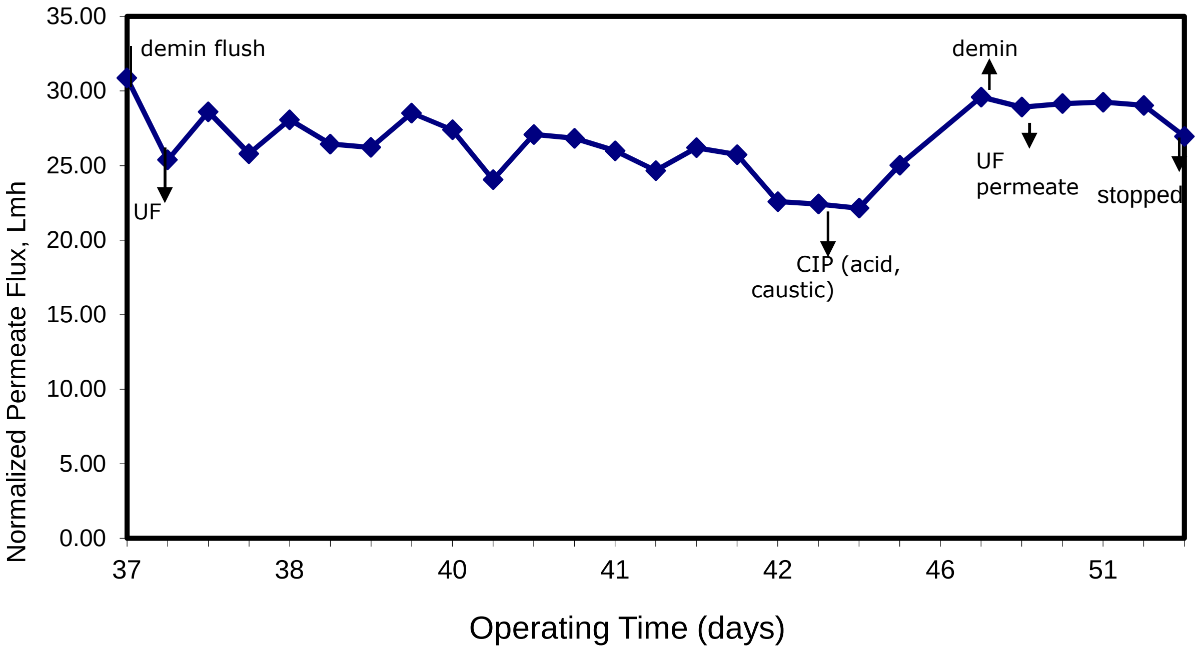


Figure 35: Normalised permeate flux as a function of time for the BW30XFR-2540 membrane with UF pre-treatment

(Note: graph is not to scale. Number of data points plotted varied from day to day of operation –see Appendix 14.1).

The initial CWF was 30.87 Lmh (day 37 when operation started). The flux then declined to 25.38 Lmh when the UF permeate (CW) was used as feed water, with the decline in permeate flux being associated with the increase in osmotic pressure of the feed water, and then increased to 28.61 Lmh (day 37). The increase in permeate flux was presumed to be due to the increase in the feed pressure (from 10 bar to 10.6 bar).The flux then stabilised at approximately 26.4 Lmh on the second day of operation (day 38 Figure 35) and remained relatively stable until it started declining on day 41; and thereafter declined to 22.58 Lmh (day 42 - 5 days of operation). Thereafter, A CIP was performed when the flux had declined to 22.58 Lmh (11.03% flux decline- from 25.38 Lmh to 22.58 Lmh). The CWF obtained after the CIP was 29.60 Lmh (day 48). This was slightly lower than the initial CWF of 30.87 Lmh obtained on day 37 when operation with pre-treated water was started. The permeate flux subsequently declined to 28.92 Lmh when the filtered CW was added on day 50 (13 days of operation). The flux remained stable at approximately 29.2 Lmh and started to decline on day 52 to 26.96 Lmh and the unit was then stopped (end of pilot plant operation). The membrane was preserved with a sodium metabisulphite (SMBS) solution prior to a membrane autopsy.

The flux decline of 11.03% prior to CIP was within the accepted industry norm of 15% flux decline prior to cleaning. The CWF regained after the CIP was lower than that of the CWF at the start of the operation (29.60 Lmh vs. 30.87 Lmh thus flux recovery of 95.89%). This was within the industry norm of a flux recovery of 95%, indicating that the CIP had been effective.

The normalised permeate flux for the RO treatment of the RO permeate as a function of time for the BW30-2540 standard membrane is shown in Figure 36. The detailed results are shown in Appendix 14 –14.2.

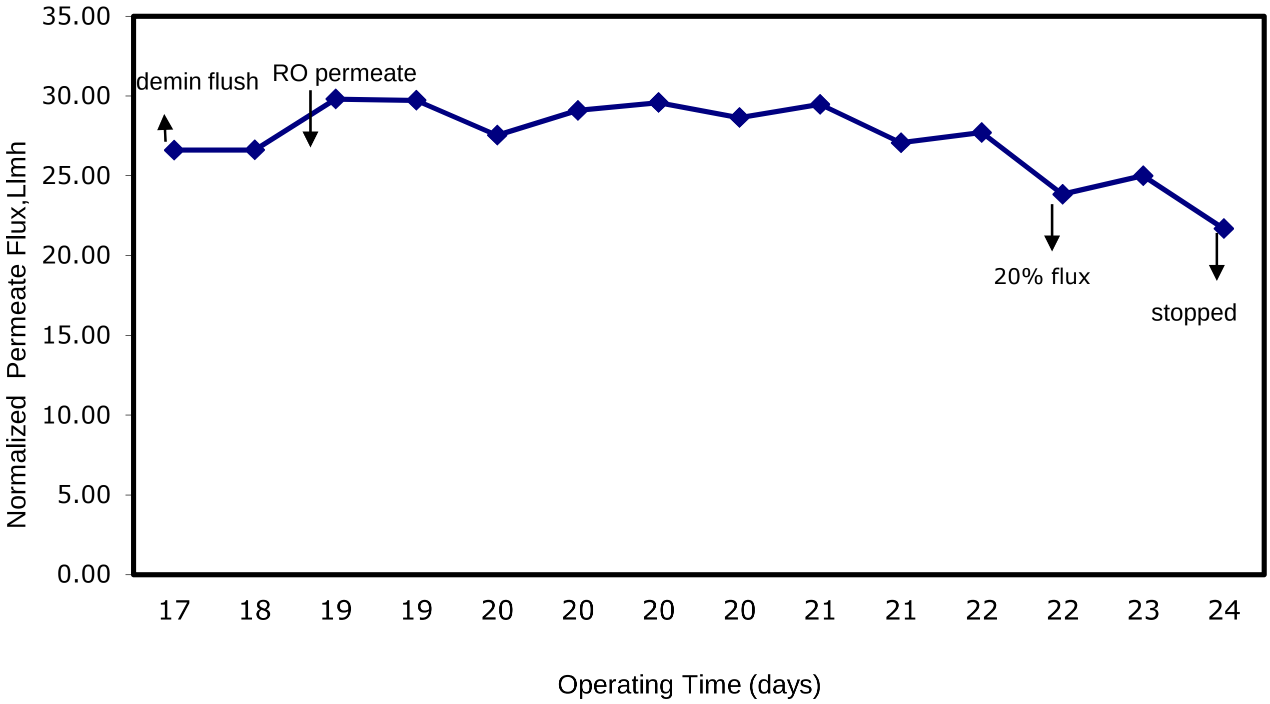


Figure 36: Normalised permeate flux for RO treatment of the RO permeate (from operation with the BW30XFR-2540 membrane) as a function of time for BW30-2540 membrane

(Note: graph is not to scale. Number of data points plotted varied from day to day of operation – see Appendix 14.2).

The initial CWF was 26.61 Lmh (day 18). The permeate flux increased to 29.73 Lmh when the RO permeate was used as feed water (day 19). The permeate flux remained at 28 Lmh (on average) after three days of operation. The flux eventually declined to 23.84 Lmh on day 22 (after 3 days of treatment of RO permeate). This represented a flux decline of 19.81% - from 29.73 Lmh to 23.84 Lmh. The unit was stopped when the flux declined to 21.69 Lmh on day 24 (end of operation of pilot plant). There was no CIP done prior to shut-down but the membrane was preserved with an SMBS solution prior to a membrane autopsy.

The normalized flux of the standard BW30-2540 membrane declined quicker than that of the low fouling membrane (3 days operation compared to 5 days operation achieved with the low fouling membrane prior to CIP). The permeate flux of the standard membrane stabilized at 28 Lmh when treating the RO permeate while the flux of the low fouling membrane stabilized at 27 Lmh when treating UF permeate. The CWF recovery obtained with the low fouling BW30XFR-2540 membrane was 95. 89%, this is within the accepted industry norm of 95%. This showed that the low fouling membrane had a slight advantage over the standard membrane in terms of operating flux as it could maintain a permeate flux which was comparable to that of the standard membrane (27 Lmh vs. 28 Lmh) while it treated a more challenging feed water. Although the advantage is slight, there is benefit in the use of the XFR membrane instead of the standard brackish water membrane as the price difference between the two membranes is not significant.

4.6.2.2 TOC Rejection of the BW30XFR-2540 and the BW30-2540 RO Membranes with Pre-treated Water

The permeate that was produced during the RO treatment of CW pre-treated with UF was collected and used as feed water to the standard BW30-2540 membrane. This was done to establish the capability of the membranes to reduce the TOC through a two-pass RO operation in order to be able to meet feed water requirements for the CEDI (final polisher). The TOC of the feed and permeate streams for the imitation of a first and second pass RO treatment is shown in Table 11.

Table 11: TOC of feed and permeate streams for the imitation of a first and second pass RO operation.

Feed TOC for First Pass RO Imitation (mg/L)	Permeate TOC for First Pass RO Imitation (mg/L)	Feed TOC for Second Pass RO Imitation (mg/L)	Permeate TOC for Second Pass RO Imitation (mg/L)	Treatment Objective (CEDI Inlet Requirement) (mg/L)
63.7	1.09	4.89	0.579	0.5
73.1	1.34	6.45	0.844	
Average TOC rejection on first pass : 98.23% and Average TOC rejection on second pass: 89.80%.		8.67	1.03	
		16	1.22	
		16	1.05	

The results (Table 11) show that the second pass RO (imitation) TOC obtained exceeded the treatment objective (0.5 mg/L TOC CEDI inlet requirement). This could be ascribed to biological activity in the feed water, which caused an increase in the TOC. Using a TOC rejection of 95% (BW30-2540 membrane) as obtained during the pilot plant operation, the second pass RO permeate was expected to show a TOC of 67 µg/kg (when feed TOC is 1.34 mg/L- see Table 11). However, much higher values of TOC were observed (4.89 mg/L to 16 mg/L – Table 11). This indicated that biological activity in the feed water should be monitored and controlled very strictly on the plant to prevent such occurrences.

It is critical that strict control of the cooling water chemistry is exercised at Komati Power Station, as the cooling water is very rich in microbes and, thus, there is a significant amount of biological activity in the cooling water circuit. This complicates the situation at Komati Power Station, as the biological activity tends to elevate the TOC concentration beyond its cycles of concentration (as the microbes die, they produce carbonaceous matter which is measured as TOC) and, consequently, the feed water to the desalination plant has high contents of organic matter as well as biological activity, which make operation of the desalination plant (UF and RO processes) challenging in terms of biofouling control and performance.

4.6.3 Reduction of organics in the Komati Power Station Desalination Plant (main plant) during the treatment of the raw water feed

UV scans were carried out on the UF feed, UF permeate, first pass RO (RO1) feed, first pass RO (RO1) permeate, first pass RO (RO1) brine, second pass RO (RO2) feed, second pass RO (RO2) permeate and second pass RO (RO2) brine to establish the effectiveness of the Komati plant process in the reduction/removal of DOC using UV absorbance from the wavelength 200 nm to 300 nm. Absorbance was measured from 220 nm to 300 nm at intervals of 10 nm.

The UV absorbance of the compounds that absorbed in the range of 220 nm to 300 nm for three different days of operation is shown in Figures 37 to 39. The UV absorbance of some of the streams was similar, therefore, there was overlap. The UV absorbance of the UF permeate overlapped with that of the RO1 feed sample while the RO1 permeate, RO2 feed, RO2 permeate and RO2 brine overlapped with one another. The detailed results are shown in Appendix 15.

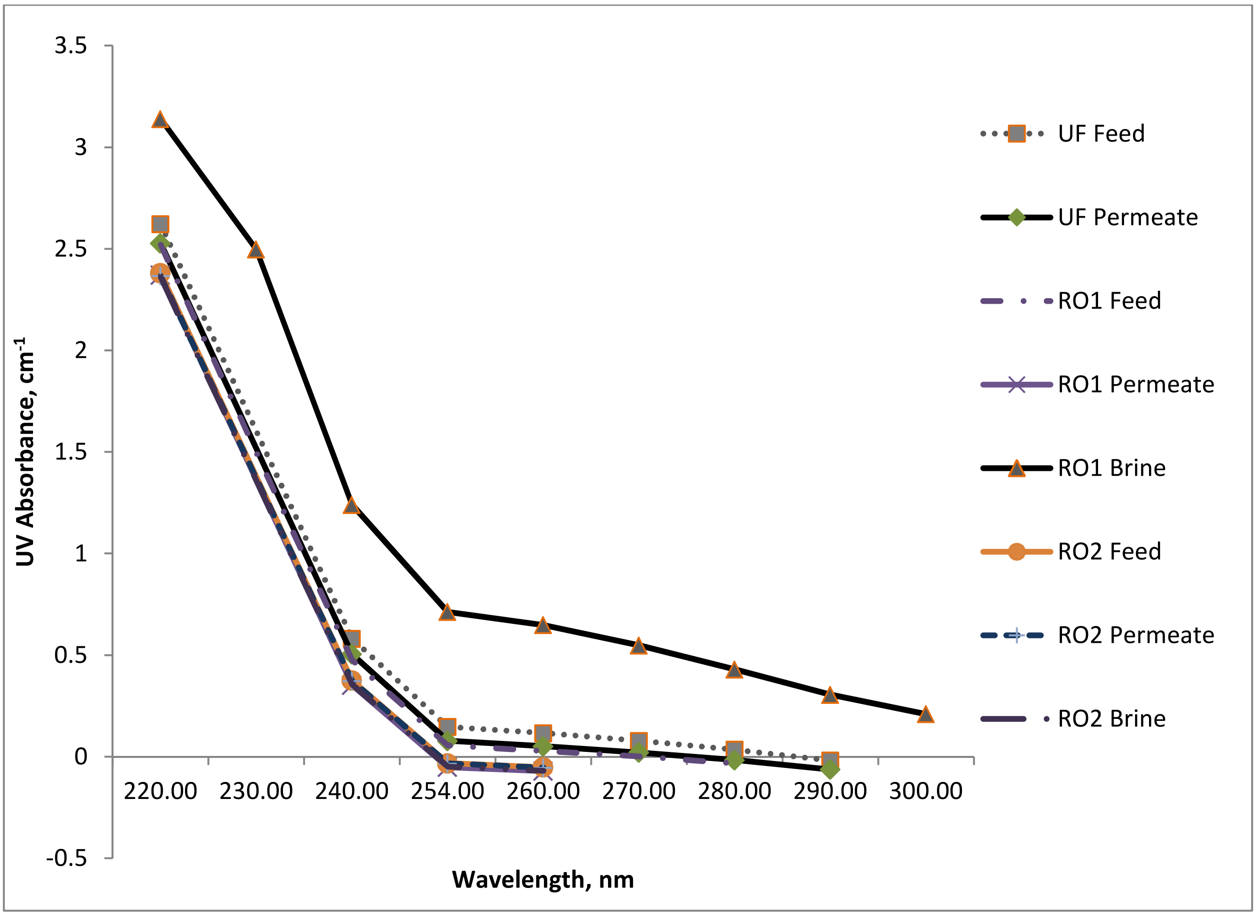


Figure 37: UV absorbance of UF and RO feed, permeate and brine for 31 March 2011

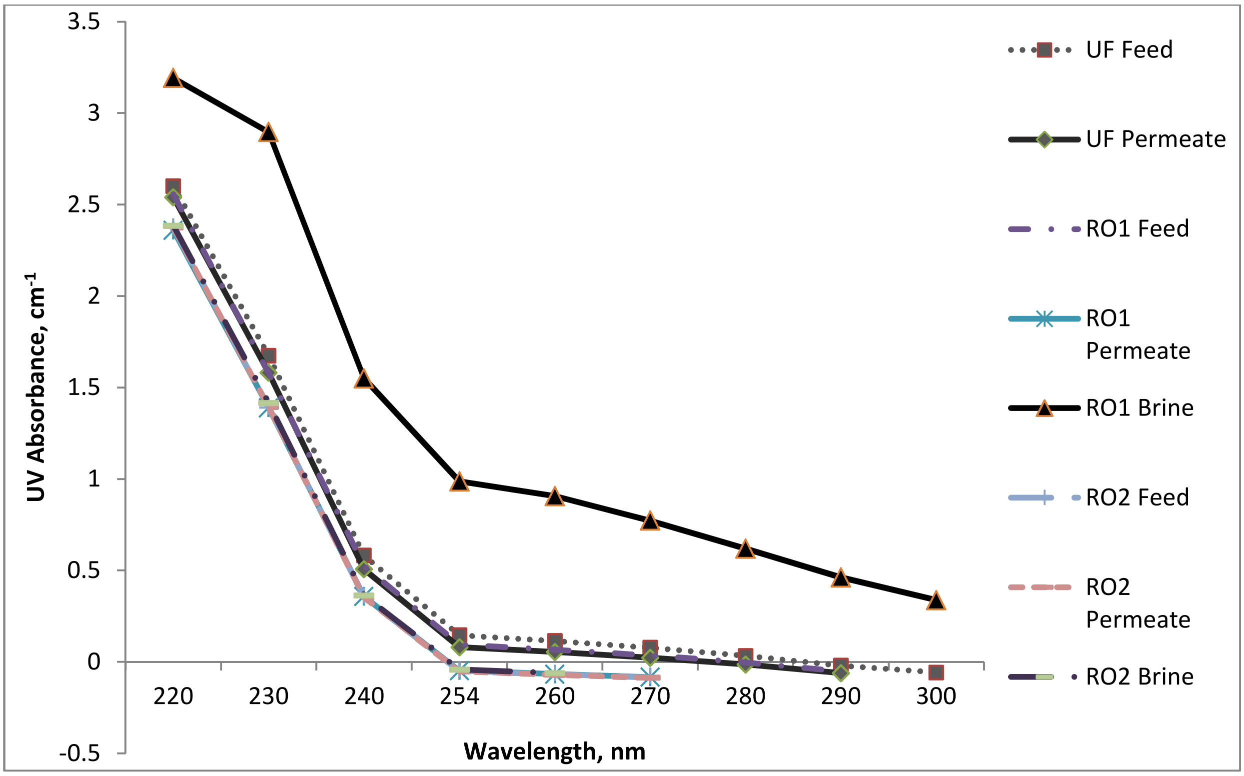


Figure 38: UV absorbance of UF and RO feed, permeate and brine for 5 April 2011

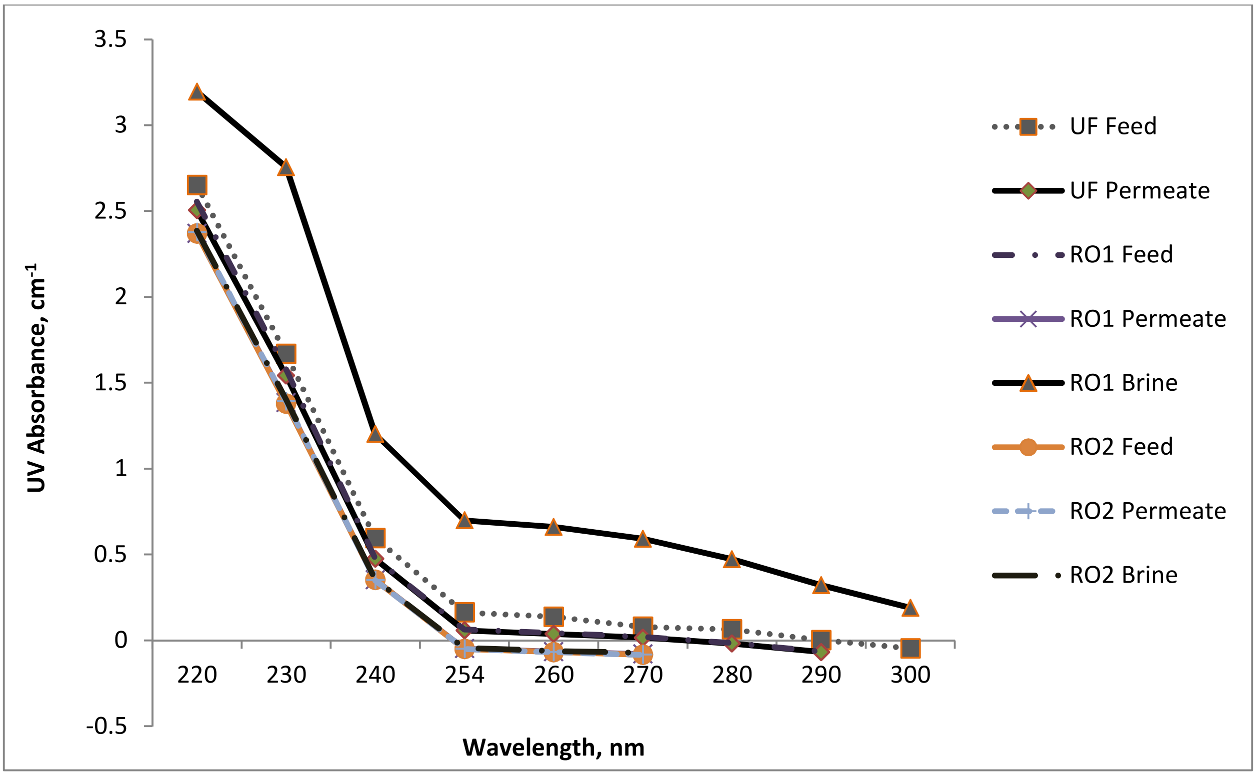


Figure 39: UV absorbance of UF and RO feed, permeate and brine for 12 April 2011

The UV absorbance of the RO1 brine stream was the highest over all the wavelengths tested, followed by the UF feed; UF permeate, RO1 feed and RO1 permeate (Figures 37 to Figure 39). This showed that the organics were removed by the UF membranes and the first pass (RO1) RO membranes. The mass balances carried out for the first pass RO show that the expected concentration of organics in the brine as measured by UV_{254} is comparable to the actual measured values as shown in Figures 37 to 39 (calculated values vs. measured values were 0.73 cm^{-1} vs. 0.71 cm^{-1}; 0.99 cm^{-1} vs. 0.98 cm^{-1} and 0.76 cm^{-1} vs. 0.70 cm^{-1}, respectively).

There were no organics measured by UV absorbance from the wavelength 254 nm to 300 nm in the second pass RO. The UV absorbing species present in the RO2 feed absorbed only in the wavelength 220 nm to 240 nm and there was little to no removal of these species with the second pass RO membrane. It is expected that when the plant is treating cooling water that the second pass RO membranes will be presented with a significant organic load as the organic loading of the cooling water is much higher than that of the raw water.

The TOC removal from the UF, RO1 and RO2 for the Komati desalination plant is shown in Table 12. The detailed results are shown in Appendix 15.

Table 12: TOC removal across the Komati desalination membrane plant

	31-Mar-11		05-Apr-11		12-Apr-11		
Process	TOC, mg/L	% TOC Removal	TOC, mg/L	% TOC Removal	TOC, mg/L	% TOC Removal	Average %TOC removal
UF feed	7.93	30.64	7.23	23.65	7.14	33.61	29.30
UF permeate	5.5		5.52		4.74		
RO1 feed	4.7	97.83	5.7	90.47	4.91	98.38	95.56
RO1 permeate	0.102		0.543		0.0794		
RO2 feed	0.13	61.08	0.847	83.23	0.0942	14.01**	52.77
RO2	0.0506		0.142		0.081		

Note: ** data point discarded

The average TOC removal with the UF was 29.3%, while the average TOC removal for RO1 and RO2 was 95.56% and 72.16%, respectively. The removal of TOC from RO2 varied. This implies that operation of the second pass RO (RO2) is not optimal and performance should be improved to match that of the first pass. Operational conditions that could be optimized for improved TOC rejection include flux, solution chemistry (e.g.

pH- NOM rejection is low at pH 4.5) and the applied (feed) pressure. Other factors that affect the rejection of organics includes charge of the membrane (a negatively charged membrane surface will repel negatively charged organic molecules), charge of the organic molecule and molecular size (large molecules are easily rejected) (Hong & Elimelech, 1997).

4.6.4 Water Quality

The RO feed and permeate water qualities during the pilot plant operation are shown in Table 13 (BW30-2540 and BW30XFR-2540).

Table 13: RO feed and permeate qualities during the pilot plant tests (BW30-2540 and BW30XFR-2540)

		Standard Membrane BW30-2540			**Low Fouling Membrane BW30XFR-2540**			**Simulated RO 2 on BW30-2540**		
Component	Unit	RO Feed 25/26- 02-2010	RO Permeate 21/26-02-2010	**% Rejection**	Feed 24/26-02/2010	Permeate 20/26-02-2010	**% Rejection**	RO 1 Permeate 22/26-02-2010	RO 2 Permeate 23/26-02/2010	**% Rejection**
Alkalinity total	mg/L CaCO3	266	5.9	**97.78**	130.00	4.20	**96.77**	31.2	16.6	**46.79**
Al	mg/L	0.04	0.005	**87.50**	<0.005	<0.005		0.04	0.005	**87.50**
Ba	mg/L	0.1	0.005	**95.00**	0.25	0.005	**98.00**	0.03	0.005	**83.33**
Be	mg/L	<0.005	0.005		<0.005	<0.005		<0.005	<0.005	
Br	mg/L	0	0		0.80	0.00	**100.00**	0	0	
B	mg/L	0.7	0.2	**71.43**	0.30	0.06	**80.00**	0.2	0.06	**70.00**
Ca	mg/L	39	0.005	**99.99**	200.00	0.005	**100.00**	6.8	0.26	**96.18**
Cd	mg/L	<0.005	<0.005		<0.005	<0.005		<0.005	<0.005	
Cl	mg/L	125	3.58	**97.14**	290.00	0.60	**99.79**	14.13	2.27	**83.93**
Conductivity	mS/cm	404	2.3	**99.43**	290.00	1.00	**99.66**	15.1	3.5	**76.82**
Co	mg/L	<0.005	<0.005		<0.005	<0.005		<0.005	<0.005	
Cr	mg/L	0.01	<0.005		<0.005	<0.005		<0.005	<0.005	
Cu	mg/L	0.03	0.02	**33.33**	0.18	0.02	**88.89**	0.02	0.02	**0.00**
DOC	mg/L	112	2.36	**97.89**	105.00	1.29	**98.77**	15.6	4.1	**73.72**

Fe	mg/L	0.11	0.02	**81.82**	0.05	0.01	**80.00**	0.03	0.02	**33.33**
F	mg/L	7.4	0.05	**99.32**	2.51	0.03	**98.80**	0.62	0.05	**91.94**
K	mg/L	23	0.49	**97.87**	38.00	0.54	**98.58**	0.68	0.84	**-23.53**
Mg	mg/L	66	0.005	**99.99**	150.00	0.005	**100.00**	8.4	0.005	**99.94**
Mn	mg/L	<0.005	<0.005		0.04	0.005	**87.50**	<0.005	<0.005	
Na	mg/L	690	4.5	**99.35**	260.00	2.50	**99.04**	8.6	7.1	**17.44**
Ni	mg/L	0.03	0.01	**66.67**	0.03	0.01	**66.67**	0.01	0.01	**0.00**
NO2	mg/L	<0.020	<0.020		1.93	0.02	**98.96**	<0.020	<0.020	
Nitrate as N	mg/L	18.43	0.29	**98.43**	33.90	0.31	**99.09**	4.64	0.05	**98.92**
OA as O2	mg/L	94	2.7	**97.13**	150.00	2.60	**98.27**	18	1.9	**89.44**
Pb	mg/L	0.02	0.02	**0.00**	0.02	0.02	**0.00**	0.02	0.02	**0.00**
pH at 25 ℃		8.34	7.23		8.27	7.75		7.9	7.89	
Reactive silica as SiO2	mg/L	5.99	0.2	**96.66**	4.10	0.01	**99.76**	0.2	0.08	**60.00**
SO4	mg/L	150	0.02	**99.99**	1190.00	0.76	**99.94**	1.37	0.43	**68.61**
Sr	mg/L	0.32	0.005	**98.44**	1.40	0.005	**99.64**	0.04	0.005	**87.50**
TOC	mg/L	108	2.61	**97.58**	112.00	1.86	**98.34**	15.4	4.65	**69.81**
UV absorbance at 254nm		2.134	N/D		4.12	0.13	**96.77**	0.534	0.096	**82.02**
Zn	mg/L	0.13	<0.005		0.64	0.005	**99.22**	<0.005	<0.005	

The low fouling membrane BW30XFR-2540 could effectively remove organics from the RO feed water as measured by the DOC rejection (98.77%), TOC rejection (98.34%), and oxygen absorbed (OA) (98.27%). A salt rejection (in terms of conductivity) of 99.66% was obtained (Table 13). The standard membrane BW30-2540 also demonstrated a good capability to remove organics from the RO feed water as measured by the DOC rejection (97.89%), TOC rejection (97.58%), and OA rejection (97.13%). A salt rejection (in terms of conductivity) of 99.43% was obtained. This showed that the performance of the two RO membranes in terms of organics rejection and salt rejection was comparable.

When a double pass RO process was imitated/mimicked by passing RO permeate produced by the low fouling membrane BW30XFR-2540 as feed to the standard membrane BW30-2540, the performance of the standard membrane was poorer than that observed when the membrane had been used for the treatment of concentrated raw and cooling water. The effectiveness of organics removal from the RO permeate as measured by DOC, TOC, OA and UV_{254} was 73.7%, 69.81%, 89.44% and 82.02%, respectively. A salt rejection (in terms of conductivity) of 76.82% was obtained. These are preliminary results for a double pass RO operation (mimicked/imitated) and further investigations should be conducted to establish reasons for the poor performance of the membrane and for possible optimization of the process to improve the performance.

4.7 ADDITIONAL PRE-TREATMENT OF THE COOLING WATER FOR OPTIMISATION FOR DOC REMOVAL

This section details the results of adsorption tests that were conducted in an effort to optimise the Komati process by further reducing the organic levels in the RO feed water in order to minimise fouling of the RO membranes.

Coagulation of the feed water improved the removal of the organics as measured by DOC from 4% (without coagulation) to approximately 35% with in-line coagulation. The organics that are not retained by the UF membrane are going through to the RO membranes (TOC >55 mg/L in UF permeate). The TOC requirement for the feed for the standard RO membrane is <6 mg/L and <10 mg/L for the low fouling RO membrane. Therefore, it was necessary to investigate other TOC removal processes that could reduce the TOC loading on the RO membranes to acceptable levels. Pilot testing of an organic scavenger and of activated carbon was conducted for the reduction of TOC in the cooling water. Samples of Chemviron activated carbon (F300 and F400) provided by Cwenga Technologies and resin samples from Rohm and Haas were tested. These were preliminary tests that were conducted to assess whether adsorption could be considered further in the identification of technologies for the reduction of TOC upstream of the RO membranes.

4.7.1 TOC Adsorption on Granular Activated Carbon

A pilot test was conducted to determine the removal of TOC from CW using GAC. GAC Filtrasorb F300 and F400 activated carbons were used. The breakthrough curves for the Filtrasorb F300 and F400 GACs are shown in Figures 40. The detailed results from which the breakthrough curves were derived are shown in Appendix 16.

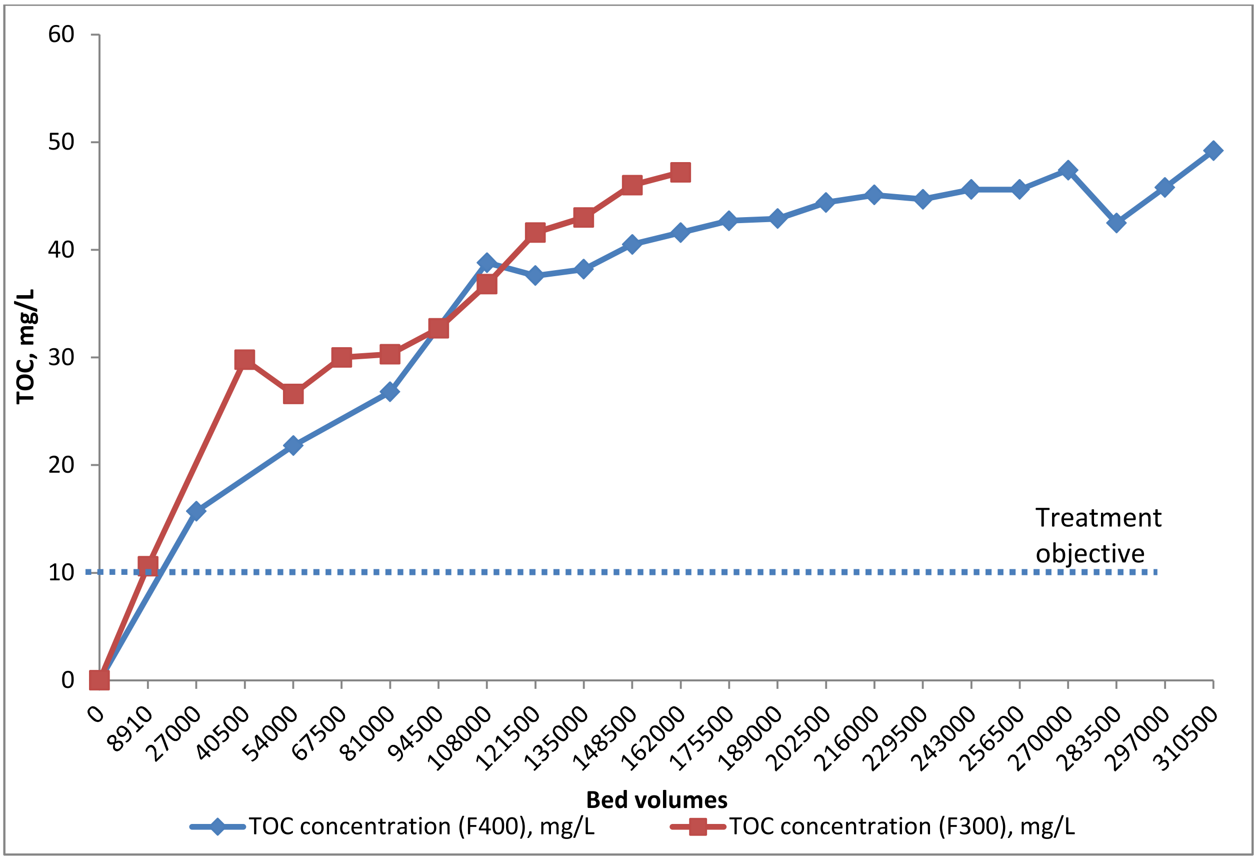


Figure 40: Breakthrough curve for F300 and F400 GAC pilot test – expressed as a function of BV,

1BV = 500 mL

Breakthrough at 10 mg/L TOC took place at 8910 bed volumes for the Filtrasorb F300. The breakthrough for the Filtrasorb 400 was already exceeded when the first sample was taken at 27000 BV, although it can be seen from Figure 40 that breakthrough was in-between 8910 and 27000BV, and is estimated at 18000BV, based on Figure 40. Although the Filtrasorb 400 had a better run length compared to the Filtrasorb 300 (18000BV vs. 8910BV), both run lengths were short. It was noted that batch jar tests should have been done as these would have helped to establish the adsorptive capacity of the activated carbon for the NOM and the breakthrough point, however, these are preliminary results and would need to be further investigated in order to confidently conclude on the economical viability of using activated carbon for the removal of organics at Komati as well as to identify other alternative adsorbents that can effectively reduce the organics. TOC removals obtained at 8910BV (F300) and 27000BV (F400) were 78.01% and 69.81%, respectively.

The use of biological activated carbon was not part of this investigation. However, it is possible that biological activated carbon might yield very different results. Bancrof *et al*. (1983) reported that bacterial TOC removal had been studied as a mechanism by which more adsorption sites on GAC could be made available for the removal of non-biodegradable compounds: thus, extending the bed life of GAC columns and reducing operating costs. Bancroft *et al.* (1983) found that 24% of TOC removal in the feedwater could be attributed to biological activity in the activated carbon bed. Babi *et al*. (2007) have reported that small portions of chlorination by-products (haloacetic acids) as well as DOC were removed due to biodegradation in the activated carbon adsorber, where dechlorination was enhanced by microbiological growth.

If biological activity does result in an extended bed life, then the economics of the biological activated carbon process might be more viable than a conventional GAC process for reducing organics.

4.7.2 TOC Adsorption on an Organic Scavenger

An acrylic, porous, anionic resin – Amberlite IRA958 Cl – was used to investigate the removal of the organics with the organic scavenger. The properties of the organic scavenger resin are shown in Appendix 17. The breakthrough curve of the organic scavenger is shown in Figure 41. The detailed results used to derive the breakthrough curve are shown in Appendix 16.

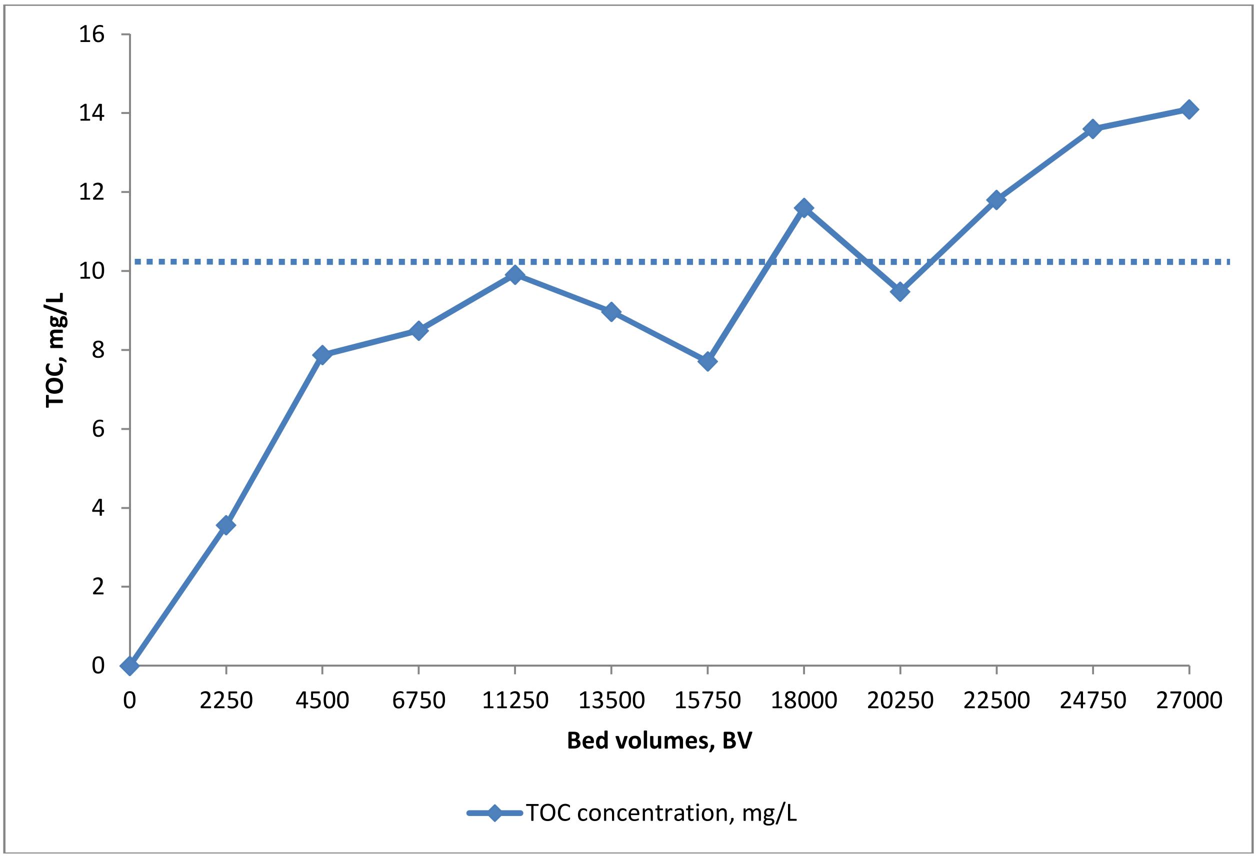


Figure 411: Breakthrough curve of the organic scavenger resin- expressed as a function of BV, 1BV = 1000 mL

A very short run length was obtained. The treatment objective of 10 mg/L was exceeded (from the second column) in less than 18000 bed volumes.

The short run length that was obtained was not surprising, as the cooling water was high in sulphates (concentration of sulphates in the CW>2 000 mg/L). According to Meyers (1995) and DeSilva (1997), the sulphate competes with organics for exchange sites on the resin and the relative selectivity of the anion resin for sulphates and organics are similar. Only a fraction of a complex organic acid has a selectivity coefficient higher than that of sulphate (Meyers, 1995). Thus, as the resin exhausts in the sulphate form, organics with selectivity less than that of sulphate are displaced from the resin (organic leakage). The quick TOC breakthrough (< 18000BV) made the organic scavenger unfeasible as a solution for Komati Power Station. In addition to the added operating cost, the organic scavenger represented an additional effluent stream in the form of the regenerant effluent. This is not in line with the Power Station’s objective of ZLED to minimise regenerant effluents that pose an additional load on the ashing system.

4.8 RO MEMBRANES AUTOPSIES

Autopsies (microscopic examination, EDX analysis, an ATR scan and microbiological analysis) were conducted on the RO membranes to identify the presence of membrane foulants.

4.8.1 Identification of Membrane Foulants

4.8.1.1 Microscopic Examination of the RO Membranes

Microscopic examination (at 100x magnification) of the active surfaces of the standard BW30-2540 and the low fouling BW30XFR-2540 membranes was carried out and this showed the presence of small, round particles, as shown in Figure 42. The feed and middle sections of the membrane are shown.

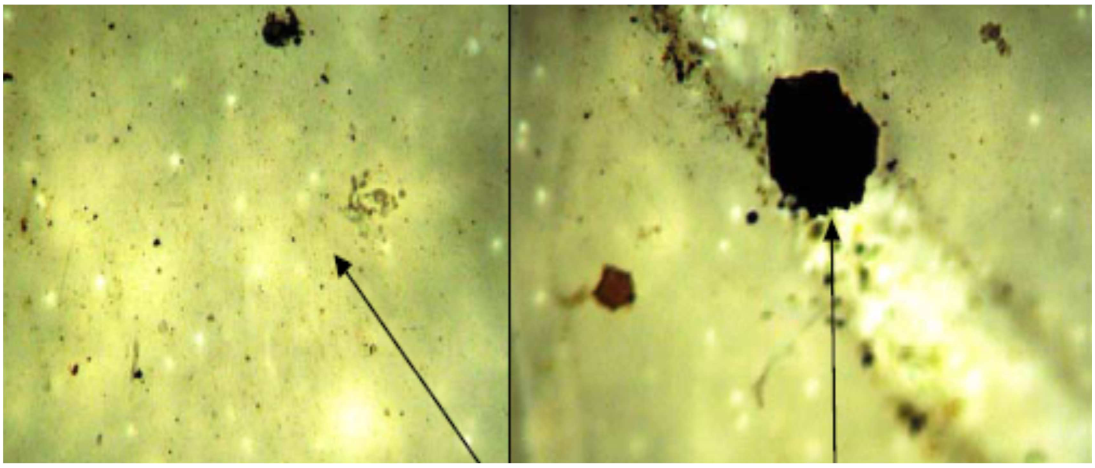


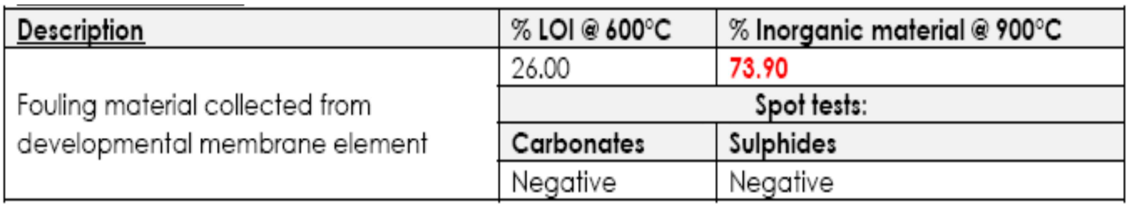
Figure 42: Particles on the active surface of the BW30XFR-2540 membrane

The deposit (Figure 42) appeared to be crystalline in nature, which could indicate that it was of inorganic nature, as organic foulants usually appear as amorphous rather than crystalline.

4.8.1.2 Spot tests, Ashing and EDX Analysis

The results of spot and ashing tests when the deposit on the membrane was ashed at 600 ℃ and 900 ℃ are shown in Table 14.

Table 14: Spot tests and ashing results for the BW30XFR-2540 membrane (Tigeli & Nonti, 2010)

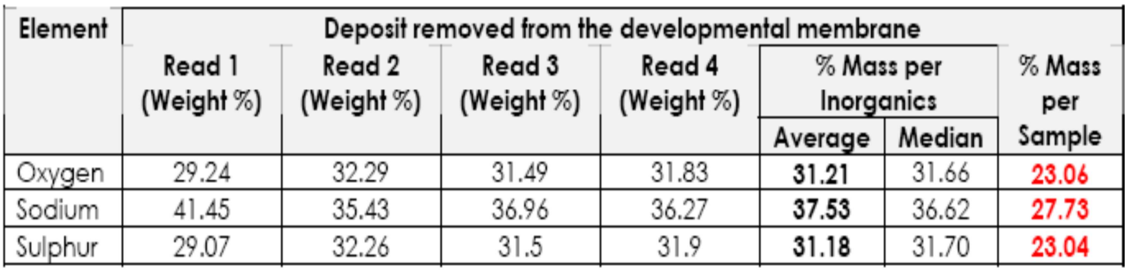


Description	% LOI @ 600°C	% Inorganic material @ 900°C
Fouling material collected from developmental membrane element	26.00	73.90
	Spot tests:	
	Carbonates	**Sulphides**
	Negative	Negative

The Loss-On-Ignition (LOI) results showed that when the sample was heated at 600 ℃, that there was a 26% loss of material indicating the presence of organics, while the ashing results showed further that the deposit collected from the low fouling BW30XFR-2540 membrane was mostly inorganic in nature (73.90%). Spot tests were negative for both carbonates and sulphides.

The EDX results of the deposit are shown in Table 15.

Table 15: EDX results of the deposit collected from the BW30XFR-2540 membrane



Element	Deposit removed from the developmental membrane						
	Read 1 (Weight %)	Read 2 (Weight %)	Read 3 (Weight %)	Read 4 (Weight %)	% Mass per Inorganics		% Mass per Sample
					Average	Median	
Oxygen	29.24	32.29	31.49	31.83	**31.21**	31.66	**23.06**
Sodium	41.45	35.43	36.96	36.27	**37.53**	36.62	**27.73**
Sulphur	29.07	32.26	31.5	31.9	**31.18**	31.70	**23.04**

The results showed that there was a presence of sodium and sulphur in the deposit. In addition to the EDX results, an attenuated total reflection (ATR) scan of the deposit was performed and this showed a sulphate peak at 1 117 cm^{-1} (see Tigeli & Nonti, 2010). This seemed to suggest that the deposit collected from the membrane could be sodium sulphate. It was surprising that the deposit collected from the surface of the membrane could be sodium sulphate, as this is a very soluble salt and was not expected to precipitate out of solution.

It was therefore argued that the deposit could arise from air oxidation of sodium metabisulphite (SMBS), which was used to preserve the membranes before they were transported to the external laboratory that was used for the autopsies. Solutions of SMBS exposed to air can be oxidised to sodium sulphate (Lee, 2009:559-560). The reaction of SMBS in water is:

$$Na_2S_2O_5 + H_2O \rightarrow 2NaHSO_3$$

Equation 1 Reaction of SMBS in water

In aqueous solutions the sodium bisulphite formed in equation 1 can oxidize readily when exposed to air as indicated by the reaction:

$$2NaHSO_3 + O_2 \rightarrow Na_2SO_4 + H_2SO_4$$

Equation 2 Air oxidation of SMBS

If sodium sulphate crystallises from an aqueous solution at room temperature or below, it does so as a heptahydrate ($Na_2SO4.7H_2O$) which is not very soluble. Heptahydrate crystals also oxidise in air to form the sulphate. A similar white crystalline deposit was observed on leaking valves vessels at the Komati MWTP on UF and RO vessels which were preserved with 1% SMBS. Analysis of the deposit using Quantitative Evaluation of Minerals by Scanning Electron Microscopy (QEMSCAN) mineral identification technique confirmed the presence of sodium sulphate (94.1%) as well as sodium sulphate with calcium impurities and sodium sulphate with Al and Si impurities. Trace proportions of sodium sulphate with chloride, iron oxide and pure sulphur were also present (van Alphen, 2012). QEMSCAN is an automated scanning electron microscope configured to automatically determine the mineralogical composition of any particulate sample (van Alphen, 2012).

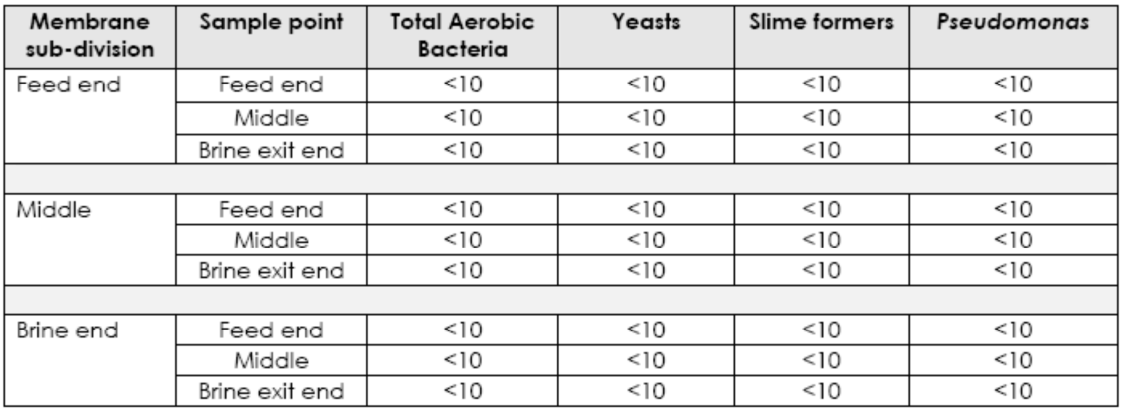
4.8.1.3 Microbiological Analysis of the Foulant

Microbial swaps results of the membranes are shown in Tables 16 (for BW30XFR-2540 membrane) and 17 (for BW30-2540 membrane).

Table 16: Plate counts for the low-fouling membrane - results expressed as CFU/50 cm^2

Membrane sub-division	Sample point	Total Aerobic Bacteria	Yeasts	Slime formers	*Pseudomonas*
Feed end	Feed end	<10	<10	<10	<10
	Middle	<10	<10	<10	<10
	Brine exit end	<10	<10	<10	<10
Middle	Feed end	<10	<10	<10	<10
	Middle	<10	<10	<10	<10
	Brine exit end	<10	<10	<10	<10
Brine end	Feed end	<10	<10	<10	<10
	Middle	<10	<10	<10	<10
	Brine exit end	<10	<10	<10	<10

Table 17: Plate counts for standard membrane - results expressed as CFU/50 cm^2



Membrane sub-division	Sample point	Total Aerobic Bacteria	Yeasts	Slime formers	*Pseudomonas*
Feed end	Feed end	<10	<10	<10	<10
	Middle	<10	<10	<10	<10
	Brine exit end	<10	<10	<10	<10
Middle	Feed end	<10	<10	<10	<10
	Middle	<10	<10	<10	<10
	Brine exit end	<10	<10	<10	<10
Brine end	Feed end	<10	<10	<10	<10
	Middle	<10	<10	<10	<10
	Brine exit end	<10	<10	<10	<10

The results in Tables 16 and 17 showed that there were no significant amounts of yeasts, aerobic bacteria, slime formers and *Pseudomonas* on the surfaces of either the standard BW30-2540 membrane or the fouling-resistant BW30XFR-2540 membrane. This was attributed to the fact that the membranes had undergone a CIP shortly before preservation and this would have removed foulants from the surfaces and also gave an indication of the success of the biocide dosing that was implemented for the control of biological activity.

The results of the autopsies did not reveal the presence of an organic foulant, as was suspected during the pilot test. However, the performance of the plant (in terms of flux decline and the increase in TOC when control of bacterial activity was not practised) indicated that the membranes were fouled. Scaling was not expected to have been responsible for the flux decline as anti-scalant was dosed in the RO feed water. Therefore, biofouling could have been responsible for the fouling of the membranes and this matter needs further investigation.

CHAPTER 5:
CONCLUSION AND RECOMMENDATIONS

The main objectives of the study were to:

- Investigate the impact of natural organic matter in the Komati Power Station raw and cooling water on membrane fouling;
- Optimise the UF and RO membrane processes to minimise fouling of the membranes.

The following conclusions can be made as a result of the investigations.

5.1 WATER QUALITY

- The cooling water quality was not determined prior to the design of the desalination plant at Komati, and this has resulted in a design base that was not suitable for the cooling water conditions on site, causing the desalination plant to be under-designed with respect to organics and oil in the cooling water.
- It was found that the raw water and, consequently, the cooling water at Komati Power Station had a high organic content, on average a TOC of 6.5 mg/L, and this peaked to values in excess of 8.5 mg/L TOC for the raw water and approximately 80 mg/L TOC for the cooling water.

5.2 WATER CHARACTERISTICS

- The DOC characterisation of the raw, concentrated raw and cooling water indicated a high content of humics (63%, 64% and 67.9%, respectively). The biopolymers (polysaccharides, carbohydrates, amino-sugars) in the raw, concentrated raw and cooling water only made up a small fraction of the DOC make-up (4.3%, 2.6% and 0.9%, respectively).
- SUVA measurements indicated that the DOC in the raw, concentrated raw and cooling water were aromatic in character (SUVA values of between 4 and 5 L/(mg*m)

correspond to DOC of the water consisting of mostly humic substances and the DOC being relatively hydrophobic and aromatic).

- The TOC:DOC ratio (≈1) indicated that the raw feed water consisted mostly of DOC and little to no POC.

5.3 Jar tests (Clarification)

- MembraQuest 3000a coagulant gave the highest turbidity removal (95.05% at a dosage of 5 mg/L) but gave the poorest DOC removal at dosages of 5 to 30 mg/L.
- Ferric chloride and alum gave the highest DOC removal (48% at a dosage of 30 mg/L ferric chloride and 47.69% at a dosage of 10 mg/L alum). Ferric chloride gave poor turbidity removals (83.55% at 30 mg/L) while alum gave turbidity removal of 91.97% at a dosage of 10 mg/L.
- Ferric chloride and alum were not selected for use as coagulants in the pilot plant due to their impact on the plant operations. These coagulants depressed the pH and would necessitate an additional process step for pH adjustment on the main plant to maintain a pH of 7.8 and they also resulted in sulphate and chloride residuals, which would have imposed increased loading of chlorides and sulphates on the desalination plant and that would require removal.
  - U3000 was selected as the coagulant of choice to be used for the in-line coagulation of the raw water. This coagulant did not affect the pH of the raw water and achieved turbidity removals of 91.62% and DOC removals of 35.39% at a dosage of 20 mg/L.

5.4 IMPACT OF ORGANICS IN THE FEED WATER ON THE UF MEMBRANE

- UF treatment of the raw feed water gave very poor DOC and turbidity removal (4% and 13%, respectively) when operated without coagulation of the feed water with a slow decline in flux. The highest flux decline (15.53%) was experienced in the first 64 hours of operation after which flux decline averaged 3.46%.
- The organics in the feed did not have a very significant fouling impact (in terms of flux decline) on the UF membrane when operated without coagulation. More than 90% of the DOC consisted of LMW fractions which were less than the MWCO of the membrane and passed through the membrane. The particulate matter in the feed was

mostly responsible for the decline in flux (and hence fouling) of the UF membrane that was experienced.

- In-line coagulation of the raw feed water resulted in improved organics removal by the UF membrane. Maximum DOC removal was 30.13% and organics removal as measured by UV254 was 50.33% (with no coagulant mixing).
- Introduction of coagulant mixing led to an increase in DOC (removal from 30.13% to 42.58%) and organics removal (48.53%) as measured by UV_{254}.
- The flux of the UF membrane could be restored with regular backwash and CEB procedures.
- Although coagulation increased the size of the pin floc, most of the DOC still passed through the UF membrane due to the small molecular size of the DOC. The UF permeate still contained TOC concentrations > 55 mg/L after in-line coagulation.
- The performance of the UF membrane (with in-line coagulation) was not sufficient to reduce the DOC levels to 6 mg/L for the standard RO membrane and 10 mg/L for the low fouling membrane which were required for the RO membranes.
- The UF membrane did not protect the RO membranes from DOC fouling because the UF itself did not remove much of the DOC as a result of the low molecular weight DOC fractions in the Komati water.

5.5 IMPACT OF ORGANICS IN THE CONCENTRATED RAW AND COOLING WATER FEED ON THE RO MEMBRANES

- The operation of the RO using pre-treated water (UF) improved the performance of the RO process in terms of the rate of membrane fouling. The flux of the standard BW30-2540 membrane declined over three days from 29.73 Lmh to 23.84 Lmh when using pre-treated water compared to a decline over two days of operation without pre-treated water. The flux of the low-fouling BW30XFR-2540 membrane declined over five days from 25.38 Lmh to 22.58 Lmh when using pre-treated water compared to four days when operated without pre-treated water.
- Fouling of the RO membranes could be ascribed to the high organic loading (>55 mg/L) on the RO membranes as the UF membrane was not successful in reducing the organic loading sufficiently to protect the RO membranes.

- The low fouling membrane had an advantage over the standard membrane in terms of flux decline (it took longer for the low fouling membrane to foul – 5 days compared to 3 days for the standard membrane when using pre-treated feed water).
- The low fouling membrane achieved better CWF recovery after the cleaning cycles (81.26% Lmh) compared to the standard membrane (77.35%) when using untreated feed water. This performance improved when pre-treated feed water was used and the low fouling membrane's CWF regained after the CIP (95.89%) was within the industry norm of a flux recovery of 95%, indicating that the CIP had been effective.
- The low fouling membrane had a slight advantage over the standard membrane in terms of operating flux as it could maintain a permeate flux which was comparable to that of the standard membrane (27 Lmh vs. 28 Lmh) while it treated a more challenging feed water.
- The DOC rejection of the low-fouling membrane was higher (average TOC rejection of 97%, maximum TOC rejection of 99%) than that of the standard membrane (average TOC rejection of 95.3%, maximum TOC rejection of 97%).
- The low fouling membrane demonstrated a better capability to treat the Komati raw and cooling water and would be expected to achieve lower operating costs for the plant (CIP costs and membrane replacement costs) while achieving better organics removal.
- Organics removal as measured by UV absorbing species from wavelengths 220 nm to 300 nm on the Komati Power Station desalination plant showed that organics were removed by the UF and RO1 process steps (the UV absorbance of the streams decreased in the order of UF feed; UF permeate, RO1 feed and RO1 permeate, indicating removal of organic species through these treatment processes). There was no organics loading on the second pass RO (RO2) from the wavelength of 254 nm to 300 nm.
- Organics removal from the raw water feed to the Komati Power Station desalination plant as measured by TOC rejection was 29.30% (average) for the UF, 95.66% for RO1 and 72.16% for RO2 processes.
- The results of the preliminary testing of the organic scavenger resin and granular activated carbon showed that these adsorbents might not be suitable for the reduction of DOC, due to the relatively quick TOC breakthrough (8910BV, approximately 18000BV and less than 18000BV for the Filtrasorb 300, Filtrasorb 400 and organic

scavenger resin) that would translate to increased regeneration/reactivation requirements as well as increased effluent generation on site.

- Membrane autopsies identified an inorganic foulant on the RO membranes proposed to be sodium sulphate. Sodium sulphate, although a soluble salt, could have resulted from the air oxidation of the SMBS used as a preservation solution, which could have crystallized in the relatively insoluble heptahydrate form at room temperature on the membrane.

5.6 RECOMMENDATIONS

- RO membrane fouling was observed despite pre-treatment of the feed water. Therefore, improved pre-treatment methods should be investigated. This can include the use of better coagulants, use of enhanced coagulation instead of in-line coagulation, advanced oxidation processes, etc.
- Coagulant mixing was not effective. Therefore, the use of a proper coagulant mixing device (such as a static mixer) would improve the in-line coagulation process and should be further invesigated.
- Better coagulant retention time is required to improve organics removal by coagulation. This would require that sand filters be retrofitted to ensure there is no carry-over of flocs to the UF membranes.
- The high TOC concentrations (3.5 - 8.5 mg/L) in the raw feed can aggravate biological fouling and it is essential that a proper regime is put in place for monitoring biological activity in the system. Biofouling and impacts thereof on plant operations and performance should be investigated further.
- An investigation should be carried out on cleaning solutions (CIP agents) to identify the most effective and efficient CIP chemicals that should be used.
- It is suggested that the Low-fouling membranes (XFR) should be installed in the first RO pass on the main plant while the second pass of the RO should use the standard membranes because the permeate from the first pass will have low levels of organics (97% average DOC rejection is obtained on the XFR membrane).
- It is suggested that further work be conducted on the use of GAC and organic scavenger resin for the removal of TOC to confirm the preliminary findings.

- It is also suggested that other DOC-reducing processes, such as biological activated carbon and Magnetic Ion-exchange (MIEX) resin, use of tighter UF or NF membranes, etc. are investigated as options for DOC removal upstream of the RO membranes.
- Pilot testing should be conducted prior to designing a treatment plant. This would ensure that the most suitable and effective technologies would be selected for the treated water quality requirements and to achieve specific treatment objectives. This should also avoid costly design modifications and retrofitting of additional processing to accommodate design deficiencies.

CHAPTER 6:

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CHAPTER 7:

APPENDICES

APPENDIX 1:

KHP Calibration Curves (UV Absorbance – DOC Relationship)

A stock solution of 1000 mg/L KHP as carbon (C) was prepared and used to prepare working solutions for the calibration curve.

The calibration curve is shown in Figure 1A.

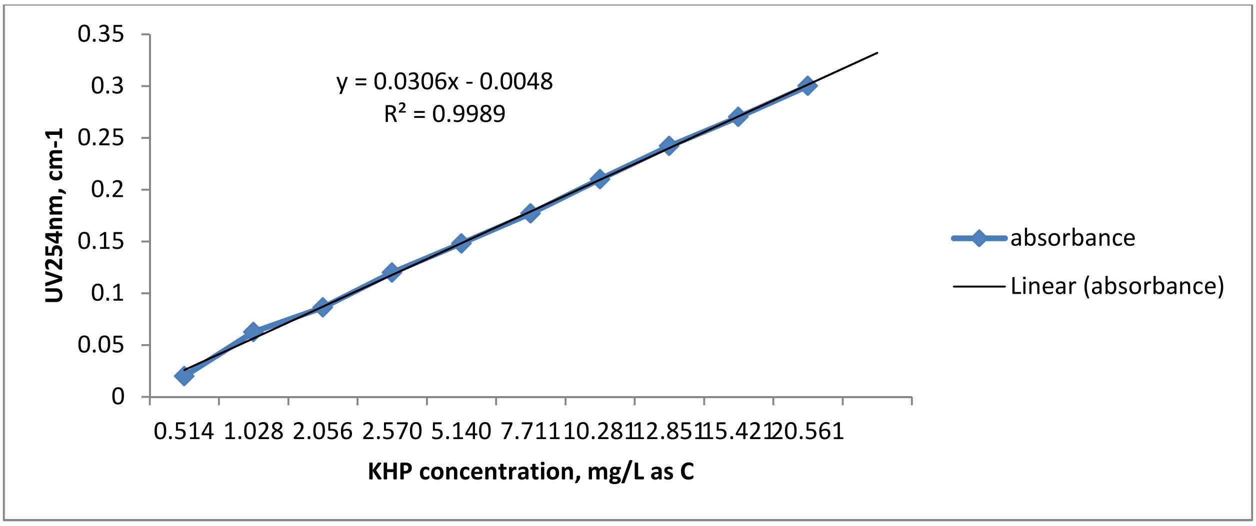


Figure 1A: KHP calibration curve

The coefficient of determination, R^2, is a measurement of variation in the dependent variable (absorbance) which can be attributed to the independent variable (KHP concentration as C) (Kornegay *et al*., 2000). In this case, 97% of the variability in absorbance could be attributed to the variation in KHP concentration as C.

DOC values of the calibration solutions were measured on a TOC analyser as a check to determine if both values correlated. Table 1A below shows measurements of DOC of some of the solutions used to draw up the calibration curve.

Table 1A: Measured DOC of some KHP solution used for calibration curve

KHP as C, mg/L	DOC, mg/L
0.514	0.142
1.028	1.06
2.056	1.99
2.570	2.58
5.140	5.05
7.711	7.53
10.281	10.2
12.851	12.7
15.421	15.4
20.561	20.2

After the correlation was established, the levels of carbon concentration could be estimated graphically, as shown in the curve or by means of the equation also shown above in Figure 1A. The calibration curve was therefore used to determine the expected DOC of the UF feed and permeate (under conditions of no coagulation), using the measured absorbance of the UF feed and permeate and then comparing the carbon concentration obtained from the calibration curve with the actual measured DOC.

The comparison of carbon concentration obtained from the calibration curve with actual measured DOC is shown in Table 1B.

Table 1B: Comparison of carbon concentration (calibration curve values vs actual measured DOC)

	Measured absorbance	Carbon concentration from calibration curve, linear relationship	Measured DOC, mg/L
UF Feed	0.155	5.30	6.02
	0.165	5.63	5.85
	0.171	5.83	6
	0.176	6.00	5.35
UF Permeate	0.181	6.17	6.08
	0.176	6.00	5.81
	0.178	6.07	5.95
	0.172	5.87	5.29

The correlation was also used to determine the DOC in the samples during the pilot runs where absorbance values were available but the TOC analyser was not available.

APPENDIX 2:
Cleaning Procedures

2.1 RO membranes CIP procedure (as per Filmtec cleaning recommendations for 2.5" membranes):

- Make up cleaning solution.

- Low-flow pumping: Pump mixed, preheated cleaning solution to the vessel at conditions of low flow rate and low pressure to displace the process water. Use only enough pressure to compensate for the pressure drop from feed to concentrate. The pressure should be low enough that essentially no or little permeate is produced. A low pressure minimises redeposition of dirt on the membrane. Dump the concentrate, as necessary, to prevent dilution of the cleaning solution.

- Recycle: After the process water is displaced, the cleaning solution will be present in the concentrate stream. Then, recycle the concentrate and permeate to the cleaning solution tank and allow the temperature to stabilise. Measure the pH of the solution and adjust the pH if needed.

- Soak: Turn the pump off and allow the elements to soak. Sometimes a soak period of about one hour is sufficient. For difficult fouling, an extended soak period is beneficial; soak the elements overnight for 10 to 15 hours. In order to maintain a high temperature during an extended soak period, use a slow recirculation rate.

- High-flow pumping: Feed the cleaning solution at the recommended rates for 30 to 60 minutes. The high flow rate flushes out the foulants removed from the membrane surface by the cleaning.

Table 2A: **Recommended flow rates during cleaning**

feed pressure[a], bar	element diameter, inches	feed flow rate per pressure vessel, m^3/h
1.5 - 4	2.5	0.7 - 1.2
1.5 - 4	4	1.8 - 2.3
1.5 - 4	4 full fit	2.7 - 3.2
1.5 - 4	6	3.6 - 4.5
1.5 - 4	8	6.0 - 10.2
1.5 - 8	8 full fit	10.2 - 12.5

[a] Dependent on number of elements in pressure vessel, amount of foulant and flow restriction of the concentrate valve

- Flush out: RO permeate or deionised water is recommended for flushing out the cleaning solution. Pre-filtered raw water or feed water should be avoided as its components may react with the cleaning solution and precipitation of foulants may occur in the membrane elements. The minimum flush out temperature is 20 ℃.

Table 2B: Recommended cleaning solutions for RO membrane organic fouling and biofouling

Organic Fouling	
Cleaning Solutions	**Solution**
Preferred	0.1 wt% NaOH @ pH 12, 30 ℃ maximum, followed by: 0.2% HCl @ pH 2, 45 ℃
Preferred	0.1 wt% NaOH 0.025 wt% Na-DDS@ pH 12, 30 ℃ maximum, followed by : 0.2% HCl @ pH 2, 45 ℃ (Na-DDS is sodium salt of dodecylsulfate)
Alternate	0.1wt% NaOH 1.0wt% Na_4EDTA @ pH 12, 30 ℃ maximum, followed by: 0.2% HCl @ pH 2, 45 ℃
Biofouling	
Preferred	0.1wt% NaOH @ pH 13, 35 ℃ maximum, followed by: 0.2% HCl @ pH 2, 45 ℃
Preferred	0.1wt% NaOH 0.025wt % Na-DDS@ pH 13, 35 ℃ maximum, followed by : 0.2% HCl @ pH 2, 45 ℃ (Na-DDS is sodium salt of dodecylsulfate)
Alternate	0.1wt% NaOH 1.0wt% Na_4EDTA @ pH 13, 35℃ maximum, followed by: 0.2% HCl @ pH 2, 45 ℃

2.2 UF CEB

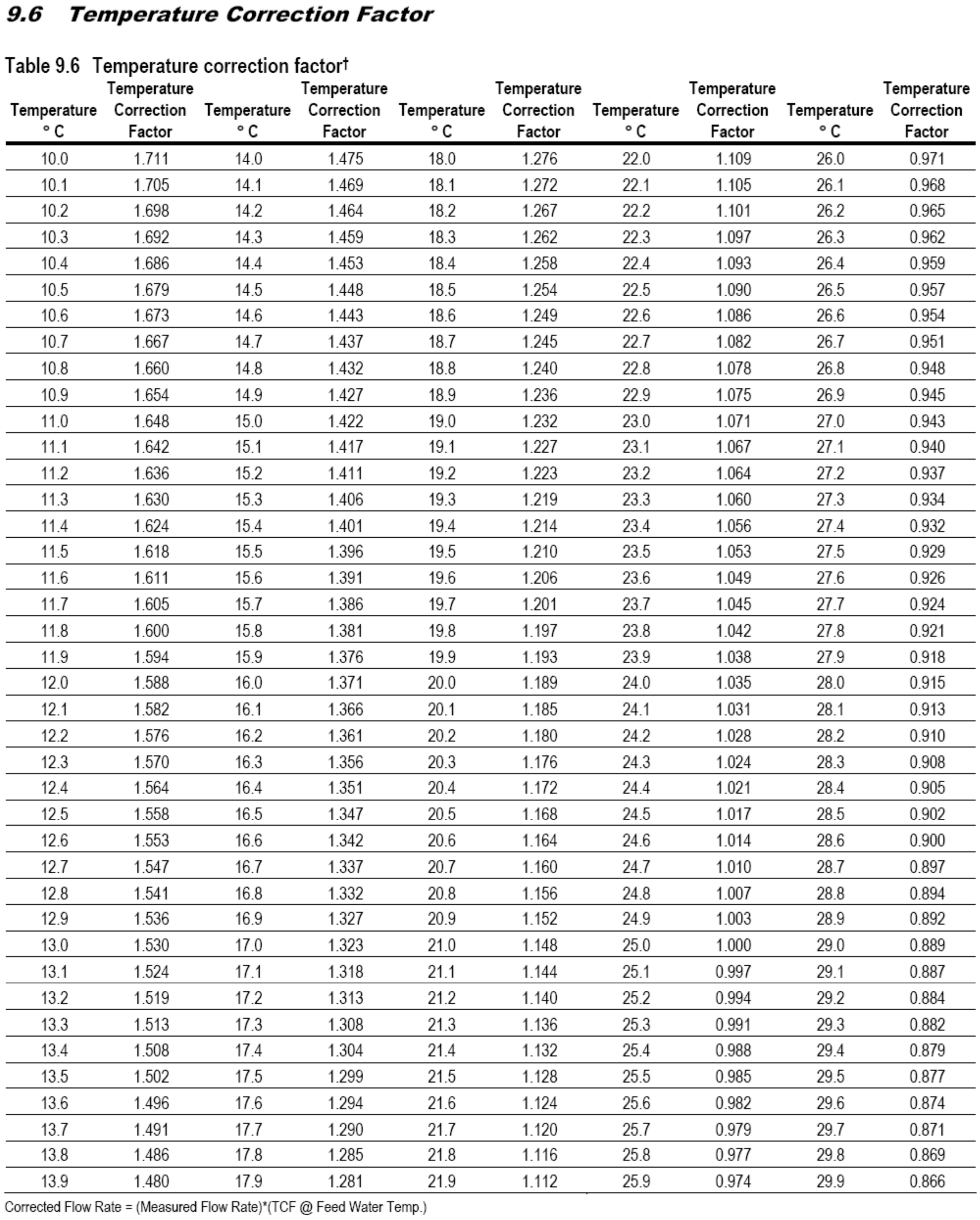
During the CEB, HCl was dosed into the backwash water stream to pH 2. The acid solution was pumped into the membrane for 60 seconds at a flow of 5 000 L/h. After 60 seconds, the acid dosing pump stopped and the membrane was soaked in the acid solution for 10 minutes.

After the soak, the membrane was rinsed with UF permeate, using the backwash pump, for 70 seconds at a flow of 10 m^3/h.

After the acid rinse down, the caustic (NaOH) as well as the sodium hypochlorite (NaOCl) dosing pumps started and dosed the chemicals into the backwash stream. The caustic was dosed to pH 12 and the NaOCl dose used was 200 mg/L.

APPENDIX 3:

Temperature Correction Factors



9.6 Temperature Correction Factor

Table 9.6 Temperature correction factor†

Temperature ° C	Temperature Correction Factor	Temperature ° C	Temperature Correction Factor	Temperature ° C	Temperature Correction Factor	Temperature ° C	Temperature Correction Factor	Temperature ° C	Temperature Correction Factor
10.0	1.711	14.0	1.475	18.0	1.276	22.0	1.109	26.0	0.971
10.1	1.705	14.1	1.469	18.1	1.272	22.1	1.105	26.1	0.968
10.2	1.698	14.2	1.464	18.2	1.267	22.2	1.101	26.2	0.965
10.3	1.692	14.3	1.459	18.3	1.262	22.3	1.097	26.3	0.962
10.4	1.686	14.4	1.453	18.4	1.258	22.4	1.093	26.4	0.959
10.5	1.679	14.5	1.448	18.5	1.254	22.5	1.090	26.5	0.957
10.6	1.673	14.6	1.443	18.6	1.249	22.6	1.086	26.6	0.954
10.7	1.667	14.7	1.437	18.7	1.245	22.7	1.082	26.7	0.951
10.8	1.660	14.8	1.432	18.8	1.240	22.8	1.078	26.8	0.948
10.9	1.654	14.9	1.427	18.9	1.236	22.9	1.075	26.9	0.945
11.0	1.648	15.0	1.422	19.0	1.232	23.0	1.071	27.0	0.943
11.1	1.642	15.1	1.417	19.1	1.227	23.1	1.067	27.1	0.940
11.2	1.636	15.2	1.411	19.2	1.223	23.2	1.064	27.2	0.937
11.3	1.630	15.3	1.406	19.3	1.219	23.3	1.060	27.3	0.934
11.4	1.624	15.4	1.401	19.4	1.214	23.4	1.056	27.4	0.932
11.5	1.618	15.5	1.396	19.5	1.210	23.5	1.053	27.5	0.929
11.6	1.611	15.6	1.391	19.6	1.206	23.6	1.049	27.6	0.926
11.7	1.605	15.7	1.386	19.7	1.201	23.7	1.045	27.7	0.924
11.8	1.600	15.8	1.381	19.8	1.197	23.8	1.042	27.8	0.921
11.9	1.594	15.9	1.376	19.9	1.193	23.9	1.038	27.9	0.918
12.0	1.588	16.0	1.371	20.0	1.189	24.0	1.035	28.0	0.915
12.1	1.582	16.1	1.366	20.1	1.185	24.1	1.031	28.1	0.913
12.2	1.576	16.2	1.361	20.2	1.180	24.2	1.028	28.2	0.910
12.3	1.570	16.3	1.356	20.3	1.176	24.3	1.024	28.3	0.908
12.4	1.564	16.4	1.351	20.4	1.172	24.4	1.021	28.4	0.905
12.5	1.558	16.5	1.347	20.5	1.168	24.5	1.017	28.5	0.902
12.6	1.553	16.6	1.342	20.6	1.164	24.6	1.014	28.6	0.900
12.7	1.547	16.7	1.337	20.7	1.160	24.7	1.010	28.7	0.897
12.8	1.541	16.8	1.332	20.8	1.156	24.8	1.007	28.8	0.894
12.9	1.536	16.9	1.327	20.9	1.152	24.9	1.003	28.9	0.892
13.0	1.530	17.0	1.323	21.0	1.148	25.0	1.000	29.0	0.889
13.1	1.524	17.1	1.318	21.1	1.144	25.1	0.997	29.1	0.887
13.2	1.519	17.2	1.313	21.2	1.140	25.2	0.994	29.2	0.884
13.3	1.513	17.3	1.308	21.3	1.136	25.3	0.991	29.3	0.882
13.4	1.508	17.4	1.304	21.4	1.132	25.4	0.988	29.4	0.879
13.5	1.502	17.5	1.299	21.5	1.128	25.5	0.985	29.5	0.877
13.6	1.496	17.6	1.294	21.6	1.124	25.6	0.982	29.6	0.874
13.7	1.491	17.7	1.290	21.7	1.120	25.7	0.979	29.7	0.871
13.8	1.486	17.8	1.285	21.8	1.116	25.8	0.977	29.8	0.869
13.9	1.480	17.9	1.281	21.9	1.112	25.9	0.974	29.9	0.866

Corrected Flow Rate = (Measured Flow Rate)*(TCF @ Feed Water Temp.)

Source: Dow Technical Manual

APPENDIX 4:

RO Test Membranes Characteristics

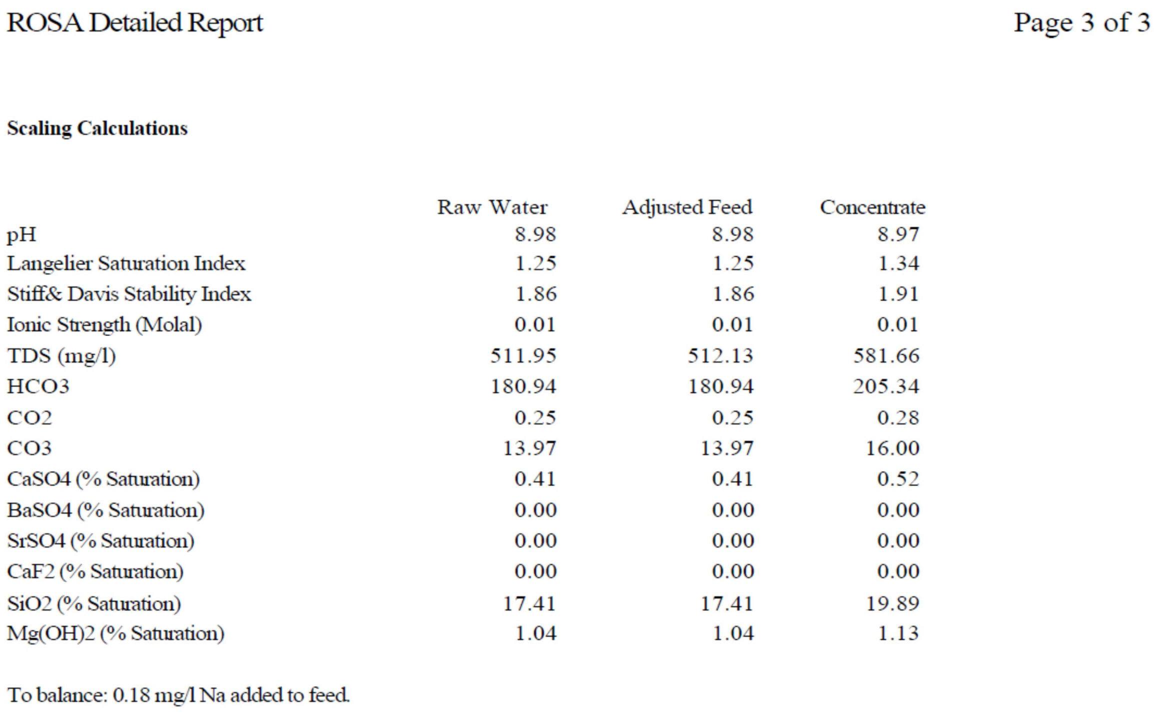
Table 4A: Characteristics of the RO Test Membranes

Membrane	Membrane type	Active area, m^2	Spacer thickness, mil	Permeate flow rate, m^3/d	Stabilized salt rejection, %
BW30-2540	Polyamide thin film composite (PA TFC)	2.6	28	3.2	99.5
BW30XFR-2540	PA TFC	2.6	34		

APPENDIX 5:

RO Feed Water Scaling Potential (ROSA modelling)

Appendix 5.1: Scaling calculations for RO Treatment of concentrated raw water feed



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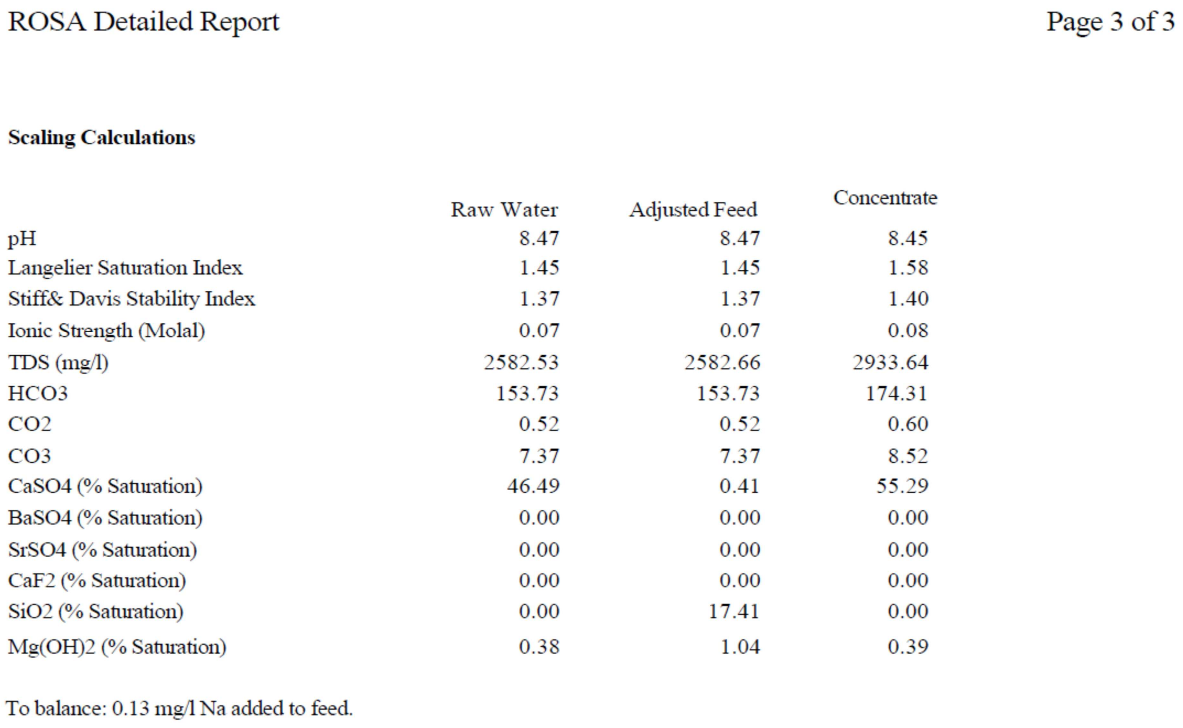
Scaling Calculations

	Raw Water	Adjusted Feed	Concentrate
pH	8.98	8.98	8.97
Langelier Saturation Index	1.25	1.25	1.34
Stiff& Davis Stability Index	1.86	1.86	1.91
Ionic Strength (Molal)	0.01	0.01	0.01
TDS (mg/l)	511.95	512.13	581.66
HCO3	180.94	180.94	205.34
CO2	0.25	0.25	0.28
CO3	13.97	13.97	16.00
CaSO4 (% Saturation)	0.41	0.41	0.52
BaSO4 (% Saturation)	0.00	0.00	0.00
SrSO4 (% Saturation)	0.00	0.00	0.00
CaF2 (% Saturation)	0.00	0.00	0.00
SiO2 (% Saturation)	17.41	17.41	19.89
Mg(OH)2 (% Saturation)	1.04	1.04	1.13

To balance: 0.18 mg/l Na added to feed.

Appendix 5.2: Scaling calculations for RO Treatment of UF filtered CW feed



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Scaling Calculations

	Raw Water	Adjusted Feed	Concentrate
pH	8.47	8.47	8.45
Langelier Saturation Index	1.45	1.45	1.58
Stiff& Davis Stability Index	1.37	1.37	1.40
Ionic Strength (Molal)	0.07	0.07	0.08
TDS (mg/l)	2582.53	2582.66	2933.64
HCO3	153.73	153.73	174.31
CO2	0.52	0.52	0.60
CO3	7.37	7.37	8.52
CaSO4 (% Saturation)	46.49	0.41	55.29
BaSO4 (% Saturation)	0.00	0.00	0.00
SrSO4 (% Saturation)	0.00	0.00	0.00
CaF2 (% Saturation)	0.00	0.00	0.00
SiO2 (% Saturation)	0.00	17.41	0.00
Mg(OH)2 (% Saturation)	0.38	1.04	0.39

To balance: 0.13 mg/l Na added to feed.

APPENDIX 6:

Normalized RO Data

Appendix 6.1: Operational data for the BW30-2540 standard brackish water membrane.

Permeate Flow,L/h	Concentrate Flow, L/h	Feed Pressure, bar	Concentrate Pressure, bar	Calculated Feed TDS, mg/L	Feed Conductivity, µS/cm	Calculated Perm TDS, mg/L	Permeate Conductivity, µS/cm	Calculated Temp (C)	Feed Temp (C)	Feed Flow, L/h	Days of Operation , d	Recovery (%)	Differential Pressure, bar	Temperature Correction Factor	Calculated Feed/Brine Avg Conc	Feed/Brine Osmotic Pressure, bar	Permeate Osmotic Pressure, bar	Net Driving Pressure, bar	Normalized Permeate Flow, L/m^{2}h	Normalized Permeate Salt Passage	
90.00	800.00	10.10	9.76	0.51	1.00	0.06	0.10	24.80	24.80	890.00	0.00	0.10	0.34	0.99	0.54	0.00	0.00	9.93	90.00	0.11	
89.00	800.00	10.00	9.66	0.51	1.00	0.06	0.10	24.50	24.50	889.00	0.00	0.10	0.34	0.99	0.54	0.00	0.00	9.83	90.73	0.11	
82.00	805.00	11.00	10.66	954.50	1899.00	17.72	37.70	24.10	24.10	887.00	0.00	0.09	0.34	0.97	1001.54	0.79	0.01	10.05	82.74	0.02	
70.00	805.00	10.90	10.56	954.50	1899.00	8.51	18.00	21.10	21.10	875.00	1.00	0.08	0.34	0.89	994.84	0.78	0.01	9.96	78.04	0.01	
72.00	805.00	11.10	10.76	954.50	1899.00	11.31	24.00	22.60	22.60	877.00	1.00	0.08	0.34	0.93	995.96	0.78	0.01	10.16	75.21	0.01	
71.10	675.00	9.60	9.34	531.22	1078.00	9.07	19.20	24.00	24.00	746.10	1.00	0.10	0.26	0.97	558.27	0.44	0.01	9.04	80.03	0.01	
72.00	673.00	9.70	9.42			8.97	19.00	24.50	24.50	745.00	1.00	0.10	0.29	0.99	0.00	0.00	0.01	9.56	75.43		
72.00	676.00	9.70	9.43	575.54	1165.00	8.97	19.00	23.70	23.70	748.00	1.00	0.10	0.27	0.96	605.16	0.48	0.01	9.10	81.23	0.01	
75.80	675.00	9.10	8.84	569.93	1154.00	23.43	49.90	21.20	21.20	750.80	2.00	0.10	0.26	0.89	600.79	0.47	0.02	8.52	98.48	0.04	
52.00	622.00	9.80	9.58	2620.13	5000.00	24.89	53.00	22.40	23.40	674.00	2.00	0.08	0.22	0.95	2726.72	2.15	0.02	7.56	71.26	0.01	
37.00	673.00	9.60	9.34	2565.23	4900.00	29.59	63.00	20.00	20.00	710.00	2.00	0.05	0.26	0.86	2634.49	2.06	0.02	7.44	57.11	0.01	
52.80	661.00	9.80	9.58	2730.11	5200.00	19.07	40.60	23.20	23.20	713.80	3.00	0.07	0.22	0.95	2836.36	2.24	0.02	7.47	73.67	0.00	
49.20	629.00	9.70	9.48		0.00	24.89	53.00	24.50	24.50	678.20	3.00	0.07	0.22	0.99	0.00	0.00	0.02	9.61	51.31		
48.00	616.00	9.60	9.38		0.00	27.71	59.00	23.80	23.80	664.00	3.00	0.07	0.22	0.96	0.00	0.00	0.02	9.51	51.64		
30.24	673.00	9.60	9.34	2592.67	4950.00	34.44	73.30	21.00	21.00	703.24	3.00	0.04	0.26	0.89	2650.07	2.08	0.03	7.42	45.36	0.01	
47.52	625.00	9.60	9.38				48.00	23.40	23.40	672.52	4.00	0.07	0.22	0.95	0.00	0.00	0.00	9.49	51.86		
47.52	629.00	9.10	8.87	2609.14	4980.00	18.46	39.30	23.70	23.70	676.52	4.00	0.07	0.23	0.96	2705.31	2.14	0.01	6.86	71.06	0.00	

Permeate Flow,L/h	Concentrate Flow, L/h	Feed Pressure, bar	Concentrate Pressure, bar	Calculated Feed TDS, mg/L	Feed Conductivity, µS/cm	Calculated Perm TDS, mg/L	Permeate Conductivity, µS/cm	Calculated Temp (C)	Feed Temp (C)	Feed Flow, L/h	Days of Operation, d	Recovery (%)	Differential Pressure, bar	Temperature Correction Factor	Calculated Feed/Brine Avg Conc	Feed/Brine Osmotic Pressure, bar	Permeate Osmotic Pressure, bar	Net Driving Pressure, bar	Normalized Permeate Flow, L/m^2h	Normalized Permeate Salt Passage	
81.00	722.00	10.40	10.11	1225.35	2415.00	941.98	1875.00	22.50	22.50	803.00	5.00	0.10	0.29	0.93	1291.65	1.02	0.74	9.98	86.36	0.74	
72.00	795.00	10.20	9.85			17.39	37.00	20.00	20.00	867.00	5.00	0.08	0.35	0.86	0.00	0.00	0.01	10.04	82.36		
59.40	804.00	10.20	9.87	1802.44	3497.00	10.75	22.81	25.10	25.10	863.40	6.00	0.07	0.33	1.00	1867.44	1.48	0.01	8.56	68.30	0.00	
55.50	805.00	10.10	9.77	1805.13	3502.00	13.96	29.68	21.50	21.50	860.50	7.00	0.06	0.33	0.90	1865.97	1.46	0.01	8.48	71.77	0.01	
54.79	804.00	10.30	9.97			9.91	21.00	22.00	22.00	858.79	8.00	0.06	0.33	0.91	0.00	0.00	0.01	10.14	58.37		
57.50	805.00	10.40	10.07	1922.69	3720.00	11.85	25.17	22.90	22.90	862.50	8.00	0.07	0.33	0.94	1989.77	1.57	0.01	8.68	69.68	0.00	
56.21	804.00	10.50	10.17			13.18	28.00	22.50	22.50	860.21	8.00	0.07	0.34	0.93	0.00	0.00	0.01	10.34	57.83		
73.89	807.00	10.60	10.28	1772.31	3441.00	16.45	35.00	30.70	30.70	880.89	9.00	0.08	0.32	1.18	1851.07	1.50	0.01	8.96	68.99	0.01	
55.06	804.00	10.20	9.87			15.05	32.00	23.10	23.10	859.06	9.00	0.06	0.33	0.94	0.00	0.00	0.01	10.05	57.26		
56.66	804.00	10.40	10.07			13.18	28.00	23.40	23.40	860.66	9.00	0.07	0.33	0.95	0.00	0.00	0.01	10.25	57.27		
59.67	804.00	10.20	9.88	1744.89	3390.00	21.98	46.80	25.80	25.80	863.67	9.00	0.07	0.32	1.02	1808.09	1.44	0.02	8.62	66.76	0.01	
58.59	804.00	10.10	9.77	1919.45	3714.00	18.93	40.30	25.40	25.40	862.59	9.00	0.07	0.33	1.01	1987.74	1.58	0.02	8.37	68.27	0.01	
50.64	802.00	10.50	10.17	1987.56	3840.00	18.37	39.10	22.90	22.90	852.64	10.00	0.06	0.33	0.94	2049.03	1.61	0.01	8.74	60.94	0.01	
60.44	802.00	10.20	9.87	1933.49	3740.00	21.60	46.00	26.10	26.10	862.44	10.00	0.07	0.33	1.03	2004.58	1.60	0.02	8.46	68.29	0.01	
61.59	804.00	10.20	9.88	2052.55	3960.00	13.99	29.73	26.30	26.30	865.59	10.00	0.07	0.32	1.04	2129.23	1.70	0.01	8.35	70.05	0.00	
55.84	804.00	10.20	9.87			13.18	28.00	22.20	22.20	859.84	11.00	0.06	0.33	0.92	0.00	0.00	0.01	10.05	59.69		
51.69	802.00	10.20	9.87	2047.13	3950.00	21.75	46.30	23.10	23.10	853.69	11.00	0.06	0.33	0.94	2111.72	1.66	0.02	8.39	64.41	0.01	
52.40	805.00	10.20	9.88	2041.71	3940.00	11.79	25.03	23.60	23.60	857.40	11.00	0.06	0.32	0.96	2106.76	1.66	0.01	8.38	64.34	0.00	
52.40	805.00	10.20	9.88	2041.71	3940.00	11.79	25.03	23.60	23.60	857.40	13.00	0.06	0.32	0.96	2106.76	1.66	0.01	8.38	64.34	0.00	
53.69	802.00	10.20	9.87	2052.55	3960.00	13.99	29.73	26.10	26.10	855.69	15.00	0.06	0.33	1.03	2119.77	1.69	0.01	8.36	61.40	0.00	
53.69	802.00	10.20	9.87	2052.55	3960.00	13.99	29.73	26.10	26.10	855.69	15.00	0.06	0.33	1.03	2119.77	1.69	0.01	8.36	61.40	0.00	

Permeate Flow,L/h	Concentrate Flow, L/h	Feed Pressure, bar	Concentrate Pressure, bar	Calculated Feed TDS, mg/L	Feed Conductivity, µS/cm	Calculated Perm TDS, mg/L	Permeate Conductivity, µS/cm	Calculated Temp (C)	Feed Temp (C)	Feed Flow, L/h	Days of Operation, d	Recovery (%)	Differential Pressure, bar	Temperature Correction Factor	Calculated Feed/Brine Avg Conc	Feed/Brine Osmotic Pressure, bar	Permeate Osmotic Pressure, bar	Net Driving Pressure, bar	Normalized Permeate Flow, L/m^{2}h	Normalized Permeate Salt Passage	
63.00	802.00	10.30	9.97	2047.13	86.60	21.75	22.98	27.10	27.10	865.00	17.00	0.07	0.34	1.06	2125.50	1.70	0.02	8.45	69.17	0.01	
63.00	802.00	10.30	9.97	2041.71	86.60	11.79	22.98	27.10	27.10	865.00	18.00	0.07	0.34	1.06	2119.88	1.69	0.01	8.45	69.20	0.00	
84.40	802.00	10.30	9.97	40.72	86.60	10.83	22.98	27.10	27.10	886.40	19.00	0.10	0.34	1.06	42.79	0.03	0.01	10.11	77.49	0.23	
81.14	808.00	10.20	9.86	42.61	90.60	8.24	17.44	26.20	26.20	889.14	19.00	0.09	0.34	1.04	44.68	0.04	0.01	10.00	77.29	0.17	
73.61	802.00	10.40	10.07			3.83	8.00	24.70	24.70	875.61	20.00	0.08	0.33	0.99	0.00	0.00	0.00	10.24	71.61		
82.15	812.00	10.30	9.97	34.77	74.00	1.30	2.65	27.00	27.00	894.15	20.00	0.09	0.33	1.06	36.47	0.03	0.00	10.11	75.66	0.03	0.97
79.74	1012.00	10.00	9.52			3.83	8.00	26.60	26.60	1091.74	20.00	0.07	0.48	1.05	0.00	0.00	0.00	9.76	76.91		
76.40	1020.00	10.10	9.62	66.22	140.30	1.20	2.44	26.10	26.10	1096.40	20.00	0.07	0.48	1.03	68.64	0.05	0.00	9.80	74.47	0.02	0.98
71.83	1015.00	10.10	9.61	74.43	157.50	1.73	3.55	23.10	23.10	1086.83	21.00	0.07	0.49	0.94	77.01	0.06	0.00	9.80	76.63	0.02	0.98
64.88	1020.00	10.20	9.71	71.33	151.00	0.92	1.84	22.20	22.20	1084.88	21.00	0.06	0.49	0.92	73.55	0.06	0.00	9.90	70.38	0.01	0.99
68.87	1020.00	10.20	9.71	69.89	148.00	0.89	1.79	23.40	23.40	1088.87	22.00	0.06	0.49	0.95	72.20	0.06	0.00	9.90	72.05	0.01	0.99
62.15	1018.00	10.30	9.97	70.85	150.00	1.21	2.44	24.40	24.40	1080.15	22.00	0.06	0.33	0.98	72.97	0.06	0.00	10.08	61.99	0.01	0.99
59.74	1010.00	10.20	9.52	66.08	140.00	1.73	3.54	22.40	22.40	1069.74	23.00	0.06	0.68	0.93	67.99	0.05	0.00	9.81	65.01	0.02	0.98
56.40	1014.00	10.10	9.62	53.70	114.00	2.27	4.68	25.20	25.20	1070.40	24.00	0.05	0.48	1.01	55.17	0.04	0.00	9.82	56.39	0.03	0.97

Appendix 6.2: Operational data for the BW30XFR-2540 low fouling membrane

Permeate Flow,L/h	Concentrate Flow, L/h	Feed Pressure, bar	Concentrate Pressure, bar	Calculated Feed TDS, mg/L	Feed Conductivity, µS/cm	Calculated Perm TDS, mg/L	Permeate Conductivity, µS/cm	Calculated Temp (C)	Feed Temp (C)	Feed Flow, L/h	Days of Operation , d	Recovery (%)	Differential Pressure, bar	Temperature Correction Factor	Calculated Feed/Brine Avg Conc	Feed/Brine Osmotic Pressure, bar	Permeate Osmotic Pressure, bar	Net Driving Pressure, bar	Normalized Permeate Flow, L/m^2h	Normalized Permeate Salt Rejection, %
96	717	10.0	9.8	1736.8	3375.0	87.1	184.0	23.3	813		12%	0.2	1.0	1848.2	1.5	0.1	8.5	96.000	5.0%	95.0%
90	720	10.0	9.8	1736.8	3375.0	87.1	184.0	21.2	810		11%	0.2	0.9	1841.1	1.4	0.1	8.5	95.758	5.0%	95.0%
88	729	9.8	9.6	1736.8	3375.0	84.7	179.0	20.5	817	1	11%	0.2	0.9	1837.7	1.4	0.1	8.3	97.855	4.9%	95.1%
92	710	10.0	9.8	1736.8	3375.0	66.6	141.0	25.1	802	1	11%	0.2	1.0	1844.8	1.5	0.1	8.5	87.391	3.5%	96.5%
89	714	10.0	9.8	1736.8	3375.0	77.5	164.0	25.6	803	1	11%	0.2	1.0	1840.8	1.5	0.1	8.5	83.176	3.9%	96.1%
83	726	10.1	9.9	1736.8	3375.0	87.6	185.0	21.8	809	2	10%	0.2	0.9	1832.5	1.4	0.1	8.6	85.637	4.6%	95.4%
88	722	10.0	9.8	1736.8	3375.0	86.6	183.0	24.1	810	2	11%	0.2	1.0	1838.6	1.5	0.1	8.5	85.901	4.5%	95.5%
88	720	10.0	9.8	1688.5	3285.0	85.2	180.0	23.9	808	2	11%	0.2	1.0	1787.7	1.4	0.1	8.6	85.991	4.6%	95.4%
78	732	9.9	9.7	1262.3	2485.0	37.7	80.3	19.3	810	3	10%	0.2	0.8	1327.3	1.0	0.0	8.8	85.274	2.8%	97.2%
89	723	10.1	9.9			54.7	116.0	22.4	812	3	11%	0.2	0.9	0.0	0.0	0.0	10.0	77.534		
88	723	10.0	9.8			58.0	123.0	22.9	811	3	11%	0.2	0.9	0.0	0.0	0.0	10.0	76.253		
92	716	10.0	9.8	1264.4	2489.0	57.0	121.0	23.7	808	3	11%	0.2	1.0	1342.4	1.1	0.0	8.9	87.091	4.3%	95.7%
90	716	10.1	9.9	1211.1	2388.0	44.6	94.8	23.1	806	3	11%	0.2	0.9	1284.2	1.0	0.0	9.0	85.410	3.5%	96.5%
80	725	10.0	9.8	1181.0	2331.0	38.1	81.0	20.2	805	4	10%	0.2	0.9	1243.9	1.0	0.0	9.0	83.522	3.0%	97.0%
87	805	10.1	9.9	1067.5	2115.0	13.2	28.0	22.3	892	4	10%	0.2	0.9	1123.3	0.9	0.0	9.1	83.993	1.2%	98.8%
91	799	10.4	10.2	340.6	700.0	11.8	25.0	23.8	890	4	10%	0.2	1.0	359.3	0.3	0.0	10.0	76.403	3.3%	96.7%
102	808	9.9	9.7			16.9	36.0	25.3	910	4	11%	0.2	1.0	0.0	0.0	0.0	9.8	83.111		
82	722	10.3	10.1	1629.5	3175.0	143.2	300.0	19.9	804	8	10%	0.2	0.9	1718.8	1.3	0.1	9.0	86.329	8.4%	91.6%
81	728	9.8	9.6			147.5	309.0	20.5	809	8	10%	0.2	0.9	0.0	0.0	0.1	9.8	76.525		
95	714	10.0	9.8	1586.2	3094.0	150.9	316.0	23.9	809	8	12%	0.2	1.0	1687.3	1.3	0.1	8.7	91.451	9.3%	90.7%

Permeate Flow,L/h	Concentrate Flow, L/h	Feed Pressure, bar	Concentrate Pressure, bar	Calculated Feed TDS, mg/L	Feed Conductivity, µS/cm	Calculated Perm TDS, mg/L	Permeate Conductivity, µS/cm	Calculated Temp (C)	Feed Temp (C)	Feed Flow, L/h	Days of Operation, d	Recovery (%)	Differential Pressure, bar	Temperature Correction Factor	Calculated Feed/Brine Avg Conc	Feed/Brine Osmotic Pressure, bar	Permeate Osmotic Pressure, bar	Net Driving Pressure, bar	Normalized Permeate Flow, L/m^{2}h	Normalized Permeate Salt Rejection, %
97	711	10.2	10.0			151.4	317.0	24.2	808	8	12%	0.2	1.0	0.0	0.0	0.1	10.2	78.664		
91	719	10.0	9.8			145.6	305.0	23.9	810	8	11%	0.2	1.0	0.0	0.0	0.1	10.0	75.984		
90	723	10.0	9.8	1464.5	2866.0	143.2	300.0	22.6	813	8	11%	0.2	0.9	1552.1	1.2	0.1	8.8	89.001	9.4%	90.6%
72	750	10.3	10.1	1538.1	3004.0	180.2	376.0	15.5	822	8	9%	0.2	0.7	1609.7	1.2	0.1	9.1	85.781	11.4%	88.6%
88	726	10.1	9.9			153.4	321.0	22.8	814	9	11%	0.2	0.9	0.0	0.0	0.1	10.1	75.177		
109	698	10.2	10.0			22.5	48.0	24.7	807	9	14%	0.2	1.0	0.0	0.0	0.0	10.1	87.933		
88	722	10.1	9.9	1027.7	2039.0	40.0	85.0	19.9	810	9	11%	0.2	0.9	1087.9	0.8	0.0	9.2	90.483	4.0%	96.0%
86	726	9.9	9.7	1061.2	2103.0	31.9	68.0	21.2	812	10	11%	0.2	0.9	1121.7	0.9	0.0	9.0	87.197	2.9%	97.1%
83	729	10.1	9.9			24.9	53.0	22.8	812	10	10%	0.2	0.9	0.0	0.0	0.0	10.0	71.598		
83	720	10.1	9.9			24.4	52.0	22.5	803	10	10%	0.2	0.9	0.0	0.0	0.0	10.0	72.252		
90	726	9.9	9.7	1335.4	2623.0	26.8	57.0	22.7	816	10	11%	0.2	0.9	1414.9	1.1	0.0	8.7	89.586	1.9%	98.1%
81	701	9.1	8.9	1408.1	2760.0	49.9	106.0	22.7	782	11	10%	0.2	0.9	1486.5	1.2	0.0	7.9	89.117	3.1%	96.9%
87	725	10.0	9.8	1242.2	2447.0	33.8	72.0	25.7	812	11	11%	0.2	1.0	1314.0	1.0	0.0	8.9	77.605	2.3%	97.7%
75	728	10.0	9.8	1382.1	2711.0	69.9	148.0	20.0	803	11	9%	0.2	0.9	1451.0	1.1	0.1	8.8	80.008	4.4%	95.6%
80	726	10.1	9.9			69.4	147.0	23.0	806	12	10%	0.2	0.9	0.0	0.0	0.1	10.1	68.347		
90	792	9.9	9.7			6.2	13.0	23.8	882	12	10%	0.2	1.0	0.0	0.0	0.0	9.8	77.121		
84	801	9.8	9.6	193.4	403.0	8.5	18.0	23.8	885	12	9%	0.2	1.0	203.2	0.2	0.0	9.5	73.949	3.8%	96.2%
50	804	10.0	9.8	104.1	219.3	3.7	7.8	20.8	854	12	6%	0.2	0.9	107.2	0.1	0.0	9.8	46.375	2.0%	98.0%
87	808	10.0	9.8	126.8	266.3	6.2	13.0	23.8	895	13	10%	0.2	1.0	133.4	0.1	0.0	9.8	74.560	4.4%	95.6%
65	798	10.0	9.8	108.3	228.1	13.4	28.5	21.8	863	13	8%	0.2	0.9	112.7	0.1	0.0	9.8	59.279	9.0%	91.0%
123	679	11.1	10.9	892.5	1780.0	177.3	370.0	23.3	802	17	15%	0.2	1.0	968.9	0.8	0.1	10.4	101	24.9%	75.1%
95	537	10.9	10.8	955.5	1901.0	215.5	448.0	22.2	632	17	15%	0.1	0.9	1035.5	0.8	0.2	10.2	82	22.7%	77.3%
87	474	9.0	8.9	1133.1	2240.0	262.9	544.0	22.2	561	17	16%	0.1	0.9	1231.3	1.0	0.2	8.2	94	21.3%	78.7%
89	473	9.2	9.1	1146.3	2265.0	329.6	678.0	23.4	562	18	16%	0.1	1.0	1247.9	1.0	0.3	8.4	90	26.0%	74.0%
88	807	10.5	10.3	987.4	1962.0	15.0	32.0	27.0	895	20	10%	0.2	1.1	1039.5	0.8	0.0	9.6	70	1.3%	98.7%
86	810	10.4	10.2	951.4	1893.0	13.3	28.4	22.9	896	22	10%	0.2	0.9	1000.1	0.8	0.0	9.5	78	1.3%	98.7%
90	805	10.1	9.9	1012.5	2010.0	10.9	23.2	21.9	895	23	10%	0.2	0.9	1066.9	0.8	0.0	9.2	87	1.1%	98.9%
87	474	9.0	8.9	1133.1	2240.0	262.9	544.0	24.2	561	29	16%	0.1	1.0	1231.3	1.0	0.2	8.2	88	20.0%	80.0%

Permeate Flow,L/h	Concentrate Flow, L/h	Feed Pressure, bar	Concentrate Pressure, bar	Calculated Feed TDS, mg/L	Feed Conductivity, µS/cm	Calculated Perm TDS, mg/L	Permeate Conductivity, µS/cm	Calculated Temp (C)	Feed Temp (C)	Feed Flow, L/h	Days of Operation , d	Recovery (%)	Differential Pressure, bar	Temperature Correction Factor	Calculated Feed/Brine Avg Conc	Feed/Brine Osmotic Pressure, bar	Permeate Osmotic Pressure, bar	Net Driving Pressure, bar	Normalized Permeate Flow, L/m^{2}h	Normalized Permeate Salt Rejection, %
89	473	9.2	9.1	1146.3	2265.0	329.6	678.0	25.4	562	31	16%	0.1	1.0	1247.9	1.0	0.3	8.4	85	24.5%	75.5%
88	807	10.5	10.3	987.4	1962.0	15.0	32.0	27.0	895	33	10%	0.2	1.1	1039.5	0.8	0.0	9.6	70	1.3%	98.7%
70	810	10.4	10.2	951.4	1893.0	13.3	28.4	23.9	880	35	8%	0.2	1.0	991.3	0.8	0.0	9.5	61	1.0%	99.0%
71	805	10.1	9.9	1012.5	2010.0	10.9	23.2	24.9	876	36	8%	0.2	1.0	1055.7	0.8	0.0	9.2	63	0.8%	99.2%
97	804	10.4	10.2			1.9	4.0	23.2	901	37	11%	0.2	0.9	0.0	0.0	0.0	10.3	80		
75	805	10.2	10.0			8.5	18.0	22.1	880	37	9%	0.2	0.9	0.0	0.0	0.0	10.1	66		
79	804	10.6	10.4	1441.1	2822.0	8.5	18.0	22.3	883	37	9%	0.2	0.9	1509.6	1.2	0.0	9.3	74	0.5%	99.5%
77	807	10.2	10.0			9.0	19.0	22.1	884	37	9%	0.2	0.9	0.0	0.0	0.0	10.1	67		
75	807	10.3	10.1	1437.4	2815.0	5.2	11.0	22.3	882	38	9%	0.2	0.9	1502.2	1.2	0.0	9.0	73	0.3%	99.7%
83	807	10.2	10.0			11.8	25.0	23.9	890	38	9%	0.2	1.0	0.0	0.0	0.0	10.1	69		
83	805	10.2	10.0			12.2	26.0	24.1	888	38	9%	0.2	1.0	0.0	0.0	0.0	10.1	68		
81	805	10.3	10.1	1554.1	3034.0	7.8	16.5	24.8	886	38	9%	0.2	1.0	1629.9	1.3	0.0	8.9	74	0.4%	99.6%
70	802	10.2	10.0	1557.3	3040.0	4.8	10.0	21.6	872	40	8%	0.2	0.9	1623.5	1.3	0.0	8.8	71	0.2%	99.8%
74	801	10.2	10.0			8.0	17.0	23.3	875	40	8%	0.2	1.0	0.0	0.0	0.0	10.1	63		
75	804	10.3	10.1	1547.7	3022.0	8.0	17.0	23.8	879	40	9%	0.2	1.0	1617.7	1.3	0.0	8.9	70	0.4%	99.6%
72	810	10.4	10.2	1893.0	3665.0	4.8	10.0	23.5	882	41	8%	0.2	1.0	1974.7	1.6	0.0	8.7	70	0.2%	99.8%
67	808	10.3	10.1	1713.2	3331.0	8.0	17.0	22.1	875	41	8%	0.2	0.9	1782.7	1.4	0.0	8.8	68	0.3%	99.7%
72	810	10.2	10.0			8.5	18.0	21.4	882	41	8%	0.2	0.9	0.0	0.0	0.0	10.1	64		
71	805	10.3	10.1	1893.0	3665.0	8.0	17.0	24.2	876	42	8%	0.2	1.0	1974.1	1.6	0.0	8.6	68	0.3%	99.7%
71	805	10.3	10.1	1621.0	3159.0	8.5	18.0	23.8	876	42	8%	0.2	1.0	1690.2	1.3	0.0	8.9	67	0.4%	99.6%
71	802	10.2	10.0			8.0	17.0	23.0	879	42	8%	0.2	0.9	0.0	0.0	0.0	10.1	59		
69	810	10.3	10.1			8.0	17.0	22.1	874	42	8%	0.2	0.9	0.0	0.0	0.0	10.2	58		
67	807	10.6	10.4			4.8	10.0	21.6	874	42	8%	0.2	0.9	0.0	0.0	0.0	10.5	58		
69	807	10.2	10.0	1553.1	3032.0	8.5	18.0	23.9	876	43	8%	0.2	1.0	1617.3	1.3	0.0	8.8	65	0.4%	99.6%
87	805	10.0	9.8	1347.1	2645.0	7.5	15.9	26.6	892	48	10%	0.2	1.0	1417.7	1.1	0.0	8.8	77	0.5%	99.5%
87	801	10.2	10.0	1280.3	2519.0	22.9	48.8	26.1	888	50	10%	0.2	1.0	1347.3	1.1	0.0	9.0	75	1.5%	98.5%
87	812	10.2	10.0	1280.3	2519.0	8.5	18.0	26.1	899	50	10%	0.2	1.0	1346.9	1.1	0.0	9.0	76	0.6%	99.4%
84	804	10.0	9.8	1280.3	2519.0	7.1	15.0	25.4	888	51	9%	0.2	1.0	1344.9	1.1	0.0	8.8	76	0.5%	99.5%
82	808	10.0	9.8	1319.0	2592.0	5.4	11.3	25.0	890	51	9%	0.2	1.0	1383.8	1.1	0.0	8.8	75	0.3%	99.7%

78	804	10.0	9.8			3.4	7.0	21.8	882	52	9%	0.2	0.9	0.0	0.0	0.0	9.9	70		

APPENDIX 7:

EXPERIMENTAL SET-UP: KOMATI DESALINATION PLANT PROCESS FLOW DIAGRAM

The process flow diagram for the Komati Power Station desalination plant is shown in Figure 7A.

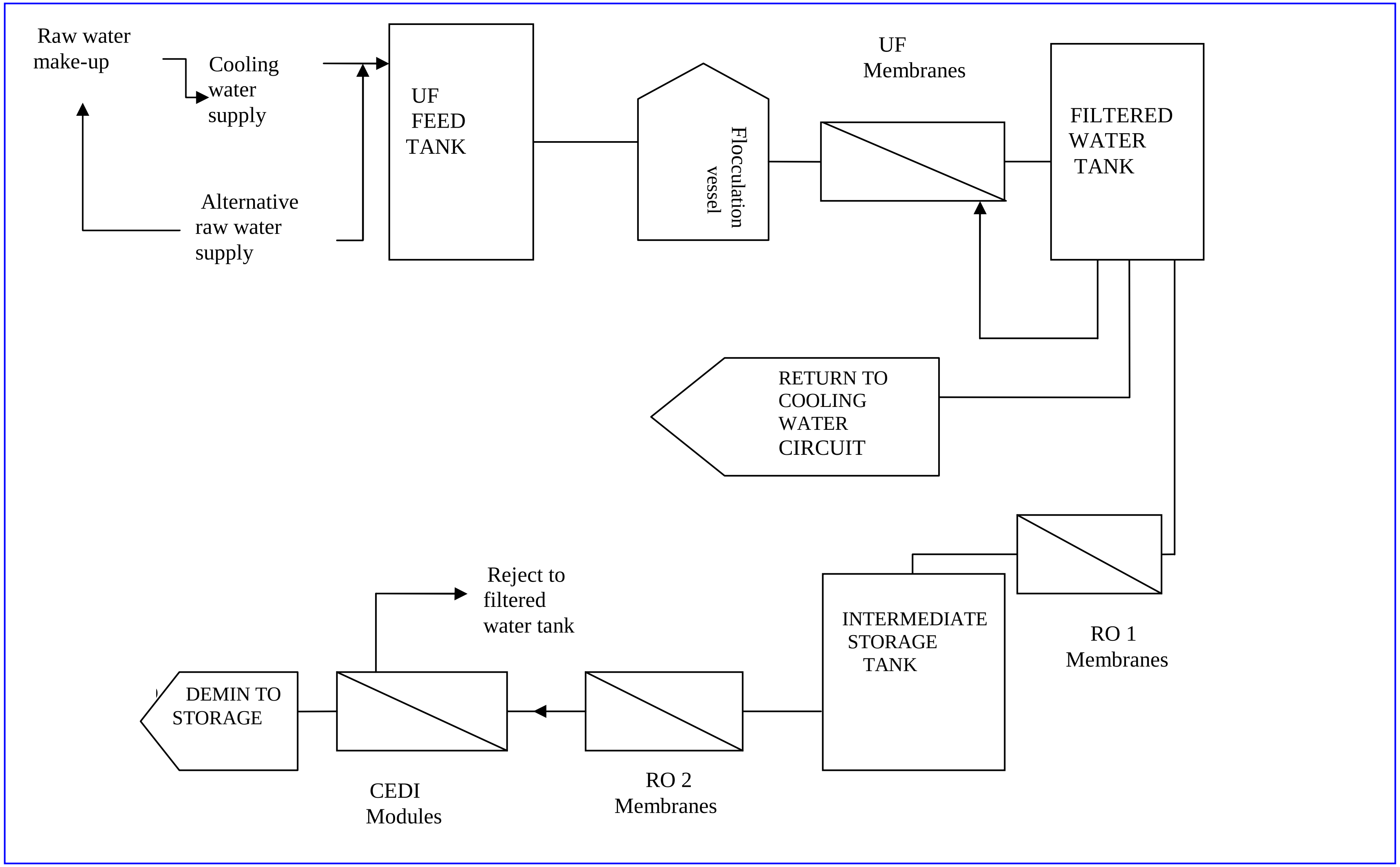

Figure 7A: Komati desalination plant process overview

The plant uses cooling tower blowdowns as feed water. In the event that cooling water is not available for use due to oil contamination, an alternative raw water supply is used to provide feed water to the plant.

The plant consists of four UF skids and three trains of double-pass three-stage RO. The final polishing of the water is achieved on four continuous electro-deionisation (CEDI) skids.

Coagulant is dosed in-line at the suction of the UF feed pumps, with a recirculation loop installed at the discharge of the feed pumps for mixing of the coagulant.

The UF is operated in the dead-end mode, inside-out configuration. The hydraulic backwashes are performed every 20 minutes and the chemically enhanced backwash (CEB) is performed twice per day using NaOH, NaOCl and HCl.

The filtered water from the UF provides:

- UF backwash water;
- Recycling (return stream) to the cooling water;
- Feed or make-up source for demineralisation.

Desalination of the filtered water feed stream is achieved on a RO plant consisting of the standard RO membranes for brackish water. In order to achieve the final water quality required for boiler feedwater make-up, a two-pass multi-staged RO array configuration is used with final polishing done through a CEDI process. De-carbonation (removal of dissolved carbon dioxide) ahead of the CEDI is achieved by inter-pass caustic injection on the RO system.

The setup of the pilot plants was derived from the configuration of the main plant.

APPENDIX 8:

FEED WATER QUALITY

Appendix 8.1 Komati Raw water TOC

Table 8A: Komati Raw Water TOC

Date	KOM-RAW- TOC as C (mg/kg)	Average	Minimum	Maximum
2008/04/07 08:00	8.14	6.292066	3.5	8.45
2008/04/14 08:00	8.2	6.292066	3.5	8.45
2008/04/21 08:00	7.57	6.292066	3.5	8.45
2008/05/05 08:00	7.59	6.292066	3.5	8.45
2008/05/12 08:00	7.6	6.292066	3.5	8.45
2008/05/19 08:00	7.61	6.292066	3.5	8.45
2008/05/26 08:00	7.91	6.292066	3.5	8.45
2008/06/02 08:00	7.42	6.292066	3.5	8.45
2008/06/09 08:00	7.83	6.292066	3.5	8.45
2008/06/23 08:00	8.19	6.292066	3.5	8.45
2008/06/30 08:00	8.03	6.292066	3.5	8.45
2008/07/07 08:00	8.45	6.292066	3.5	8.45
2008/07/14 08:00	8.21	6.292066	3.5	8.45
2008/07/21 08:00	7.75	6.292066	3.5	8.45
2008/07/28 08:00	7.35	6.292066	3.5	8.45
2008/08/04 08:00	7.95	6.292066	3.5	8.45
2008/08/11 08:00	7.43	6.292066	3.5	8.45
2008/08/18 08:00	7.8	6.292066	3.5	8.45
2008/10/20 08:00	5.82	6.292066	3.5	8.45
2008/10/27 08:00	6.44	6.292066	3.5	8.45
2008/11/03 08:00	5.61	6.292066	3.5	8.45
2008/11/10 08:00	5.93	6.292066	3.5	8.45
2008/11/17 08:00	6.29	6.292066	3.5	8.45
2008/12/08 08:00	5.32	6.292066	3.5	8.45
2009/01/05 08:00	5.87	6.292066	3.5	8.45
2009/01/12 08:00	5.96	6.292066	3.5	8.45
2009/01/19 08:00	6.38	6.292066	3.5	8.45
2009/01/26 08:00	6.23	6.292066	3.5	8.45

2009/02/02 08:00	6.56	6.292066	3.5	8.45
2009/02/09 08:00	6.25	6.292066	3.5	8.45
2009/02/16 08:00	5.85	6.292066	3.5	8.45
2009/02/23 08:00	6.54	6.292066	3.5	8.45
2009/03/02 08:00	6.21	6.292066	3.5	8.45
2009/03/09 08:00	6.76	6.292066	3.5	8.45
2009/03/16 08:00	5.14	6.292066	3.5	8.45
2009/03/23 08:00	7.6	6.292066	3.5	8.45
2009/03/30 08:00	7.3	6.292066	3.5	8.45
2009/04/06 08:00	6.72	6.292066	3.5	8.45
2009/04/20 08:00	7.34	6.292066	3.5	8.45
2009/05/18 08:00	7.02	6.292066	3.5	8.45
2009/05/25 08:00	6.59	6.292066	3.5	8.45
2009/06/01 08:00	6.52	6.292066	3.5	8.45
2009/06/08 08:00	7.06	6.292066	3.5	8.45
2009/06/15 08:00	6.74	6.292066	3.5	8.45
2009/06/22 08:00	6.63	6.292066	3.5	8.45
2009/06/29 08:00	6.74	6.292066	3.5	8.45
2009/07/06 08:00	5.24	6.292066	3.5	8.45
2009/07/13 08:00	6.73	6.292066	3.5	8.45
2009/07/20 08:00	5.38	6.292066	3.5	8.45
2009/07/27 08:00	6.55	6.292066	3.5	8.45
2009/08/03 08:00	6.43	6.292066	3.5	8.45
2009/08/17 08:00	5.5	6.292066	3.5	8.45
2009/08/24 08:00	6.71	6.292066	3.5	8.45
2009/08/31 08:00	6.62	6.292066	3.5	8.45
2009/09/07 08:00	6.44	6.292066	3.5	8.45
2009/09/14 08:00	6.47	6.292066	3.5	8.45
2009/09/21 08:00	5.76	6.292066	3.5	8.45
2009/09/28 08:00	4.16	6.292066	3.5	8.45
2009/10/05 08:00	5.67	6.292066	3.5	8.45
2009/11/02	5.71	6.292066	3.5	8.45
2009/11/09	5.74	6.292066	3.5	8.45
2009/11/16	5.68	6.292066	3.5	8.45
2009/11/23	5.54	6.292066	3.5	8.45
2009/11/30	5.82	6.292066	3.5	8.45
2009/12/07	5.73	6.292066	3.5	8.45
2009/12/14	3.5	6.292066	3.5	8.45
2009/12/28	5.74	6.292066	3.5	8.45

2010/01/04	5.85	6.292066	3.5	8.45
2010/01/11	5.78	6.292066	3.5	8.45
2010/01/18	5.78	6.292066	3.5	8.45
2010/01/25	5.95	6.292066	3.5	8.45
2010/02/01	6.15	6.292066	3.5	8.45
2010/02/08	6.07	6.292066	3.5	8.45
2010/02/15	6.17	6.292066	3.5	8.45
2010/02/22	6.24	6.292066	3.5	8.45
2010/03/01	6.06	6.292066	3.5	8.45
2010/03/08	6.09	6.292066	3.5	8.45
2010/03/15	5.97	6.292066	3.5	8.45
2010/03/29	5.88	6.292066	3.5	8.45
2010/04/12	6.12	6.292066	3.5	8.45
2010/04/12	5.86	6.292066	3.5	8.45
2010/04/26	5.69	6.292066	3.5	8.45
2010/05/03	6.33	6.292066	3.5	8.45
2010/05/10	5.92	6.292066	3.5	8.45
2010/05/17	6.03	6.292066	3.5	8.45
2010/05/24	5.73	6.292066	3.5	8.45
2010/05/31	5.33	6.292066	3.5	8.45
2010/06/07	6.01	6.292066	3.5	8.45
2010/06/14	5.63	6.292066	3.5	8.45
2010/06/21	6.26	6.292066	3.5	8.45
2010/06/28	6.23	6.292066	3.5	8.45
2010/07/05	6.46	6.292066	3.5	8.45
2010/07/12	6.05	6.292066	3.5	8.45
2010/07/19	6.32	6.292066	3.5	8.45
2010/08/16	5.54	6.292066	3.5	8.45
2010/08/23	4.63	6.292066	3.5	8.45
2010/08/30	5.52	6.292066	3.5	8.45
2010/09/06	5.46	6.292066	3.5	8.45
2010/09/13	5.36	6.292066	3.5	8.45
2010/09/20	5.06	6.292066	3.5	8.45
2010/09/27	5.52	6.292066	3.5	8.45
2010/10/04	5.38	6.292066	3.5	8.45
2010/10/11	5.95	6.292066	3.5	8.45
2010/10/18	6.26	6.292066	3.5	8.45
2010/10/25	5.34	6.292066	3.5	8.45

2010/11/01	6.86	6.292066	3.5	8.45
2010/11/08	5.52	6.292066	3.5	8.45
2010/11/15	5.6	6.292066	3.5	8.45
2010/11/22	5.73	6.292066	3.5	8.45
2010/11/29	5.76	6.292066	3.5	8.45
2010/12/06	5.79	6.292066	3.5	8.45
2010/12/13	5.66	6.292066	3.5	8.45
2010/12/20	5.88	6.292066	3.5	8.45
2011/01/03	5.33	6.292066	3.5	8.45
2011/01/10	5.76	6.292066	3.5	8.45
2011/01/17	6.31	6.292066	3.5	8.45
2011/01/24	7.24	6.292066	3.5	8.45
2011/01/31	6.68	6.292066	3.5	8.45
2011/02/07	6.89	6.292066	3.5	8.45
2011/02/14	7.12	6.292066	3.5	8.45
2011/02/21	5.66	6.292066	3.5	8.45

Appendix 8.2 Komati Cooling Water West Forebay TOC

Table 8B: Cooling water west TOC

KOM-MAIN-CW-WEST/TOC as C (mg/L)				
Date	**TOC**	**Minimum**	**Average**	**Maximum**
2009/09/18 08:00	71.9	20.5	53.04143	94.8
2009/05/25 08:00	71.2	20.5	53.04143	94.8
2009/06/01 08:00	36.9	20.5	53.04143	94.8
2009/06/08 08:00	52.3	20.5	53.04143	94.8
2009/06/15 08:00	73.4	20.5	53.04143	94.8
2009/06/22 08:00	78.3	20.5	53.04143	94.8
2009/06/29 08:00	64	20.5	53.04143	94.8
2009/07/06 08:00	74.6	20.5	53.04143	94.8
2009/07/13 08:00	73.1	20.5	53.04143	94.8
2009/07/20 08:00	74.8	20.5	53.04143	94.8
2009/07/27 08:00	73.2	20.5	53.04143	94.8
2009/08/03 08:00	78.7	20.5	53.04143	94.8
2009/08/10 08:00	74	20.5	53.04143	94.8
2009/08/17 08:00	72.9	20.5	53.04143	94.8
2009/08/24 08:00	69.6	20.5	53.04143	94.8
2009/08/31 08:00	47.7	20.5	53.04143	94.8
2009/09/09 08:00	21.3	20.5	53.04143	94.8
2009/10/05	28.1	20.5	53.04143	94.8
2009/11/02	40.7	20.5	53.04143	94.8
2009/11/09	40.3	20.5	53.04143	94.8
2009/11/16	49.2	20.5	53.04143	94.8
2009/11/23	48	20.5	53.04143	94.8
2009/11/23	48	20.5	53.04143	94.8
2009/12/07	53	20.5	53.04143	94.8
2009/12/14	52	20.5	53.04143	94.8
2010/01/04	63.5	20.5	53.04143	94.8
2010/01/11	68.9	20.5	53.04143	94.8
2010/02/15	20.8	20.5	53.04143	94.8
2010/02/22	24.2	20.5	53.04143	94.8
2010/03/01	28.5	20.5	53.04143	94.8
2010/03/08	24.9	20.5	53.04143	94.8
2010/03/15	21.2	20.5	53.04143	94.8

2010/03/29	20.5	20.5	53.04143	94.8
2010/04/12	25	20.5	53.04143	94.8
2010/04/19	21	20.5	53.04143	94.8
2010/05/10 08:00	25.20	20.5	53.04143	94.8
2010/05/17 08:00	23.00	20.5	53.04143	94.8
2010/05/24 08:00	21.20	20.5	53.04143	94.8
2010/06/21 08:00	24.70	20.5	53.04143	94.8
2010/06/28 08:00	30.10	20.5	53.04143	94.8
2010/07/05 08:00	39.90	20.5	53.04143	94.8
2010/07/12 08:00	47.00	20.5	53.04143	94.8
2010/07/19 08:00	41.00	20.5	53.04143	94.8
2010/08/16 08:00	64.70	20.5	53.04143	94.8
2010/08/23 08:00	54.30	20.5	53.04143	94.8
2010/08/30 08:00	66.00	20.5	53.04143	94.8
2010/09/06 08:00	60.60	20.5	53.04143	94.8
2010/09/13 08:00	51.20	20.5	53.04143	94.8
2010/09/20 08:00	59.10	20.5	53.04143	94.8
2010/09/27 08:00	63.00	20.5	53.04143	94.8
2010/10/04 08:00	58.40	20.5	53.04143	94.8
2010/10/11 08:00	64.00	20.5	53.04143	94.8
2010/10/18 08:00	64.90	20.5	53.04143	94.8
2010/11/01 08:00	58.50	20.5	53.04143	94.8
2010/11/08 08:00	62.00	20.5	53.04143	94.8
2010/11/15 08:00	63.80	20.5	53.04143	94.8
2010/11/22 08:00	61.20	20.5	53.04143	94.8
2010/11/29 08:00	51.70	20.5	53.04143	94.8
2010/12/06 08:00	56.00	20.5	53.04143	94.8
2010/12/13 08:00	52.40	20.5	53.04143	94.8
2010/12/20 08:00	56.20	20.5	53.04143	94.8
2010/12/27 08:00	59.10	20.5	53.04143	94.8
2011/01/03 08:00	67.00	20.5	53.04143	94.8
2011/01/10 08:00	23.00	20.5	53.04143	94.8
2011/01/17 08:00	57.80	20.5	53.04143	94.8
2011/01/24 08:00	85.00	20.5	53.04143	94.8
2011/01/31 08:00	58.00	20.5	53.04143	94.8
2011/02/07 08:00	86.40	20.5	53.04143	94.8
2011/02/14 08:00	76.00	20.5	53.04143	94.8
2011/02/21 08:00	94.80	20.5	53.04143	94.8

APPENDIX 9:

TURBIDITY AND DOC REMOVAL FROM RAW WATER (JAR TEST RESULTS)

Raw Water Data	
pH	7.79 at 22.1deg celc
K25, µS/cm	185.6
Turbidity, NTU	2.18
TOC, mg/L	5.73
Absorbance, cm^{-1}	0.176
DOC, mg/L	5.43
POC	0.3

	Membra Quest 3000a						
Coagulant Dose, mg/L	**pH**	**Conductivity, µS/cm**	**Turbidity, NTU**	**Absorbance at 254nm, cm^{-1}**	**DOC, mg/L**	**Turbidity Removal, %**	**DOC Removal, %**
5	7.82	188.9	0.276	0.13	4.98	87.34	8.29
10	7.78	191.1	1.06	0.12	4.43	51.38	18.42
15	7.8	190.8	0.321	0.12	5.22	94.40	3.87
20	7.78	189.3	0.275	0.11	5.03	94.94	7.37
25	7.72	191.3	0.347	0.18	36.4ppb/9.55ppm	84.08	
30							

	PAC						
Coagulant Dose, mg/L	**pH**	**Conductivity, µS/cm**	**Turbidity, NTU**	**Absorbance at 254nm, cm^{-1}**	**DOC, mg/L**	**Turbidity Removal, %**	**DOC Removal, %**
5	7.45	190.2	0.108	0.13	5.03	95.05	7.37
10	7.38	194	0.155	0.14	5.08	92.89	6.45
15	7.3	193.2	0.366	0.12	5.09	83.21	6.26
20	7.23	196.2	0.139	0.11	5.05	93.62	7.00
25	7.25	197.4	0.147	0.10	4.33	93.26	20.26
30	7.21	194.5	0.141	0.10	4.26	93.53	21.55

Raw Water Data	
pH	7.71 at 21.4deg celc
K25, µS/cm	208.7
Turbidity, NTU	2.34
TOC, mg/L	4.96
Absorbance, cm^{-1}	0.118
DOC, mg/L	6.16
POC	-1.2

	U3000						
Coagulant dose, mg/L	**pH**	**Conductivity, µS/cm**	**Turbidity, NTU**	**Absorbance at 254nm, cm^{-1}**	**DOC, mg/L**	**Turbidity Removal, %**	**DOC Removal, %**
5	7.7	190.3	1.96	0.16	5.49	16.24	10.88
10	7.6	188.3	0.137	0.11	4.36	94.15	29.22
15	7.53	211.8	0.225	0.11	4.13	90.38	32.95
20	7.44	193.7	0.196	0.09	3.98	91.62	35.39
25	7.38	199.6	0.63	0.09	4.18	73.08	32.14
30	7.35	200.5	0.149	0.08	3.69	93.63	40.10

	FeCl3						
Coagulant dose, mg/L	**pH**	**Conductivity, µS/cm**	**Turbidity, NTU**	**Absorbance at 254nm, cm^{-1}**	**DOC, mg/L**	**Turbidity Removal, %**	**DOC Removal, %**
5	6.88	215.4	0.911	0.25	4.66	61.07	24.35
10	6.63	204.4	0.885	0.14	4.39	62.18	28.73
15	6.61	203.8	0.267	0.12	4.08	88.59	33.77
20	6.55	199.1	0.253	0.12	4.8	89.19	22.08
25	6.16	207.2	0.394	0.07	2.6	83.16	57.79
30	4.91	288.5	0.385	0.24	3.16	83.55	48.70

Raw Water Data	
pH	8.16 at 17.5deg celc
K25, µS/cm	
Turbidity, NTU	2.99
TOC, mg/L	3.84
Absorbance, cm^{-1}	0.157
DOC, mg/l	4.32

	Alum						
Coagulant dose, mg/L	**pH**	**Conductivity, µS/cm**	**Turbidity, NTU**	**Absorbance at 254nm, cm^{-1}**	**DOC, mg/L**	**Turbidity Removal, %**	**DOC Removal, %**
5	6.9		0.302	0.09	3.14	89.90	27.31
10	6.23		0.24	0.06	2.26	91.97	47.69
15	5.51		0.67	0.07	2.54	77.59	41.20
20	7.04		0.39	0.07	3.79	86.96	12.27
25	6.16		2.41	0.07	3.66	19.40	15.28
30							

APPENDIX 10:

WATER CHARACTERISTICS DURING UF TREATMENT

Appendix 10.1 Water Characteristics During UF Treatment Without Coagulation

Time, hours	**Feed Pressure, kPa**	**Permeate flow, L/h**	**Feed DOC, mg/L**	**Permeate DOC, mg/L**	**Feed Absorbance, cm^{-1}**	**Permeate Abs, cm^{-1}**	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	**FEED SUVA (ratio of UV$_{254}$/DOC)**	**Permeate SUVA**	**%SUVA Removal**
0		3000									
6		3000	6.02	5.85	0.139	0.142	0.99	1.01	2.31	2.40	-4.06
7		3150									
8		3250									
11		3250	5.05	6.2	0.155	0.181	1.19	0.99	3.07	2.98	3.01
12		3350									
17	18	3500	5.61	5.53	0.161	0.145	0.99	1.00	2.87	2.62	8.64
22	18	3350									
					0.163						
28		3350									
34	20	3500									
50	20	3400									
59		3281.8182									
64	18	3000	5.5	6	0.165	0.176	0.92	1.02	3.00	2.93	2.22
68	20	3500	5.8	6	0.165	0.176	1.01	1.01	2.84	2.93	-3.11
83	10	3500									
99		3500									

Time, hours	**Feed Pressure, kPa**	**Permeate flow, L/h**	**Feed DOC, mg/L**	**Permeate DOC, mg/L**	**Feed Absorbance, cm^{-1}**	**Permeate Abs, cm^{-1}**	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	**FEED SUVA (ratio of UV$_{254}$/DOC)**	**Permeate SUVA**	**%SUVA Removal**
109		3400									
123		3500									
167		3450									
179	20	3500									
210	22	3500	6	6.09	0.171	0.178	0.97	0.98	2.85	2.92	-2.56
218	22	3250						1.01			
235		3410									
244	30	3550						1.01			
247	30	3500	5.35	5.22	0.159	0.163	0.99	1.00	2.97	3.12	-5.07
289	30	3550									
295	25	3500									
319	22	3450	5.36	5	0.162	0.154	1.07	1.01	3.02	3.08	-1.91
349	22	3550									
359	22	3550									
392		3521.4286									
427	10?	3550									
444	22	3500	5.87	5.9	0.172	0.162	1.05	1.03	2.93	2.75	6.29
494	22	3400									
511											
545	22	3500	5.9	5.89	0.178	0.163	1.05	1.04	3.02	2.77	8.27
584	22	3500									
612	20	3550	5.77	5.85	0.174	0.162	1.04	1.01	3.02	2.77	8.17
637	20	3500									
		3491.6667									
642	25	3500	5.99	5.82	0.172	0.165			2.87	2.84	1.27
676	25	3000									
684											
700	25	3500	6.8	6.23	0.194	0.181	0.86	1.00	2.85	2.91	-1.84
721	25	3500									

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Feed DOC, mg/L	Permeate DOC, mg/L	Feed Absorbance, cm^{-1}	Permeate Abs, cm^{-1}	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	FEED SUVA (ratio of UV_{254}/DOC)	Permeate SUVA	%SUVA Removal
726	25	3400									
732	25	3500									
754	22	3500	5.88	5.82	0.177	0.165		1.01	3.01	2.84	5.82
767	20	3500						1.04			
782	20	3500	5.35	5.77	0.176	0.172	0.93	1.00	3.29	2.98	9.39
810	20	3400	6.73	6.42	0.192	0.185	0.98	1.04	2.85		100.00
		3430									
816	30	3500									
818		3500									
821		3600	4.11	4.88	0.13	0.149	1.03	1.02	3.16	3.05	3.47
824		3500									
827		3550									
830		3500									
832		3600									
835		3550	4.34	4.74	0.162	0.154	0.90	1.02	3.73	3.25	12.96
838		3500									
841											
844		3550									
846		3500									
849		3550									
852		3480	4.34	4.74	0.162	0.154	0.90	1.02	3.73	3.25	12.96
855		3500									
858		3500									
860		3500									
863		3480									
866		3490									
869		3500	4.6	4.73	0.146	0.15	1.01	1.01	3.17	3.17	0.08
872		3520									
874		3500									
		3517.619									
877	20	3500									

Time, hours	**Feed Pressure, kPa**	**Permeate flow, L/h**	**Feed DOC, mg/L**	**Permeate DOC, mg/L**	**Feed Absorbance, cm^{-1}**	**Permeate Abs, cm^{-1}**	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	**FEED SUVA (ratio of UV_{254}/DOC)**	**Permeate SUVA**	**%SUVA Removal**
880		3500									
883		3520									
886		3550									
888		3500									
891		3500	3.97	4.5	0.114	0.12	1.03	1.03	2.87	2.67	7.13
894		3500									
897		3550					1.00	1.01			
899		3500									
900		3500									
		3512									

Appendix 10.2: Water Characteristics During UF Treatment With Coagulation

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Feed DOC, mg/L	Permeate DOC, mg/L	Feed Absorbance, cm^{-1}	Permeate Abs, cm^{-1}	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	FEED SUVA (ratio of UV$_{254}$/DOC)	Permeate SUVA	% SUVA REMOVAL
8.00	22.00	3500.00									
12.00		3550.00									
17.00		3550.00	3.90	2.84	0.15	0.10	0.99	1.16	3.72	3.35	10.03
21.00		3550.00									
25.00		3550.00									
29.00		3500.00									
33.00		3550.00									
37.00		3500.00	4.16	3.28	0.15	0.10	0.93	1.04	3.49	3.11	10.78
41.00		3550.00									
45.00		3520.00									
50.00		3500.00									
54.00		3500.00	4.06	2.88	0.15	0.10	0.95	1.26	3.57	3.51	1.81
58.00		3500.00									
62.00		3550.00									
66.00		3520.00	3.96	3.32	0.15	0.11	0.97	1.04	3.66	3.31	9.51
70.00		3550.00									
74.00		3500.00									
79.00		3500.00									
83.00		3550.00	4.26	2.84	0.15	0.10	0.91	1.26	3.40	3.35	1.72
83.00	22.00	3600.00									
87.00		3500.00									
91.00		3600.00									
95.00		3550.00									
99.00		3500.00									
103.00		3550.00									

Time, hours	**Feed Pressure, kPa**	**Permeate flow, L/h**	**Feed DOC, mg/L**	**Permeate DOC, mg/L**	**Feed Absorbance, cm^{-1}**	**Permeate Abs, cm^{-1}**	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	**FEED SUVA (ratio of UV_{254}/DOC)**	**Permeate SUVA**	**% SUVA REMOVAL**
107.00		3500.00	4.45	3.76	0.13	0.11	1.06	1.12	3.01	2.85	5.50
111.00		3500.00									
116.00		3500.00									
120.00		3500.00									
124.00		3550.00									
128.00		3520.00									
132.00		3500.00									
132.00	22.00	3500.00	3.70	2.50	0.15	0.10	1.01	1.24	4.14	3.96	4.24
136.00		3550.00									
140.00		3500.00									
161.00		3550.00									
165.00		3550.00									
194.00		3500.00									
198.00		3500.00	4.15	2.85	0.15	0.10	0.90	1.08	3.69	3.47	5.78
202.00		3550.00									
207.00		3520.00									
211.00		3500.00									
215.00		3520.00	4.24	2.26	0.15	0.08	0.88	1.16	3.61	3.36	6.81
219.00		3550.00									
223.00		3550.00									
227.00		3520.00									
231.00		3520.00	3.99	2.40	0.15	0.08	0.94	1.19	3.83	3.21	16.33
236.00		3520.00									
240.00		3540.00									
244.00		3550.00									
248.00		3520.00	4.09	2.87	0.15	0.11	0.92	1.06	3.74	3.66	2.20
256.00		3500.00									
264.00		3500.00									
314.00		3500.00	4.04	3.18	0.13	0.09	1.03	1.06	3.09	2.86	7.51
322.00		3500.00									
331.00		3500.00									
339.00		3500.00									
347.00		3500.00	4.10	3.36	0.13	0.10	1.01	1.14	3.05	2.83	7.26

Time, hours	**Feed Pressure, kPa**	**Permeate flow, L/h**	**Feed DOC, mg/L**	**Permeate DOC, mg/L**	**Feed Absorbance, cm^{-1}**	**Permeate Abs, cm^{-1}**	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	**FEED SUVA (ratio of UV$_{254}$/DOC)**	**Permeate SUVA**	**% SUVA REMOVAL**
355.00		3500.00	4.74	3.87	0.13	0.10	0.88	0.75	2.64	2.56	3.00
363.00		3520.00									
430.00		3500.00	4.94	2.78	0.14	0.08	0.94	1.10	2.75	2.70	2.00
438.00		3520.00									
446.00		3500.00	4.94	2.72	0.14	0.07	0.94	1.00	2.75	2.61	5.18
455.00		3550.00									
463.00		3500.00	5.04	2.76	0.14	0.07	0.92	0.97	2.70	2.54	6.01
512.00		3520.00	5.04	2.97	0.14	0.08	0.92	1.33	2.70	2.59	3.92
513.00	38.00	3500.00									
534.00		3500.00	4.32	2.76	0.16	0.08	0.89	0.84	3.63	3.04	16.26
538.00		3500.00									
546.00		3520.00									
550.00		3520.00									
555.00		3500.00									
563.00		3500.00									
571.00		3490.00	4.88	5.12	0.16	0.16	0.86	0.82	3.30	3.13	5.28
579.00		3490.00									
600.00		3500.00	4.32	4.02	0.16	0.14	0.95	0.92	3.66	3.41	6.82
612.00		3480.00	4.32	4.02	0.16	0.14	0.95	0.90	3.66	3.41	6.82
621.00		3500.00									
629.00		3500.00									
637.00		2600.00									
664.00		2600.00									
667.00		2600.00	4.07	2.82	0.14	0.09	0.94	0.97	3.42	3.05	10.70
671.00		2550.00									
675.00		2550.00									
683.00		2550.00									
692.00		2530.00									
700.00		2530.00	4.47	3.01	0.14	0.09	0.86	0.96	3.11	2.89	7.05
708.00		2600.00									
717.00		2550.00	4.47	2.99	0.14	0.08	0.86	1.07	3.11	2.78	10.73
725.00		2600.00									

Time, hours	**Feed Pressure, kPa**	**Permeate flow, L/h**	**Feed DOC, mg/L**	**Permeate DOC, mg/L**	**Feed Absorbance, cm^{-1}**	**Permeate Abs, cm^{-1}**	FEED RATIO OF TOC:DOC	PERMEATE RATIO OF TOC:DOC	**FEED SUVA (ratio of UV$_{254}$/DOC)**	**Permeate SUVA**	**% SUVA REMOVAL**
733.00		2600.00	4.47	3.02	0.14	0.09	0.86	0.93	3.11	2.91	6.29
738.00		2700.00									
742.00		2690.00									
750.00		2690.00	4.32	2.76	0.16	0.09			3.68	3.08	16.32
759.00		2500.00									
792.00		2690.00									
800.00		2800.00					0.93	1.05			

APPENDIX 11:

UF TREATMENT WITHOUT COAGULATION

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Flux, $L/m^2.h$	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	%DOC removal, minimum	%DOC removal, maximum	Feed Turbidity, NTU	Permeate Turbidity, NTU	Turbidity Removal, %	Turbidity Removal, %, minimum	Turbidity Removal, %, maximum	Turbidity Removal, %, average	Normalized Flow	Normalized flux, L/m^2h	Comments
0.00		3000.00	75.00									3.72	13.09	8.90	2871.00	71.78	Temp≈26.5deg Celcius
																	Tripped on feed tank low level
6.00		3000.00	75.00	6.02	5.85	1.17	-0.17	4.01	2.42	2.33	3.72	3.72	13.09	8.90	2871.00	71.78	
7.00		3150.00	78.75			2.82	-0.17	4.01						8.90	3014.55	75.36	
																	Backwash
8.00		3250.00	81.25				-0.17	4.01						8.90	3110.25	77.76	
11.00		3250.00	81.25	5.05	6.20	0.00	-0.17	4.01	3.30	2.98	9.70	3.72	13.09	8.90	3110.25	77.76	
12.00		3350.00	83.75				-0.17	4.01				3.72	13.09	8.90	3205.95	80.15	opened feed pump disch IV almost 100%
17.00	18.00	3500.00	87.50	5.61	5.53	0.18	-0.17	4.01	1.91	1.66	13.09	3.72	13.09	8.90	3349.50	83.74	
22.00	18.00	3350.00	83.75			1.43	-0.17	4.01						8.90	3205.95	80.15	
																	backwash
28.00		3350.00	83.75				-0.17	4.01						8.90	3205.95	80.15	
34.00	20.00	3500.00	87.50				-0.17	4.01						8.90	3349.50	83.74	
50.00	20.00	3400.00	85.00				-0.17	4.01						8.90	3253.80	81.35	
59.00		3281.82	82.05				-0.17	4.01						8.90	3140.70	78.52	CEB

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Flux, L/m^2.h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	%DOC removal, minimum	%DOC removal, maximum	Feed Turbidity, NTU	Permeate Turbidity, NTU	Turbidity Removal, %	Turbidity Removal, %, minimum	Turbidity Removal, %, maximum	Turbidity Removal, %, average	Normalized Flow	Normalized flux L/m^{2}h,	Comments
64.00	18.00	3000.00	75.00	5.50	6.00	1.98	-0.17	4.01	1.58	1.44	8.86	3.72	13.09	8.90	2829.00	70.73	Plant started at 11H00. Temp≈27 deg Celc??
																	CEB
68.00	20.00	3500.00	87.50	5.80	6.00	0.68	-0.17	4.01	1.62	1.43	11.73	3.72	13.09	8.90	3300.50	82.51	
83.00	10.00	3500.00	87.50				-0.17	4.01							3300.50	82.51	
99.00		3500.00	87.50				-0.17	4.01							3300.50	82.51	
109.00		3400.00	85.00				-0.17	4.01	1.77	1.59	10.17	3.72	13.09	8.90	3206.20	80.16	
123.00		3500.00	87.50				-0.17	4.01							3300.50	82.51	
																	CEB
167.00		3450.00	86.25				-0.17	4.01	1.35	1.21	10.37	3.72	13.09	8.90	3253.35	81.33	
179.00	20.00	3500.00	87.50				-0.17	4.01							3300.50	82.51	
210.00	22.00	3500.00	87.50	6.00	6.09	0.00	-0.17	4.01	1.62	1.53	5.56	3.72	13.09	8.90	3300.50	82.51	
218.00	22.00	3250.00	81.25			0.83	-0.17	4.01							3064.75	76.62	
235.00		3410.00	85.25														CEB
244.00	30.00	3550.00	88.75			0.83	-0.17	4.01							3347.65	83.69	
247.00	30.00	3500.00	87.50	5.35	5.22	0.38	-0.17	4.01	7.30	6.83	6.44	3.72	13.09	8.90	3202.50	80.06	Temp: 28deg celc
289.00	30.00	3550.00	88.75			2.43	-0.17	4.01							3248.25	81.21	
295.00	25.00	3500.00	87.50				-0.17	4.01							3202.50	80.06	

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Flux, L/m^2.h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	%DOC removal, minimum	%DOC removal, maximum	Feed Turbidity, NTU	Permeate Turbidity, NTU	Turbidity Removal, %	Turbidity Removal, %, minimum	Turbidity Removal, %, maximum	Turbidity Removal, %, average	Normalized Flow	Normalized flux, L/m^{2}h	Comments
319.00	22.00	3450.00	86.25	5.36	5.00	1.22	-0.17	4.01	1.24	1.19	4.03	3.72	13.09	8.90	3156.75	78.92	
																	Backwash
349.00	22.00	3550.00	88.75				-0.17	4.01							3248.25	81.21	
359.00	22.00	3550.00	88.75				-0.17	4.01							3248.25	81.21	
392.00		3521.43	88.04														Backwash
427.00	10?	3550.00	88.75				-0.17	4.01							3248.25	81.21	
444.00	22.00	3500.00	87.50	5.87	5.90	2.77	-0.17	4.01	2.13	1.99	6.57	3.72	13.09	8.90	3318.00	82.95	Temp 26.8
494.00	22.00	3400.00	85.00				-0.17	4.01							3223.20	80.58	
511.00																	CEB
545.00	22.00	3500.00	87.50	5.90	5.89	3.87	-0.17	4.01	2.42	2.23	7.85	3.72	13.09	8.90	3318.00	82.95	
584.00	22.00	3500.00	87.50				-0.17	4.01							3318.00	82.95	
612.00	20.00	3550.00	88.75	5.77	5.85	1.33	-0.17	4.01	1.55	1.41	9.03	3.72	13.09	8.90	3365.40	84.14	
637.00	20.00	3500.00	87.50				-0.17	4.01							3318.00	82.95	
		3491.67	87.29				-0.17	4.01							3310.10	82.75	
642.00	25.00	3500.00	87.50	5.99	5.82	2.85	-0.17	4.01	1.78	1.60	10.11	3.72	13.09	8.90	3559.50	88.99	temp 24.5
676.00	25.00	3000.00	75.00				-0.17	4.01							3051.00	76.28	
684.00																	CEB
700.00	25.00	3500.00	87.50	6.80	6.23	-0.17	-0.17	4.01	2.04	1.86	8.82	3.72	13.09	8.90	3559.50	88.99	

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Flux, L/m^2.h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	%DOC removal, minimum	removal, maximum	Feed Turbidity, NTU	Permeate Turbidity, NTU	Turbidity Removal, %	Turbidity Removal, %, minimum	Turbidity Removal, %, maximum	Turbidity Removal, %, average	Normalized Flow	Normalized flux, L/m^{2}h	Comments
721.00	25.00	3500.00	87.50				-0.17	4.01							3559.50	88.99	
726.00	25.00	3400.00	85.00				-0.17	4.01							3457.80	86.45	
732.00	25.00	3500.00	87.50				-0.17	4.01							3559.50	88.99	
																	CEB
754.00	22.00	3500.00	87.50	5.88	5.82	0.68	-0.17	4.01	2.11	1.90	9.95	3.72	13.09	8.90	3559.50	88.99	
767.00	20.00	3500.00	87.50			4.01	-0.17	4.01							3559.50	88.99	
782.00	20.00	3500.00	87.50	5.35	5.77	0.40	-0.17	4.01	2.23	1.98	11.12	3.72	13.09	8.90	3559.50	88.99	
810.00	20.00	3400.00	85.00	6.73	6.42	3.48	-0.17	4.01							3471.40	86.79	
		3430.00	85.75				-0.17	4.01							3502.03	87.55	temp 24.4
816.00	30.00	3500.00	87.50				-0.17	4.01							3573.50	89.34	
818.00		3500.00	87.50				-0.17	4.01							3675.60	91.89	
821.00		3600.00	90.00	4.11	4.88	2.35	-0.17	4.01	1.23	1.10	10.57	3.72	13.09	8.90	3573.50	89.34	
824.00		3500.00	87.50				-0.17	4.01							3624.55	90.61	
827.00		3550.00	88.75				-0.17	4.01							3573.50	89.34	
830.00		3500.00	87.50				-0.17	4.01							3675.60	91.89	
832.00		3600.00	90.00				-0.17	4.01							3624.55	90.61	
835.00		3550.00	88.75	4.34	4.74	2.05	-0.17	4.01	1.50	1.34	10.67	3.72	13.09	8.90	3573.50	89.34	
838.00		3500.00	87.50				-0.17	4.01									
841.00			0.00				-0.17	4.01							3624.55	90.61	

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Flux, L/m^2.h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	%DOC removal, minimum	%DOC removal, maximum	Feed Turbidity, NTU	Permeate Turbidity, NTU	Turbidity Removal, %	Turbidity Removal, %, minimum	Turbidity Removal, %, maximum	Turbidity Removal, %, average	Normalized Flow	Normalized flux, L/m^{2}h	Comments
844.00		3550.00	88.75				-0.17	4.01							3573.50	89.34	
846.00		3500.00	87.50				-0.17	4.01							3624.55	90.61	
849.00		3550.00	88.75				-0.17	4.01							3553.08	88.83	
852.00		3480.00	87.00	4.34	4.74	2.05	-0.17	4.01	1.50	1.34	10.67	3.72	13.09	8.90	3573.50	89.34	
855.00		3500.00	87.50				-0.17	4.01							3573.50	89.34	
858.00		3500.00	87.50				-0.17	4.01							3573.50	89.34	
860.00		3500.00	87.50				-0.17	4.01							3608.50	90.21	temp 24.1
863.00		3480.00	87.00				-0.17	4.01							3587.88	89.70	
866.00		3490.00	87.25				-0.17	4.01							3598.19	89.95	
869.00		3500.00	87.50	4.60	4.73	0.86	-0.17	4.01	1.28	1.16	9.38	3.72	13.09	8.90	3608.50	90.21	
872.00		3520.00	88.00				-0.17	4.01							3629.12	90.73	
874.00		3500.00	87.50				-0.17	4.01							3608.50	90.21	
		3517.62	87.94				-0.17	4.01							3626.67	90.67	
877.00	20.00	3500.00	87.50				-0.17	4.01							3549.00	88.73	temp 24.6
880.00		3500.00	87.50				-0.17	4.01							3549.00	88.73	
883.00		3520.00	88.00				-0.17	4.01							3569.28	89.23	
886.00		3550.00	88.75				-0.17	4.01							3599.70	89.99	
888.00		3500.00	87.50				-0.17	4.01							3549.00	88.73	

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Flux, L/m^2.h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	%DOC removal, minimum	%DOC removal, maximum	Feed Turbidity, NTU	Permeate Turbidity, NTU	Turbidity Removal, %	Turbidity Removal, %, minimum	Turbidity Removal, %, maximum	Turbidity Removal, %, average	Normalized Flow	Normalized flux L/m^{2}h,	Comments
891.00		3500.00	87.50	3.97	4.50	3.17	-0.17	4.01	1.33	1.20	9.77	3.72	13.09	8.90	3549.00	88.73	
894.00		3500.00	87.50				-0.17	4.01							3549.00	88.73	
897.00		3550.00	88.75				-0.17	4.01							3599.70	89.99	
899.00		3500.00	87.50				-0.17	4.01							3549.00	88.73	
900.00		3500.00	87.50				-0.17	4.01							3549.00	88.73	
		3512.00	87.80				-0.17	4.01							3561.17	89.03	

APPENDIX 12:

UF TREATMENT WITH COAGULATION

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	Feed Absorbance, cm^{-1}	Permeate Abs, cm^{-1}	Abs Removal, cm^{-1}	Normalized Flow	Normalized Flux, L/m^2h	Comments
8	22	3500							3500	87.5	corrected flow rate = measured flow rate * (TCF @ feedwater temp
12		3550							3550	88.75	temp 25
17		3550	3.9	2.84	14.50777	0.145	0.095	34.482759	3550	88.75	
21		3550							3550	88.75	
25		3550							3550	88.75	
29		3500							3500	87.5	
33		3550							3550	88.75	
37		3500	4.16	3.28	11.65803	0.145	0.102	29.655172	3500	87.5	
41		3550							3550	88.75	
45		3520							3520	88	
50		3500							3500	87.5	
54		3500	4.06	2.88	5.699482	0.145	0.101	30.344828	3500	87.5	
58		3500							3500	87.5	
62		3550							3550	88.75	
66		3520	3.96	3.32	10.62176	0.145	0.11	24.137931	3520	88	
70		3550							3550	88.75	
74		3500							3500	87.5	
79		3500							3500	87.5	
83		3550	4.26	2.84	7.253886	0.145	0.095	34.482759	3550	88.75	
83	22	3600							3546	88.65	temp 25.5
87		3500							3447.5	86.1875	
91		3600							3546	88.65	
95		3550							3496.75	87.41875	
99		3500							3447.5	86.1875	
103		3550							3496.75	87.41875	

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	Feed Absorbance, cm^{-1}	Permeate Abs, cm^{-1}	Abs Removal, cm^{-1}	Normalized Flow	Normalized Flux, L/m^2h	Comments
107		3500	4.45	3.76	15.50562	0.134	0.107	20.149254	3447.5	86.1875	
111		3500							3447.5	86.1875	
116		3500							3447.5	86.1875	
120		3500							3447.5	86.1875	
124		3550							3496.75	87.41875	
128		3520							3467.2	86.68	
132		3500							3447.5	86.1875	
132	22	3500	3.7	2.5	17.6	0.153	0.099	35.294118	3559.5	88.9875	temp 24.5
136		3550							3610.35	90.25875	
140		3500							3559.5	88.9875	
161		3550							3610.35	90.25875	
165		3550							3610.35	90.25875	
194		3500							3559.5	88.9875	
198		3500	4.15	2.85	17.6	0.153	0.099	35.294118	3559.5	88.9875	
202		3550							3610.35	90.25875	
207		3520							3579.84	89.496	
211		3500							3559.5	88.9875	
215		3520	4.24	2.26	30.13333	0.153	0.076	50.326797	3579.84	89.496	
219		3550							3610.35	90.25875	
223		3550							3610.35	90.25875	
227		3520							3579.84	89.496	
231		3520	3.99	2.4	24	0.153	0.077	49.673203	3579.84	89.496	
236		3520							3579.84	89.496	
240		3540							3600.18	90.0045	
244		3550							3610.35	90.25875	
248		3520	4.09	2.87	19.2	0.153	0.105	31.372549	3579.84	89.496	
256		3500							3559.5	88.9875	
264		3500							3559.5	88.9875	
314		3500	4.04	3.18	19.23077	0.125	0.091	27.2	3559.5	88.9875	
322		3500							3559.5	88.9875	
331		3500							3559.5	88.9875	
339		3500							3559.5	88.9875	
347		3500	4.1	3.36	7.692308	0.125	0.095	24	3559.5	88.9875	

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	Feed Absorbance, cm^{-1}	Permeate Abs, cm^{-1}	Abs Removal, cm^{-1}	Normalized Flow	Normalized Flux, L/m^{2}h	Comments
355		3500	4.74	3.87	30.52885	0.125	0.099	20.8	3559.5	88.9875	
363		3520							3579.84	89.496	
430		3500	4.94	2.78	34.19355	0.136	0.075	44.852941	3559.5	88.9875	
438		3520							3579.84	89.496	
446		3500	4.94	2.72	41.72043	0.136	0.071	47.794118	3559.5	88.9875	
455		3550							3610.35	90.25875	
463		3500	5.04	2.76	42.58065	0.136	0.07	48.529412	3559.5	88.9875	
512		3520	5.04	2.97	14.83871	0.136	0.077	43.382353	3579.84	89.496	
513	38	3500							3559.5	88.9875	
534		3500	4.32	2.76	39.58333	0.157	0.084	46.496815	3559.5	88.9875	
538		3500							3559.5	88.9875	
546		3520							3579.84	89.496	
550		3520							3579.84	89.496	
555		3500							3559.5	88.9875	
563		3500							3559.5	88.9875	
571		3490	4.88	5.12	0.947867	0.161	0.16	0.621118	3549.33	88.73325	coagulant finished
579		3490							3549.33	88.73325	coagulant finished
600		3500	4.32	4.02	10.67961	0.158	0.137	13.291139	3559.5	88.9875	coagulant finished
612		3480	4.32	4.02	12.13592	0.158	0.137	13.291139	3539.16	88.479	coagulant finished
621		3500							3559.5	88.9875	
629		3500							3559.5	88.9875	
637		2600							3203.2	80.08	temp=19
664		2600							3203.2	80.08	
667		2600	4.07	2.82	28.64583	0.139	0.086	38.129496	3203.2	80.08	
671		2550							3141.6	78.54	
675		2550							3141.6	78.54	
683		2550							3141.6	78.54	
692		2530							3116.96	77.924	

Time, hours	Feed Pressure, kPa	Permeate flow, L/h	Feed DOC, mg/L	Permeate DOC, mg/L	DOC removal, %	Feed Absorbance, cm^{-1}	Permeate Abs, cm^{-1}	Abs Removal, cm^{-1}	Normalized Flow	Normalized Flux, L/m^2h	Comments
700		2530	4.47	3.01	25	0.139	0.087	37.410072	3116.96	77.924	opened recirc valve 100%
708		2600							3203.2	80.08	
717		2550	4.47	2.99	16.66667	0.139	0.083	40.28777	3141.6	78.54	
725		2600							3203.2	80.08	
733		2600	4.47	3.02	26.5625	0.139	0.088	36.690647	3203.2	80.08	
738		2700							3326.4	83.16	
742		2690							3314.08	82.852	
750		2690	4.32	2.76	30.84577	0.159	0.085	46.540881	3314.08	82.852	
759		2500							3080	77	
792		2690							3314.08	82.852	
800		2800			20.57899				3449.6	86.24	

APPENDIX 13:

RO TREATMENT WITHOUT PRE-TREATMENT

Appendix 13.1: RO Treatment for operation of the standard BW30-2540 membrane without pre-treatment

Permeate Flow, L/h	Feed Flow, L/h	Days of Operation, d	Normalized Permeate Flow, L/h	Normalized Flux L/m^{2}h,	Feed TOC, mg/L	Permeate TOC, mg/L	Brine TOC, mg/L	% TOC rejection
90.00	890.00	0.00	90.00	34.62				
89.00	889.00	0.00	90.73	34.90				
82.00	887.00	0.00	82.74	31.82				
70.00	875.00	1.00	78.04	30.01				
72.00	877.00	1.00	75.21	28.93				
71.10	746.10	1.00	80.03	30.78	70.60	2.20	72.60	96.88
72.00	745.00	1.00	75.43	29.01				
72.00	748.00	1.00	81.23	31.24	74.20	1.77	76.10	97.61
75.80	750.80	2.00	98.48	37.88	71.80	2.86	72.20	96.02
52.00	674.00	2.00	71.26	27.41				
37.00	710.00	2.00	57.11	21.97				
52.80	713.80	3.00	73.67	28.33	59.50	2.10	61.90	96.47
49.20	678.20	3.00	51.31	19.73	62.70	1.95	66.80	96.89
48.00	664.00	3.00	51.64	19.86				
30.24	703.24	3.00	45.36	17.45	61.60	2.16	65.80	96.49
47.52	672.52	4.00	51.86	19.94				
47.52	676.52	4.00	71.06	27.33	64.70	2.41	66.90	96.28
81.00	803.00	5.00	86.36	33.21				
72.00	867.00	5.00	82.36	31.68				
59.40	863.40	6.00	68.30	26.27	56.00	3.88	59.20	93.07
55.50	860.50	7.00	71.77	27.60	26.00	5.81		77.65
54.79	858.79	8.00	58.37	22.45				
57.50	862.50	8.00	69.68	26.80	59.20	1.61	61.00	97.28
56.21	860.21	8.00	57.83	22.24				

73.89	880.89	9.00	68.99	26.54	58.60	2.48	63.90	95.77
55.06	859.06	9.00	57.26	22.02				
56.66	860.66	9.00	57.27	22.03				
59.67	863.67	9.00	66.76	25.68	63.30	2.10	64.10	96.68
58.59	862.59	9.00	68.27	26.26	61.90	2.20		96.44
50.64	852.64	10.00	60.94	23.44	63.80	1.99	66.00	96.88
60.44	862.44	10.00	68.29	26.27	63.60	2.71		95.74
61.59	865.59	10.00	70.05	26.94	63.70	2.71		95.75
55.84	859.84	11.00	59.69	22.96				
51.69	853.69	11.00	64.41	24.77	65.40	2.12	67.60	96.76
52.40	857.40	11.00	64.34	24.75	65.80	2.08		96.84
52.40	857.40	13.00	64.34	24.75				
53.69	855.69	15.00	61.40	23.61				
53.69	855.69	15.00	61.40	23.61				
63.00	865.00	17.00	69.17	26.61				
63.00	865.00	18.00	69.20	26.62				

Appendix 13.2: RO treatment for the low fouling BW30XFR-2540 membrane without pre-treatment

Permeate Flow, L/h	Feed Flow, L/h	Days of Operation, d	Normalized Permeate Flow, L/h	Normalized Flux, L/m^{2}h	Feed TOC, mg/L	Permeate TOC, mg/L	Brine TOC, mg/L	% TOC rejection
96	813		96.000	36.92				
90	810		95.758	36.83	103	15.6	101	84.85
88	817	1	97.855	37.64				
92	802	1	87.391	33.61	77.218155	7.13	86.3	90.77
89	803	1	83.176	31.99	90.946488	2.27	102	97.50
83	809	2	85.637	32.94	483	15.6	581	96.77
88	810	2	85.901	33.04	482	13.7	576	97.16
88	808	2	85.991	33.07				
78	810	3	85.274	32.80				
89	812	3	77.534	29.82	359	7.5	379	97.91
88	811	3	76.253	29.33				
92	808	3	87.091	33.50	393	48.8	362	87.58
90	806	3	85.410	32.85				
80	805	4	83.522	32.12	387	10.1	427	97.39
87	892	4	83.993	32.31				
91	890	4	76.403	29.39	83.2	4.92	89.9	94.09
102	910	4	83.111	31.97				
82	804	8	86.329	33.20	88.7	15.4	94.7	82.64
81	809	8	76.525	29.43				
95	809	8	91.451	35.17	96	0.673	94.9	99.30
97	808	8	78.664	30.26				
91	810	8	75.984	29.22				
90	813	8	89.001	34.23	94.8	13.3	85.7	85.97

72	822	8	85.781	32.99			77.9	
88	814	9	75.177	28.91				
109	807	9	87.933	33.82			95.8	
88	810	9	90.483	34.80	94.1	3.79	94.1	95.97
86	812	10	87.197	33.54	109	3.04	108	97.21
83	812	10	71.598	27.54				
83	803	10	72.252	27.79				
90	816	10	89.586	34.46	116	3.81	114	96.72
81	782	11	89.117	34.28	80.2	12.5	87.4	84.41
87	812	11	77.605	29.85	114	6.03	108	94.71
75	803	11	80.008	30.77	108	14.4	102	86.67
80	806	12	68.347	26.29				
90	882	12	77.121	29.66				
84	885	12	73.949	28.44	31.6	1.06	35.1	96.65
50	854	12	46.375	17.84	9.5	0.268	10.6	97.18
87	895	13	74.560	28.68	9.99	0.347	11.2	96.53
65	863	13	59.279	22.80	10.8	0.364	12	96.63
123	802	17	101	38.80	92.9	30.4		67.28
95	632	17	82	31.56	92	36.9		59.89
87	561	17	94	35.98	45.3	0.316		99.30
89	562	18	90	34.51	89.5	0.071		99.92
88	895	20	70	27.04	94.5	2.47		97.39
86	896	22	78	29.92	92.9	1.95		97.90
90	895	23	87	33.46	90.4	1.97		97.82
87	561	29	88	33.90	45.3	0.316		99.30
89	562	31	85	32.54	89.5	0.071		99.92
88	895	33	70	27.04	94.5	2.47		97.39
70	880	35	61	23.61	92.9	1.95		97.90
71	876	36	63	24.10	90.4	1.97		97.82
97	901	37	80	30.87				

APPENDIX 14:

RO TREATMENT WITH PRE-TREATMENT

Appendix 14.1: RO treatment for BW30-2540 standard membrane with pre-treatment

Permeate Flow, L/h	Feed Flow, L/h	Days of Operation, d	Normalized Permeate Flow, L/h	Normalized Flux, L/m^2h	Feed TOC, mg/L	Permeate TOC, mg/L	Brine TOC, mg/L	% TOC rejection
84.40	886.40	19.00	77.49	29.80	4.89	0.58	5.14	88.16
81.14	889.14	19.00	77.29	29.73	6.45	0.84	7.26	86.91
73.61	875.61	20.00	71.61	27.54				
82.15	894.15	20.00	75.66	29.10	8.67	1.03	9.55	88.12
79.74	1091.74	20.00	76.91	29.58				
76.40	1096.40	20.00	74.47	28.64	16.00	1.22	16.90	92.38
71.83	1086.83	21.00	76.63	29.47	16.00	1.05	16.80	93.44
64.88	1084.88	21.00	70.38	27.07	16.10	1.04	17.30	93.54
68.87	1088.87	22.00	72.05	27.71				
62.15	1080.15	22.00	61.99	23.84				
59.74	1069.74	23.00	65.01	25.00				
56.40	1070.40	24.00	56.39	21.69				

Appendix 14.2: RO treatment for the low fouling BW30XFR-2540 membrane with pre-treatment

Permeate Flow, L/h	Feed Flow, L/h	Days of Operation, d	Normalized Permeate Flow, L/h	Normalized Flux, L/m^{2}h	Feed TOC, mg/L	Permeate TOC, mg/L	Brine TOC, mg/L	% TOC rejection
75	880	37	66	25.38				
79	883	37	74	28.61	57.6	2.17	60.9	96.23
77	884	37	67	25.80				
75	882	38	73	28.08	55.1	0.873	59.8	98.42
83	890	38	69	26.45				
83	888	38	68	26.22				
81	886	38	74	28.52	61.1	1.36	64.3	97.77
70	872	40	71	27.40	63.7	1.2	66	98.12
74	875	40	63	24.05				
75	879	40	70	27.09	62.9	1.18	65.2	98.12
72	882	41	70	26.83	63.4	1.15	66.7	98.19
67	875	41	68	25.99	65.1	1.08	67.5	98.34
72	882	41	64	24.67				
71	876	42	68	26.20	63.4	1.15	66.7	98.19
71	876	42	67	25.73	64.6	1.22	69.3	98.11
71	879	42	59	22.58				
69	874	42	58	22.42				
67	874	42	58	22.15				
69	876	43	65	25.03	63.7	1.09	66.7	98.29
		46						
87	892	48	77	29.60	73.1	1.34		98.17
87	888	50	75	28.92				
87	899	50	76	29.16				
84	888	51	76	29.25				
82	890	51	75	29.04				
78	882	52	70	26.96				

APPENDIX 15:

REDUCTION OF ORGANICS ACROSS THE KOMATI DESALINATION PLANT (UV SCANS)

		wavelength, nm								
	sample point	**220**	**230**	**240**	**254**	**260**	**270**	**280**	**290**	**300**
UV absorbance, cm^{-1}	UF Feed 31 March 2011	2.6223		0.581	0.1472	0.1165	0.0785	0.0351	-0.0189	
	UF Permeate 31 March 2011	2.5272		0.5046	0.078	0.0518	0.0208	-0.0159	-0.063	
	RO1 Feed 31 March 2011	2.5196		0.4765	0.0547	0.029	0.0012	-0.0325		
	RO1 Permeate 31 March 2011	2.3716		0.3527	-0.0515	-0.0706				
	RO1 Brine 31March 2011	3.1392	2.4966	1.2399	0.7126	0.6484	0.548	0.4295	0.3054	0.2097
	RO2 Feed 31 March 2011	2.3805		0.3752	-0.0321	-0.052				
	RO2 Permeate 31 March 2011	2.3672		0.3733	-0.0332	-0.0537				
	RO2 Brine 31 March 2011	2.3606		0.3557	-0.0495	-0.0687				
		wavelength, nm								
	sample point	**220**	**230**	**240**	**254**	**260**	**270**	**280**	**290**	**300**
UV absorbance, cm^{-1}	UF Feed 05 April 2011	2.6015	1.6746	0.5818	0.1457	0.1148	0.0778	0.0331	-0.0203	-0.0578
	UF Permeate 05 April 2011	2.54	1.5804	0.5066	0.0809	0.0544	0.0231	-0.0142	-0.0615	
	RO1 Feed 05 April 2011	2.5561	1.5777	0.5236	0.095	0.0674	0.0353	-0.0038	-0.0524	
	RO1 Permeate 05 April 2011	2.3601	1.3889	0.3575	-0.0471	-0.0662	-0.0827			
	RO1 Brine 05 April 2011	3.1918	2.896	1.5499	0.9864	0.9065	0.7723	0.619	0.4621	0.3379
	RO2 Feed 05 April 2011	2.3728	1.3898	0.3543	-0.0497	-0.0685	-0.0844			
	RO2 Permeate 05 April 2011	2.3747	1.3938	0.3519	-0.0518	-0.0714	-0.0865			
	RO2 Brine 05 April 2011	2.3845	1.4151	0.3641	-0.0416	-0.0609				
		wavelength, nm								
	sample point	**220**	**230**	**240**	**254**	**260**	**270**	**280**	**290**	**300**
UV absorbance, cm^{-1}	UF Feed 12 April 2011	2.6515	1.6683	0.5946	0.1619	0.1372	0.0778	0.063	0.0008	-0.0478
	UF Permeate 12 April 2011	2.5051	1.5417	0.4748	0.058	0.0376	0.0157	-0.0185	-0.0682	
	RO1 Feed 12 April 2011	2.555	1.5777	0.4784	0.062	0.0418	0.0193	-0.0153	-0.0653	
	RO1 Permeate 12 April 2011	2.3698	1.3842	0.3529	-0.0487	-0.0664	-0.08			
	RO1 Brine 12 April 2011	3.1949	2.7556	1.2013	0.6981	0.6606	0.5913	0.4717	0.3212	0.1888
	RO2 Feed 12 April 2011	2.3692	1.378	0.3531	-0.05	-0.0683	-0.0837			
	RO2 Permeate 12 April 2011	2.3773	1.3915	0.3506	-0.0505	-0.069	-0.0836			
	RO2 Brine 12 April 2011	2.3849	1.4011	0.3546	-0.0453	-0.0612	-0.0732			

APPENDIX 16:

DATA USED TO DERIVE BREAKTHROUGH CURVES

Table 16A: DATA FOR DERIVATION OF BREAKTHROUGH CURVE - FILTRASORB F300

	Time, minutes	Time, hours	Column outlet TOC concentration, mg/L	Volume of water, Litres	Bed volumes, litres	%DOC removal
CW (initial TOC concentration = 48.2 mg/L	0	0		0	0	0
Sample 1	19.8	0.33	10.6	4500	8910	78.01
Sample 2	90	1.5	29.8	20250	40500	38.17
Sample 3	120	2	26.6	27000	54000	44.81
Sample 4	150	2.5	30	33750	67500	37.76
Sample 5	180	3	30.3	40500	81000	37.14
Sample 6	210	3.5	32.7	47250	94500	32.16
Sample 7	240	4	36.8	54000	108000	23.65
Sample 8	270	4.5	41.6	60750	121500	13.69
Sample 9	300	5	43	67500	135000	10.79
Sample 10	330	5.5	46	74250	148500	4.56
Sample 11	360	6	47.2	81000	162000	2.07

Table 16B: DATA FOR DERIVATION OF BREAKTHROUGH CURVE - FILTRASORB F400

Time, minutes	**Time, hours**	**Column outlet TOC concentration, mg/L**	**UV254, cm^{-1}**	**Volume, Litres**	**Bed volumes, Litres**	**% DOC removal**
0	0	0		0	0	0
60	1	15.7	0.305	13500	27000	69.80769
120	2	21.8	0.45	27000	54000	58.07692
180	3	26.8	0.557	40500	81000	48.46154
240	4	38.8	0.84	54000	108000	25.38462
270	4.5	37.6	0.815	60750	121500	27.69231
300	5	38.2	0.823	67500	135000	26.53846
330	5.5	40.5	0.873	74250	148500	22.11538
360	6	41.6	0.904	81000	162000	20
390	6.5	42.7	0.928	87750	175500	17.88462
420	7	42.9	0.935	94500	189000	17.5
450	7.5	44.4	0.965	101250	202500	14.61538
480	8	45.1	0.985	108000	216000	13.26923
510	8.5	44.7	0.986	114750	229500	14.03846
540	9	45.6	1.009	121500	243000	12.30769
570	9.5	45.6	1.012	128250	256500	12.30769
600	10	47.4	1.067	135000	270000	8.846154
630	10.5	42.5	0.957	141750	283500	18.26923
660	11	45.8	1.025	148500	297000	11.92308
690	11.5	49.2	1.1	155250	310500	5.384615
CW initial TOC	52 mg/L					

Table 16C: DATA FOR DERIVATION OF BREAKTHROUGH CURVE – ORGANIC SCAVENGER RESIN

Time, minutes	**TOC concentration, mg/L**	V (L itres)	Bed volumes	**% TOC removal**
0	0	0	0	0
10	3.56	2250	2250	94.32217
20	7.87	4500	4500	87.44817
30	8.49	6750	6750	86.45933
50	9.91	11250	11250	84.19458
60	8.97	13500	13500	85.69378
70	7.71	15750	15750	87.70335
80	11.6	18000	18000	81.4992
90	9.48	20250	20250	84.88038
100	11.8	22500	22500	81.18022
110	13.6	24750	24750	78.30941
120	14.1	27000	27000	77.51196
CW initial TOC	62.7			

APPENDIX 17:

PROPERTIES OF THE ORGANIC SCAVENGER RESIN

Table 17A: Properties of the organic scavenger resin

Item	Details
Matrix	Cross-linked acrylic macro reticular structure
Functional group structure	Quaternary ammonium
Total exchange capacity	≥0.8 q/L (Cl form)
Moisture-holding capacity	66 to 72% (Cl form)
Specific gravity	1.05 to 1.08 (Cl form)