MEng Research Dissertation

Desktop Study on Novel Treatment Techniques to Treat Industrial Fertilizer Effluent

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Abstract

Fertilizer production is a massive global industry with the global consumption of the three main fertilizer nutrients, nitrogen, phosphate and potassium estimated at 187 million tonnes in 2016 with an anticipated annual growth of approximately 2% for the foreseeable future. In 2016 the global fertilizer market was estimated to have an overall market value of 141 billion US dollars.

Fertilizer production produces significant liquid waste as process water used for the various separations, cleaning, emulsifying and dilution processes absorbs various nutrients and contaminants from these production processes. This liquid waste has characteristically high concentrations of nutrients derived from the base fertilizer, such as various dissolved phosphate compounds for phosphate-based fertilizer production or dissolved nitrogenous compounds for nitrogen based fertilizer production. These contaminants are inherently nutrients that could be recovered for beneficial re-use. The phosphate and potassium minerals used in fertilizer production are obtained from ores mined from the earth, thus the re-use of these mineral present particular significance when taking into accounting the declining global supply of these ores. Furthermore, if these liquid wastes are not disposed of correctly they can lead to detrimental environmental impacts such as eutrophication and ecological degradation in water courses.

This study addresses this problem by presenting three novel treatment techniques to treat the liquid waste produced from a fertilizer production plant. A liquid waste sample obtained from a particular fertilizer production plant producing primarily nitrogen-based fertilizer is used as a design basis to evaluate the three presented treatment techniques. The techniques are evaluated based on their economic feasibility, technical feasibility and resource recovery ability.

The three treatment techniques studied were the Sharon-Anammox bioreaction process, electrodialysis with struvite recovery process and combined forward-reverse osmosis process. The technical feasibility of the processes was primarily evaluated based on the effluent water quality from the treatment systems. The effluent quality index (EQI) was used as a comparative measure of the effluent quality of the processes.

All three processes were found to perform inadequately from a technical feasibility perspective as demonstrated by the negative EQI values obtained for the processes. The Sharon-Anammox bioreaction process was found to perform poorly because its application is limited to treatment of waste streams containing high ammonia concentrations such as in conventional domestic waste. Therefore, the Sharon-Anammox process was not suited to the fertilizer effluent which also contained high nitrates, phosphates and total dissolved solids.

The electrodialysis process performed poorly as it was unable to effectively remove the ammonium cations from the process water. The combined forward-reverse osmosis process performed poorly because a resource recovery step was not included to treat the concentrated waste stream discharged from the forward osmosis step of the process. It was identified that a similar struvite recovery step should be added to the combined forward-reverse osmosis process to improve the technical feasibility of the process and to provide the process with resource recovery capabilities.

From an economic feasibility perspective, it was found that the addition of the struvite recovery setup to the electrodialysis process increased the capital costs of the process to between 300% and 500% of the other two options. However, with the omission of the struvite recovery setup the capital costs of all three processes were in a similar range.

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Declaration

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Jordache R. Fortuin

Date: 10/08/2018

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LIST OF ABBREVIATIONS

COD:	Chemical oxygen demand
FS:	Feed solution
DS:	Draw solution
RO:	Reverse osmosis
FO:	Forward osmosis
NF:	Nanofiltration
TDS:	Total Dissolved Solids
CAN:	Calcium ammonium nitrate
FSA:	Free and saline ammonia
CAS:	Conventional activated sludge
CM:	Standard anion exchange membrane
AM:	Anionic exchange membrane
MVA:	Monovalent anion exchange membrane
TSS:	Total Suspended Solids

LIST OF NOMENCLATURE

Dolomite:	Calcium Magnesium Carbonate ($CaMg(CO_2)_3$)
Effluent:	Raw waste water directly from the fertilizer plant. Typically referring to the fertilizer waste water prior to the treatment plant.
Treated Effluent:	Effluent that has been treated. Typically at the discharge of the treatment plant.

1. INTRODUCTION

Effluent treatment is generally an expensive process, requiring energy intensive and often complex machinery. With global phenomena such as climate change, industrialization and rapid population growth the demand for water is expected to increase, while available fresh water is expected to decline (W. W. Li, Yu, & Rittman, 2015). Thus, it is necessary to develop more efficient water treatment processes in which we are able to minimize energy requirements and maximize resource recovery.

Industrial effluent treatment is often complex because effluent concentration profiles differ from industry to industry and often even from plant to plant. In this study, suitable waste water treatment techniques for treatment of fertilizer manufacturing plant effluent are reviewed. Data made available from a prominent South African fertilizer production facility has been used as an input for a design basis where necessary. The source of the data will remain confidential to comply with non-disclosure agreements.

Fertilizer production is a massive global industry with the global consumption of the three main fertilizer nutrients, nitrogen, phosphate and potassium as potash (K_2O) estimated at 186.67 million tonnes in 2016 and anticipated annual growth of approximately 2% for the foreseeable future (FAO, 2017). In 2016, the global fertilizer market was estimated to have an overall market value of 141 billion US dollars (Mordor_Intelligence, 2017).

Process water from fertilizer production plants are typically rich in dissolved nitrogen and phosphorus-based nutrients arising from contact with the respective mineral forms of these nutrients in the fertilizer production process. The process water obtained from fertilizer processes are ideal for stimulating cell growth as these nutrients are key components in the cell growth process. This is often seen detrimentally in natural ecosystems where waste water discharged into a natural water system leads to excessive plant and algae growth in a process referred to as eutrophication (Mishra, Nayak, Guru, & Rath, 2010).

The process water from the current fertilizer facility is currently discharged into a holding dam. A sample was taken from the effluent contained in the dam. The sample was analysed and found to contain high concentrations of nitrates, phosphates and sulphates and a relatively low pH as a result of the inorganic acids produced from the plant.

The purpose of this study is to present potential treatment options in treating effluent water from fertilizer production plants in general. Fertilizer industrial effluent has a specific concentration profile containing high nitrogen and phosphorus and low organic carbon, therefore, conventional activated sludge effluent treatment processes will not perform adequately. Treatment processes are required that focus on removal of the nitrogen and phosphorus-based nutrients in the absence of organic carbon. A further treatment requirement is the potential re-use of the recovered resources as potential commercial products. A range of treatment techniques were reviewed in various literature sources to find potential treatment techniques that would meet the above process requirements. From these literature sources three treatment techniques were chosen. The three chosen treatment techniques are listed below:

- Combined Sharon-Anammox bioreaction (van Dongen, Jetten, & van Loosdrecht, 2001).
- Electrodialysis with struvite precipitation (Zhang et al., 2013).
- Combined forward and reverse osmosis membrane filtration (Vallidares Linares et al., 2014).

2. LITERATURE REVIEW

2.1 Fertilizer production

2.1.1 Background

Fertilizers are substances added to soil to improve the health and growth of plants. In general, phosphorus, nitrogen and potassium containing compounds make up the primary nutrients in fertilizers as these nutrients are essential for intracellular functions and therefore plant wellbeing. Different ratios of these primary minerals are blended to make up specific recipes for particular vegetation requirements. Furthermore, minerals such as calcium, magnesium and sulphur are added in small amounts to fertilizers as well as micronutrients such as iron, chlorine, copper, manganese, zinc, molybdenum and boron. Fertilizers come in many forms and depending on application requirements it can be in liquid, solid or slurry form (Price, 2006).

2.1.2 Raw materials

The raw materials for fertilizer production can be obtained from natural resources such as sodium nitrates, seaweed, bones, guano, potash and various mineral rocks (limestone, phosphate rocks, dolomite, etc). Also, raw materials for fertilizer can be obtained via chemical synthesis using processes such as the Haber-Bosch and Ostwald processes (Silberberg, 2007). In particular, nitrogen containing compounds such as ammonia, nitric acid, ammonium nitrate and ammonium phosphate are synthesised using chemical processes. This is generally more cost effective than obtaining nitrogen from naturally occurring raw materials (Silberberg, 2007).

2.1.3 Calcium ammonium nitrate

One of the major fertilizers being produced at the site where the sample water was obtained from is calcium ammonium nitrate (CAN). Fertilizer grade CAN typically contains 8% calcium and between 21-27% nitrogen (Price, 2006). The raw materials for CAN include ammonium nitrate, dolomite, steam and cooling water (Chemicals_Technology, 2017).

The primary ingredient used in the CAN manufacturing process is molten ammonium nitrate produced from a 3-step chemical process as follows:

1. Haber-Bosch process: used to synthesize ammonia from atmospheric nitrogen and hydrogen obtained from natural gas, according to the following reaction (Silberberg, 2007):

$$N_2(g) + 3\,H_2(g) \leftrightharpoons 2NH_3(g) \qquad [1]$$

2. Ostwald process: used to synthesize nitric acid from ammonia via a three step reaction as follows (World_of_chemicals, 2017):

Step 2a: $4NH_3(g) + 5\,O_2(g) \leftrightharpoons 4NO(g) + 6\,H_2O(g)$ [2]

Step 2b: $2NO(g) + O_2(g) \leftrightharpoons 2NO_2(g)$ [3]

Step 2c: $3NO_2(g) + H_2O(l) \leftrightharpoons 2HNO_3(l) + NO(g)$ [4]

3. Acid-base reaction with nitric acid and ammonia to produce ammonium nitrate as follows (Silberberg, 2007):

$$HNO_3(l) + NH_3(g) \leftrightharpoons 2NH_4NO_3(l) \qquad [5]$$

For the specific plant where the sample was obtained, the above reactions are optimized to produce a liquefied molten solution containing approximately 85% ammonium nitrate. The molten ammonium nitrate is then further concentrated to between 98.5%-99.5% ammonium nitrate in a series of evaporation processes. The final concentrated melt is then mixed with dolomite in a homogenizing tank and then pumped to a fluidized drum granulator which is the first step in the granulation circuit. The granulated product obtained from the granulation circuit is then screened, cooled, coated and stored for packaging (Chemicals_Technology, 2017).

The evaporated process water from this process is a major component of the process water discharged to the site dams and will therefore form a major component in the effluent characteristics.

2.2 Methods for treating fertilizer effluent

2.2.1 Sharon-Anammox treatment

Anammox treatment is a biological treatment using various strands of bacteria that convert ammonium and nitrite to nitrogen gas. While the Anammox process has taken place in nature for centuries, it is a fairly new technology and not completely understood at present. However, it has incurred great interest in recent years because of its ability to remove nitrogen from waste water more efficiently and at a lower cost than conventional nitrification-denitrification systems, which is used to remove nitrogen in the majority of waste water treatment plants around the world (Paulsrud & Szatkowska, 2014).

Potential cost savings for the Anammox treatment system versus the conventional nitrification-denitrification systems is as a result of the following differences:

- Anammox treatment is an anaerobic process and therefore it does not require oxygen. Whereas for conventional nitrification-denitrification systems, huge amounts of oxygen is supplied for nitrifying ammonia to nitrate prior to denitrification in the anoxic reactor.
- Anammox bacteria are autotrophic and therefore they fix carbon from inorganic carbon dioxide from the atmosphere. Heterotrophic bacteria used in the denitrification of conventional waste water treatment systems utilize organic carbon as substrate from the waste water. Therefore, if the N-C ratio is not adequate, additional carbon must be dosed in the form of acetate, methanol, ethanol, etc. This has particular significance for certain industrial wastes such as landfill leachate, fertilizer plant effluent, fisheries and abattoirs that have characteristically high nitrogen contents and low carbon contents (Paulsrud & Szatkowska, 2014).
- The reaction mechanisms of the Anammox process results in a net removal of CO_2, unlike conventional nitrification-denitrification systems that result in a net production of CO_2. This will become increasingly significant with stricter carbon emissions taxes being placed on industries because of global pressure resulting from climate change.

2.2.1.1 Anammox microbiology

Anammox microbiology is fairly complicated and at present not very well understood. The Anammox bacteria share a number of properties with both eukaryotes and archaea. The cell structure of the bacteria are split into three separate compartments by bilayer membranes which consists of a cell wall, paryphoplasm, riboplasm and anammoxosome (Kartal, van Niftrik, Keltjens, Op den Camp, & Jetten, 2012). The Anammox reaction has been found to take place in the anammoxosome, which occupies most of the bacterial cell volume (Lindsay et al., 2001).

Currently 10 Anammox species have been identified. Known species can be divided into five genera: (1) Kuenenia, including one species, K. stuttgartiensis; (2) Brocadia, including three species, B. anammoxidans, B. fulgida and B. sinica; (3) Anammoxoglobus, which includes one species, A.

propionicus; (4) Jettenia, also including one species, J. asiatica; and (5) Scalindua, which has four known species, S. brodae, S. sorokinii, S. wagneri and S. profunda (Kartal et al., 2013). Phylogenetic analyses of these species places them all within the phylum *planctomycete.* All five of these genera share unique physiological and morphological features, with the key genera being the anammoxosome in order for an effective anommoxic reaction to take place (Paulsrud & Szatkowska, 2014).

2.2.1.2 Anammox reaction mechanisms

The overall Anammox reaction is given as follows (Paulsrud & Szatkowska, 2014):

$$NH_4{}^+ + NO_2{}^- \leftrightharpoons N_2 + 2H_2O \qquad [6]$$

Reaction 6 is a catabolic reaction used by the Anammox bacteria to generate energy for cell growth. Ammonium ($NH_4{}^+$) is the electron donor and nitrite ($NO_2{}^-$) the electron acceptor in the catabolic redox reaction. The energy generated in this reaction is used together with an inorganic carbon source (CO_2) to generate new bacteria cells in a process known as anabolism. The yield of Anammox bacteria are significantly less than that of heterotrophic bacteria in conventional activated sludge processes (Kartal et al., 2012).

Three half-reactions make up this overall reaction, these three reactions are as follows (Paulsrud & Szatkowska, 2014):

$$NO_2^- + 2H^+ + e^- \leftrightharpoons NO + H_2O \qquad [7]$$

$$NO + NH_4^+ + 2H^+ + 3e^- \leftrightharpoons N_2H_4 + H_2O \qquad [8]$$

$$N_2H_4 \leftrightharpoons N_2 + 4H^+ + 4e^- \qquad [9]$$

In reaction 7 nitrite is reduced to nitric oxide by nitrate reductase. Ammonium is then combined with nitric oxide by hydrazine hydrolase to form hydrazine in reaction 8. Hydrazine is then finally oxidized to nitrogen gas by hydrazine/hydroxylamine oxidoreductase in reaction 9 (Kartal et al., 2011). Szatkowska and Paulsrud (2014) report that these reactions occur within the anammoxosome, a specialized pseudo-organelle within the bacteria cell.

Strous et al. (1998) proposed the following overall cell synthesis stoichiometric reaction:

$$NH_4{}^+ + 1.32NO_2{}^- + 0.066HCO_3{}^- + 0.13H^+ \\ \leftrightharpoons 1.02N_2 + 0.26NO_3{}^- + 0.066CH_2O_{0.5}N_{0.15} + 2.03H_2O \qquad [10]$$

Of particular importance in the cell synthesis reaction is the nitrite to ammonia ratio (NAR). The above reaction proposed by Strous et al. (1998) has a nitrite to ammonia ratio of 1.32, which is in agreement with various stoichiometric quotients reported by other researchers. A value of 1.3 has gained general consensus as being acceptable amongst researchers (Paulsrud & Szatkowska, 2014).

2.2.1.3 Anammox bioreactor configurations and process performance

From the reactions presented in the previous section we can see that nitrite as well as ammonia is required for Anammox bacteria to synthesize and produce nitrogen gas. Ammonia is readily available in most waste waters and particularly domestic waste. However, nitrite seldom exists in high quantities in waste water. Hence, Anammox process configurations generally always include partial conversion of ammonia to nitrite which is achieved in aerobic conditions in the presence of ammonium oxidizing bacteria (Paulsrud & Szatkowska, 2014).

For the aerobic conditions required for nitritation an additional dissolved oxygen (DO) source is required in addition to oxygen diffusion from atmospheric oxygen. Anammox bacteria are obligate anaerobes. Therefore, the presence of DO inhibits the functioning of the Anammox bacteria (van Dongen et al., 2001). Thus, the DO concentration is a key control parameter in the Anammox process. The DO concentration in the nitritation reaction zone must be controlled, not only to maintain a DO range that results in effective oxidation of ammonia to nitrite but also to ensure that no DO is carried over into the anaerobic reaction zone. This is one of the major considerations in the process configuration for Anammox processes. There are currently many system configurations utilizing Anammox bacteria with the major differences being if the nitritation process and anaerobic process are undertaken in separate reactors and what type of growth medium is used for the bacteria. Granular sludge, activated sludge and biofilms have all been used as growth mediums in Anammox processes (Paulsrud & Szatkowska, 2014).

A few of the mainstream reactor configurations for the Anammox process are the DEMON, SHARON-Anammox, ANITA-mox and DeAm-mon processes (Paulsrud & Szatkowska, 2014).

The DEMON process is based on a suspended growth activated sludge process which takes place in a Sequencing Batch Reactor (SBR). Nitritation and Anammox occur simultaneously in different zones within the reactor. Aeration is controlled intermittently within a narrow pH range and low oxygen content (Wett, Nyhuis, Takacs, & Murthy, 2010; Wett et al., 2013). A hydro-cyclone is used to separate the heavier Anammox bacteria from the other nitrifying bacteria, to allow for preferential recycling of the Anammox bacteria back into the reactor (Wett et al., 2010).

The SHARON-Anammox process is a two-stage suspended growth process. The first stage is a SHARON (Single reactor system for High-Activity-Ammonium-Removal-Over-Nitrite) reactor. In the SHARON reactor control of temperature, pH, retention time and DO concentration allows for the preferential growth of nitrifying bacteria and minimization of nitrate formation. Van der Star, et al. (2007) report effective anammox reaction at a pH between 7-8 and temperature between 30-40°C . Ammonium is oxidized to nitrites in the Sharon reactor according to the following reaction:

$$NH_4^+ + 1.5O_2 \leftrightharpoons 2H^+ + NO_2^- + H_2O \qquad [11]$$

The SHARON reactor is operated as a completely mixed reactor to allow for uniform bio-reaction. From the SHARON reactor a nitrite enriched discharge stream proceeds to stage two which is the Anammox reactor. For the SHARON-Anammox process an up-flow solids granulation process or biofilm (Fixed or moving) bioreactor is used for the Anammox reactor in order to aid in retaining the slow growing Anammox bacteria (van der Star et al., 2007).

The ANITA-mox and DeAm-mon processes both utilize carrying media in order to produce biofilm growth on the protected surface area of the media. Thereby sheltering the Anammox bacteria within the media under the biofilm and effectively retaining Anammox bacteria in the system. Nitritation takes place in the outer biofilm layer while the anammox reaction takes place within the inner biomass (Paulsrud & Szatkowska, 2014).

2.2.2 Electrodialysis with Struvite Precipitation

Struvite is a phosphate-based mineral of the orthophosphate group. It is composed of magnesium, ammonium, orthophosphate and water. The chemical composition indicates that magnesium, ammonium and orthophosphate are required in a molar ratio of 1:1:1. These minerals are then surrounded by six water molecules to form $MgNH_4PO_4 \cdot 6H_2O$, according to the following reaction (Malanda, Randall, & von Blottnitz, 2016):

$$Mg^{2+} + NH_4^+ + HPO_4^{2-} + 6H_2O \rightleftharpoons MgNH_4PO_4.6H_2O \qquad [12]$$

Struvite is a stable white orthorhombic crystal and often encountered in waste water treatment plant anaerobic digesters, sludge treatment facilities or pipes. Struvite was initially encountered as an operational problem because of the scaling it caused in waste water treatment plant process equipment and tanks. Subsequently, it has been identified as an effective slow release fertilizer and much research has gone into struvite recovery as a resource; in particular struvite recovery from source-separated urine streams (Etter, Tilley, Khadka, & Udert, 2011).

Because of its stable solid form, struvite can be easily recovered from process water using a number of physical separation techniques. Given the general reliability of these physical separation techniques, the extent of struvite recovery is largely dependent on the degree of struvite formation from the process water. Factors influencing struvite formation are reviewed in the subsequent sections.

2.2.2.1 Process water composition

As indicated previously, struvite formation occurs with a 1:1:1 ratio of magnesium, phosphorus and nitrogen. Therefore, these minerals are required in equal molar proportions in the process water for a high degree of struvite formation. This is one of the limiting factors in source-separated urine streams which contain high concentrations of ammonia and orthophosphate but relatively low concentrations of magnesium. Therefore, magnesium is usually dosed into struvite precipitation reactors for source-separated urine streams.

A typical composition of human urine from a male is given by David Putnam (1971) to contain 9.3g/L urea, orthophosphate as phosphorus at 0.47g/L and magnesium as ionic magnesium at 0.02 g/L. Where urea is converted to ammonia and bicarbonate by Urease enzymes, for every mole of urea, two moles of ammonia are formed (Marsh, Sims, & Mulvaney, 2005). In comparison, effluent from fertilizer plants have differing compositions depending on what specific fertilizer is being manufactured. However, for the effluent obtained from the current fertilizer plant, the composition was found to contain 5.2 g/L ammonia, 0.11 g/L phosphate, and 0.38 g/L magnesium. This composition has a more favourable magnesium concentration compared to human urine, yet phosphate may be a limiting factor in the struvite formation for the fertilizer effluent sample.

Although the struvite compound contains a 1:1:1 molar ration of magnesium, ammonium and phosphate, in practice differing ratios of the compounds have been found to favour struvite formation. Krähenbühl (2016) found that the optimum molar ratio for struvite formation was 1.2:3:1, while Antonini (2011) found that in the presence of excess ammonium and using a Mg:P ratio of 1.5:1 a precipitation efficiency of 97% in terms of Phosphorus was achieved; where the precipitate was found to be comprised of 85% struvite.

2.2.2.2 Effect of pH on struvite formation

With the exception of the process water composition, pH has the strongest effect on struvite formation. This is because the speciation of both PO_4^{3-} and NH_4^+ is strongly affected by the pH of the system. A pH range between 7 and 11 has been found to be optimum for struvite precipitation (Doyle & Parsons, 2002). The effect of pH on carbonate, phosphate and ammonium speciation is reviewed below (Malanda et al., 2016):

Carbonate

The carbonate system is mediated by total inorganic carbon (in particular carbon dioxide) dissolving in water to form carbonic acid, bicarbonate and carbonate ions according to the overall equilibrium expression (Malanda et al., 2016):

$$CO_2(l) + H_2O \leftrightarrow H_2CO_3 \leftrightarrow H^+ + HCO_3^- \leftrightarrow H^+ + CO_3^{2-} \qquad [13]$$

At a pH range between 7 and 11 the dominant total inorganic carbon species is bicarbonate.

Phosphate

Aqueous phosphate speciates into four forms dependent on the system pH: phosphate (PO_4^{3-}), hydrogen phosphate (HPO_4^{2-}), dihydrogen phosphate ($H_2PO_4^-$) and trihydrogen phosphate or phosphoric acid (H_3PO_4). The dissociation of phosphoric acid into the related phosphate species occurs according to the following reaction (Malanda et al., 2016):

$$H_3PO_4 \leftrightarrow H^+ + H_2PO_4^- \leftrightarrow H^+ + HPO_3^{2-} \leftrightarrow H^+ + PO_4^{3-} \qquad [14]$$

With an increase in pH shifting the equilibrium progressively to the right of the above equation. At a pH range between 7 and 11 the dominant phosphate species is hydrogen phosphate (HPO_4^{2-}).

Ammonia

Inorganic nitrogen speciates between ammonia and ammonium dependant on the system pH according to the following equilibrium equation (Malanda et al., 2016):

$$NH_3 + H_2O \leftrightarrow NH_4^+ + OH^- \qquad [15]$$

At pH levels above 8 the equilibrium rapidly shifts to ammonium. At pH concentrations above 11 all inorganic nitrogen speciates into ammonium, and ammonia concentration is zero.

2.2.2.3 Effect of temperature on struvite formation

The various ions making up the struvite compound are held together by hydrogen bonds which can be easily broken at high temperatures. It has been identified that struvite is thermally unstable at temperatures above 50°C (Sarkar, 1991). When subjected to elevated temperatures for sustained durations, ammonia and water tend to evaporate from the struvite compound forming various dehydrated forms of struvite; including magnesium hydrogen phosphate trihydrate ($MgNPO_4 \cdot 3H_2O$) and magnesium hydrogen phosphate monohydrate or dittmarite ($MgNH_4PO_4 \cdot H_2O$) (Antonini et al., 2011).

2.2.2.4 Effect of competing ions on struvite formation

The presence of various ions influence the effectiveness of Struvite formation. In particular calcium, potassium, sodium, sulphate and carbonate compete to form various other complexes, such as the following examples: calcium phosphate, calcium carbonate and potassium phosphate. Malanda et al. (2016) found that a Ca:Mg ratio higher than 1:1 inhibits struvite formation and favours calcium phosphate formation.

2.2.2.5 Additional considerations for struvite formation

Furthermore, there are other more minor factors that can influence struvite formation and should be considered when designing a struvite precipitation reactor; these include (Antonini et al., 2011):

- Rate of mixing
- Composition of dosed chemicals ($MgCl_2$, MgO, $MgSO_4$)
- Seeding and supersaturation

2.2.2.6 Electrodialysis background

Generally speaking electrodialysis is a membrane process where ions are transported through semi-permeable membranes driven by an electric potential between the membranes (Lenntech, 2017). The semi-permeable membranes are either cationic or anionic selective, preferentially allowing either cations or anions to pass through the membrane. A number of membranes of differing ionic selectivity can be put in a row to achieve multiple ionic separations into different streams.

The semi-permeable membranes are prone to fouling and neutralization when in contact with certain chemicals such as large organic molecules, iron oxides and manganese oxides (Lenntech, 2017). Furthermore, colloids and particles, typically bigger than 10 µm, need to be removed prior to the waste stream being fed to the membrane arrangement as these will clog the pours of the membrane. Thus, pre-treatment of the waste water is often required prior to electrodialysis to ensure the absence of such particles and chemicals.

2.2.2.7 Combined struvite precipitation with electrodialysis

Zhang et al. (2013) conducted an electrodialysis experiment on anaerobic effluent water in order to concentrate phosphates for a feed into a struvite precipitation reactor. With the desired outcome that the concentrated phosphate stream would increase the efficiency of the struvite reactor. Zhang et al. (2013) made use of a three membrane stack: a standard cation exchange membrane (CM), Standard anion exchange membrane (AM) and a monovalent selective anion exchange membrane (MVA). Each membrane had an active surface area of 0.0064 m^2 and a spacer channel width of 1mm between each membrane to avoid possible blockage from precipitates. In total three cell trios were utilized, with each cell having the following membrane sequence CM-AM-MVA-CM. In total with all three cell trios four CM membranes, three AM and three MVA membranes were used with each cell having a feed, product and brine stream. The membranes used by Zhang et al. (2013) were obtained from PCD GmbH in Germany.

Using an anaerobic effluent stream as the feed stream into the electrodialysis arrangement and a struvite reactor effluent as the product stream where the struvite reactor effluent is circulated through the struvite reactor and electrodialysis setup, a concentration of 1.5 mmol/L (145 mg/L) phosphate was achieved after 14 hours of operation from an initial concentration of 0.93 mmol/L (90 mg/L). After 62 hours a concentration of 6.64 mmol/L (642 mg/L) was achieved. The experiment utilized a low current density of 31.25 A/m^2 with a current supply of 0.2A. With a 5 volt power supply and at an applied power of 1 kWh the experiment was found to produce 60g of Phosphate; resulting in a power to phosphate conversion of 60 g/kWh.

2.2.3 Combined forward and reverse osmosis (FO-RO)

2.2.3.1 Membrane filtration background

Filtration techniques such as nanofiltration (NF), reverse osmosis (RO) or forward osmosis (FO) could be used to recover dissolved ammonium and phosphate ions. These filtration techniques also use semi-permeable membranes to achieve ion separation. However, the driving force when using NF and RO is pressure rather than electrical potential or osmotic gradient as used for electrodialysis and FO, respectively.

For nanofiltration, or reverse osmosis, a high-pressure pump is used to pump the feed waste water into the membrane vessels and through the membrane. Depending on the pore size or selectivity of the membrane more pressure may be required. For a NF membrane the pore size of the membrane is in the nano range (1×10^{-7} to 1×10^{-9} m) while for a RO membrane the pore size is two orders of magnitudes smaller (1×10^{-9} to 1×10^{-11} m). Thus the pressure requirements for a RO plant is typically higher than that of a NF plant for the same flow rate and recovery (Coday et al., 2014).

Conversely, FO requires almost no external hydraulic pressure. Separation of water and dissolved solutes are driven by osmotic pressure gradient in FO treatment. Water is extracted from a lower osmotic pressure feed solution (FS) into a higher osmotic pressure draw solution (DS). The effectiveness of the FO process is driven by the osmotic pressure difference between the FS and DS selected on either side of the semi-permeable membrane (Vallidares Linares et al., 2014). The FO process results in concentration of the FS and dilution of the DS.

All membranes used in membrane filtration systems are prone to fouling from colloids and silt. However, the decline in flux for FO membranes is markedly lower than that of NF and RO membranes for the same feed water. This is because the feed water is not under pressure in the case of FO and any foulants on the FO membranes are as a result of natural flow diffusion. Experiments conducted by Valladares Linares et al. (2014) on FO-RO systems indicated that FO membranes showed an almost complete recovery after hydraulic cleaning; while no noticeable changes were observed for RO membranes. Furthermore, certain membranes are sensitive to typical chemicals used in waste water treatment; such as alum, ferric chloride and cationic polyelectrolytes (DOW, 2002). Thus, pre-treatment and post-treatment of waste water may be required when using membrane filtration techniques. The pre-treatment and post-treatment of the waste water generally requires the addition of certain chemicals to adjust the pH of the waste water; or to prevent scaling. These chemicals can inhibit resource recovery from the concentrated stream. For example, the addition of anti-scalants that is typically added to reverse osmosis systems, prevents precipitation in the brine stream. Therefore, if precipitation is required after filtration further chemical addition or evaporation of the brine stream is required to obtain precipitate.

2.2.3.2 Membrane filtration combination techniques in practice

Various treatment combinations incorporating either NF or RO have been utilized in past experiments for nitrogen and phosphorus recovery. Nunes and Peimmann (2001) used a gas permeable membrane operated under vacuum pressure to successfully recover ammonia gas from an acid solution. Kurama et al. (2002) were able to achieve a 96.9% ammonium recovery from an ammonium rich effluent stream using an RO system. Mondor et al. (2008) were able to produce a concentrated nitrogen stream containing 13 000 mgNH_3-N/L using an electro-dialysis and RO combination setup treating piggery effluent. Sengupta et al. (2015) report that permeate concentrations as low as 0.008 mg P/L have been achieved using RO systems for certain industrial effluents. Generally, high degrees of nitrogen and phosphorus recovery can be achieved by using membrane filtration systems.

2.2.3.3 FO-RO treatment of waste water

Combining RO with FO has potential cost saving and resource recovery benefits for nutrient rich industrial streams. Typically an indirect desalination process is used when configuring an FO-RO process for the primary objective of treating waste water (Vallidares Linares et al., 2014). In the indirect configuration a high salinity water such as sea water or brackish water is used as the DS and a waste water as the FS. The high salinity DS results in a net flow from the FS to the DS across the semi-permeable membrane. This results in dilution of the DS and concentration of the FS. The DS is then sent to a RO system where it is desalinated to produce clean product water. The dilution of the DS prior to feeding it to the RO system is desirable as it decreases the pumping energy required to overcome the osmotic pressure difference. Li et al. (2014) reported a dilution of the Total Dissolved Solids (TDS) of the sea water draw solution from 35 600 mg/L to 11 900 mg/L, which resulted in a corresponding drop in osmotic pressure of 17.43 bar. The drop in the pressure requirement of the RO system correspondingly resulted in a drop in specific energy consumption of more than 50%. Using an indirect desalination process is particularly desirable in water scarce regions and along coastal cities, where treatment of waste water has the valuable by-product of producing potable water.

The concentrated feed stream obtained from the FO-RO system can then be used for resource recovery. For example, when using treated effluent from a digester or domestic waste water treatment plant as the feed stream; the concentrated feed stream can be sent for further anaerobic digestion as the concentrated solution increases the efficacy of digestion and biogas production. There is little research conducted using dilute fertilizer process water as a feed steam for an FO process. Though, there is potential to concentrate this stream; resulting in elevated concentrations of nitrates and phosphate in the discharged feed stream. The concentrated feed stream is then primed for resource recovery by precipitation techniques or as direct use as a liquid fertilizer. Valladares Linares et al. (2014) report complete rejection of phosphate and moderate rejection of nitrates by FO membranes; furthermore near complete rejection of a range of metals was detected (Pb, Zn, Cu, Cd).

An alternative treatment setup may also be utilized for RO-FO treatment where concentrated fertilizer is used as the DS and a waste water or sea water is used as the FS. The diluted fertilizer discharged from the FO-RO process can then be directly applied as liquid fertilizer. This configuration finds specific application in fertilizer irrigation in water scarce agricultural regions. Preliminary tests conducted by Phuntsho et al. (2011) show that 1 kg of commercial fertilizer can extract 11-29 L of fresh water from sea water.

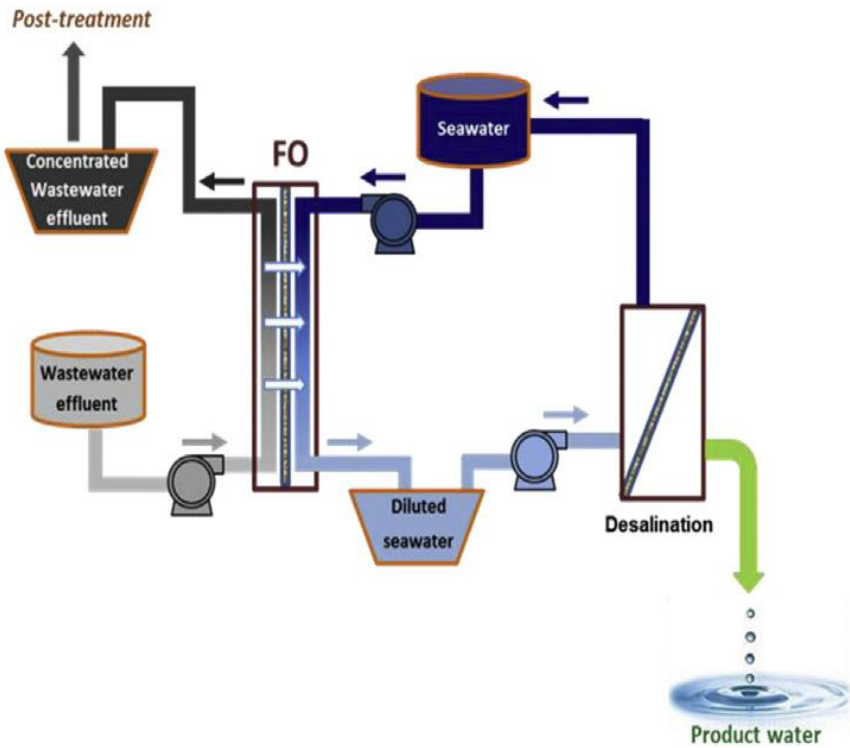


Figure 1: Layout of indirect FO -RO process (Z. Li et al., 2014)

3. STUDY PROBLEM

Water is ubiquitous in the everyday functioning of human life. It is difficult to find a part of the human life where water, in some form or another, is not involved. Water demands are increasing steadily with the rapidly growing world population and increasing industrialization of many countries. This presents a problem as water is often left contaminated after it is utilized in most processes. Usage of water in industrial processes presents a particular problem as the contaminants left in the water varies from industry to industry, resulting in the method of water treatment varying on a case by case basis.

This study looks in particular at the fertilizer production industry. Amongst others, the major contaminants encountered in the fertilizer industry are inorganic nitrogen (ammonia and nitrates), inorganic phosphates, potassium, magnesium and calcium. However, these contaminants, as indicated by the industry, are inherently nutrients. Therefore, the potential exists to utilize treatment of this waste water as a resource recovery process.

Various methods have been presented for utilizing treatment of nitrogen and phosphorus rich waste waters such as fertilizer for resource recovery. These resources include mineral rich precipitates (struvite, calcium phosphate, potassium phosphate, etc), combustible gases (methane and hydrogen) and clean reusable water.

This study aimed to present an overview of three novel treatment techniques for nitrogen and phosphorus rich waste waters, which included an economic analysis, technical feasibility and resource recovery comparison. All three of these criteria are important in the overall evaluation of a water treatment system, however the primary function of a waste water treatment system is to treat water and produce effluent quality within the legislative guidelines. Therefore, the technical performance indicated by effluent quality has been used as the priority performance criteria.

4. STUDY OBJECTIVES

The primary objective of this study was to provide a critical evaluation of three novel treatment techniques for treatment of industrial effluent from a fertilizer producing plant. These treatment techniques were primarily evaluated according to the following criteria:

- Overall technical feasibility with an emphasis on effluent water quality
- Economic feasibility
- Ability to recover resources

5. METHODOLOGY

5.1 Overall study methodology

In order to achieve the study objectives as outlined in section 4 the following methodology was utilized:

1. Appropriate novel treatment techniques were selected based on a wide range of existing technologies that could be potentially used to treat fertilizer effluents. The treatment techniques were selected based on their relative novelty and reported treatment capabilities. The three chosen treatment techniques are as follows:
    a. Sharon-Anammox treatment
    b. Struvite precipitation with electrodialysis
    c. Combined forward and reverse osmosis
2. A design basis was selected in order to evaluate the treatment techniques according to the criteria identified in section 4.
3. Process flow diagrams and process descriptions where generated for each treatment technique to provide a high-level understanding of what processes and equipment are involved in the treatment techniques.
4. Based on the design basis, mass balances were generated for the treatment techniques. The mass balances were used in the technical feasibility and performance evaluation by applying an effluent quality index to the treated water quality. The various physical separations and chemical reactions involved in the mass balance were derived from data and information obtained from literature, previous experimental setups and operational plants.
5. The treatment techniques' ability to recover resources was also evaluated by studying the mass balances derived from the design basis.
6. The treatment processes were then evaluated on a cost basis by determining CAPEX and OPEX costs for the processes. CAPEX costs were determined by using available equipment and infrastructure costs obtained from suppliers. Where real costs were not available costs were obtained from literature. OPEX costs were determined by using available costs for consumables, utilities and labour. Operational costs from literature were used in cases where real costs were not available.

All mass balances used in this study were based on reactions and data obtained from literature. No experimental validations were performed, because the scope of the research work only included a desktop study. Experimental validations are advised for future research.

5.2 Water quality and volume design basis

A water quality design basis was chosen on which to evaluate the three treatment options. The water quality data for the design basis influent was selected based on actual data obtained from a fertilizer production plant. The key water quality values for the process effluent are provided in table 1, where the values presented are average values of a time series spanning from the 15th of January 2015 to the 26th of May 2015. In addition to the historical data obtained from the fertilizer manufacturer, a grab sample was taken from the site and analysed at UCT's water lab. The grab sample analysis data was used to obtain water quality values that were not included in the historical data series. The water quality data for the historical series and the grab sample can be viewed in Appendix A-5

Table 1: Process effluent water quality for fertilizer production plant in South Africa

Parameter	Minimum Concentration	Maximum Concentration	Average Concentration	Unit
Electrical Conductivity	171	2 390	1 790	mS/m
Total Dissolved Solids*	-	-	8 940	mg/L
Ammonium (NH_4-N)	4 130	6 740	5 190	mg/L
Nitrate (NO_3-N)	4 990	11 500	7 250	mg/L
Nitrite (NO_2-N)**	-	-	0.49	mg/L
Phosphorus (P)	127	586	397	mg/L
Phosphate (PO_4-P)**	-	-	114	mg/L
Potassium (K)	230	365	325	mg/L
Zinc (Zn)	5	10	7.95	mg/L
Sulphur (S)	322	564	402	mg/L
Sulphate (SO_4)**	-	-	98	mg/L
pH	2.2	5.08	3.53	
Chemical Oxygen Demand (COD)**	-	-	193	mg/L
Volatile Fatty Acids (VFA)	-	-	0	mg/L
Calcium (Ca)**	-	-	936	mg/L
Magnesium (Mg)**	-	-	376	mg/L
Alkalinity as $CaCO_3$ ***	-	-	50	mg/L

*Value calculated based on EC, according to the following calculation: EC (in mS/m) ×10/2 ≈ TDS (in mg/L.
**No minimum or maximum values are available as these analyses were obtained from a grab sample analysed in the UCT water lab, as opposed to the historical data obtained from the fertilizer manufacturer. The grab sample report can be viewed in appendix A-5. The historical data obtained from the fertilizer manufacturer can also be viewed in appendix A-5.
*** Alkalinity value assumed.

An effluent volumetric flow rate was also required as part of the design basis as this forms an integral part in equipment and reactor sizing for the various treatment processes. A flow rate of **100 m^3/d** was used for process calculations and sizing purposes of the various treatment options.

5.3 Effluent quality performance index

The quality of the treated water discharged from a treatment plant is the primary selection criteria when evaluating an effluent treatment system. In order to more clearly define the performance of the treatment processes according to treated water quality an effluent quality index (EQI) was applied to the various treatment processes. De Ketele et al. (2017) define an EQI quantitatively according to the following equation:

$$EQI = \frac{1}{T.1000}\int_{to}^{tend}\left[\beta_{COD}\left(Limit_{COD} - COD_{(t)}\right) + \beta_{FSA}\left(Limit_{FSA} - FSA_{(t)}\right) + \beta_{OP}\left(Limit_{OP} - OP_{(t)}\right) + \beta_{NO}\left(Limit_{NO} - NO_{(t)}\right) + \beta_{TSS}\left(Limit_{TSS} - TSS_{(t)}\right)\right] Q_{e(t)}dt \quad [16]$$

Where, EQI = Effluent quality index (kg pollution/d)

T = the time step over which the EQI is evaluated (d),

Qe = the effluent volumetric flow rate (m^3/d),

$Limit_x$ = regulation concentration limit for discharge into a water source,

$X_{(t)}$ = concentration of contaminant of concern in effluent,

($X_{(t)}$= $COD_{(t)}$ for COD, $X_{(t)}$= FSA $_{(t)}$ for ammonia, $X_{(t)}$= $NO_{(t)}$ for total nitrates and nitrites, $X_{(t)}$= $TSS_{(t)}$ for total suspended solids, $X_{(t)}$= OP $_{(t)}$ for orthophosphate.)

β = Contaminant weighting factor

The beta factor is a weighting factor used to determine the weight of each pollutant in contrast to COD which is used as a reference point. The beta factor of COD is thus 1, the beta factors of the other contaminants are calculated as follows (De Ketele et al., 2017):

$$\beta_X = \frac{COD_{limit}}{limit_x} \qquad [17]$$

Where $limit_x$ is the regulation limit for the specific contaminant of concern. The regulation limits for the various contaminants of concern can be found in legislation documented by the relevant national water department. For this study, the regulation limits for discharge of effluent stipulated by the South African Department of Water Affair's general authorisations (as defined in terms of section 39 of the national water act-Act No.36 of 1998) were used.

The EQI values of each contaminant can be calculated individually by splitting equation 16 according to the various contaminants. For example, the EQI value for only COD can be calculated as follows:

$$EQI = \frac{1}{T.1000}\int_{to}^{tend}\left[\,\beta_{COD}\left(Limit_{COD} - COD_{(t)}\right)\right] Q_{e(t)}dt \qquad [18]$$

The EQI values calculated for the specific contaminants can then be split into two parts, EQI_{pos} and EQI_{neg}. If the EQI value of a specific contaminant is positive, then this is added to EQI_{pos} value and vice versa. A negative EQI value indicates that the pollutant exceeds the regulation concentration limit and a positive EQI value indicates that the pollutant falls within the regulation concentration limit.

Once the EQI values of all the contaminants are determined, the total of the negative values can be summed to determine the overall EQI_{neg} value and the total positive values can be summed to determine the over EQI_{pos} value. If the cumulative EQI_{neg} value is zero, this indicates no pollutants exceed the regulatory limit. The lower the EQI_{neg} value the worse the effluent quality and the higher EQI_{pos} value the better the effluent quality. The overall EQI value can then be easily determined by summing EQI_{neg} and EQI_{pos}.

A maximum EQI value can be determined by setting the value of all contaminants of concern equal to zero. The maximum EQI value is an ideal case and is unlikely considering all contaminants need to have a zero value, however the closer the overall EQI value is to the maximum EQI value, the better the quality of the water.

In this study, COD and suspended solids treatment was not a focus point as the waste water contained high concentrations of dissolved contaminants, low COD and negligible suspended solids. The treatment techniques were selected accordingly in order to remove the dissolved contaminants from the waste water. Thus, equation 16 was modified by removing the COD and suspended solids components and adding an additional dissolved contaminants measurement in the form of electrical conductivity. The beta factors and regulation concentration limits of the various contaminants are summarized in table 2. Where the beta values were calculated according to equation 17 by dividing the regulation COD limit (30 mg/L COD) by the regulation limit of the specific contaminant of concern indicated as C_{max} in table 2. The EQI has been applied to the treated water quality of the selected treatment techniques in this study in order to provide a quantitative measure of the technical performance of the techniques.

Table 2: Contaminant beta values and concentration limits

Parameter	Beta	C_{max}*	Unit
COD	1	30	mg/L
FSA	30	1	mg/L
OP	30	1	mg/L
NO	20	1.5	mg/L
EC	0.6	50	mS/m

*Max concentrations obtained from revision of general authorisations in terms of section 39 of the national water act, 1998 (Act No.36 of 1998)

5.4 Reactor Sizing

5.4.1 Sharon-Anammox reactor sizing

The Sharon and Anammox reactors were sized according to the following equation:

$$Vx = \frac{\mathrm{Qe}}{Rs} \qquad [19]$$

Where, V_x = reactor volume,

Q_e = the effluent volumetric flow rate (m^3/d),

R_S = reactor residence time.

As experimental or simulation data was unavailable for the required residence times, values obtained from various literature sources in which experiments were done on the processes were used. The literature values for the residence times are provided in Table 3.

5.4.2 Struvite precipitator reactor sizing

The precipitator reactor for the electrodialysis with struvite precipitation process was sized using a first order rate equation assuming the reactor operates as a plug flow reactor. The first order rate equation is given as follows (Sikosana, Randall, & Von Blottnitz, 2017):

$$\partial V = \frac{Q}{k(1-Y)}\partial Y \qquad [20]$$

Where V = reactor volume,

Q = volumetric flow rate (m^3/hr),

Y = conversion,

k = reaction constant.

Sikosana et al. (2017) used a reaction rate of 7.9 hr^{-1} to determine the required reaction volume for a struvite precipitator using the above equation. It is assumed that a value of 7.9 hr^{-1} will provide sufficient accuracy for the purposes of sizing the reactor for this study.

6. DISCUSSION

6.1 Option A: the Sharon-Anammox process

6.1.1 Overview

Option A proposed using a Sharon-Anammox process to treat the fertilizer waste water. The Sharon-Anammox process is a two-step reaction process. The first step includes partial oxidation of ammonia to nitrite in an aerated reactor. The second step includes anoxic conversion of ammonia and nitrite to nitrogen gas. The Sharon-Anammox process is a biologically based process that utilizes autotrophic nitrifying organisms for ammonia oxidation in the first reactor and autotrophic Anammox bacteria in the second reactor (van der Star et al., 2007).

6.1.2 Design basis

The conceptual design for option A considered treating 100 m^3/d of fertilizer plant effluent as outlined in the methodology section (section 5). Furthermore, option A considered treating all the fertilizer plant effluent through a two-stage Sharon-Anammox bioreactor process.

The key design criteria for option A are summarized in Table 3 below.

Table 3: Option A: Key Design Parameters

Description	Unit	Value	Comment/source
Influent volumetric flow	m^3/d	100	
Influent water quality		As per table 1	
Hydraulic retention time of Sharon Reactor	hr	24	(van Dongen et al., 2001)
Required ammonium to nitrate ratio to Anammox reactor		1: 1.3	(Strous et al., 1998)
Oxygen transfer efficiency		0.25	(Tchobanoglous, Burton, & Stensel, 2003)
Hydraulic retention time of Anammox reactor	hr	5	(Strous et al., 1998)
Anammox reactor type		MBBR	(van Dongen et al., 2001)

6.1.3 Process description

Figure 2 displays the overall process design of the Sharon-Anammox process. The process and related equipment displayed in Figure 2 has been generated for the purposes of this conceptual design. The combined Sharon-Anammox process primarily constitutes two separate reactor tanks as displayed in Figure 2. The Sharon reactor is the first reactor in the process. Raw influent is fed directly into the SHARON reactor where it is mixed and aerated. In the Sharon reactor aeration and mixing can be achieved by a number of methods. These include: mechanical mixing and aeration by impeller type mixers or aerators, air sparging or injection by pipe manifolds and aeration by diffuser systems to name a few. A fine bubble diffuser system was assumed to provide sufficient aeration and mixing in the SHARON reactor for this study. A pH correction dosing system was included in the process as indicated in Figure 2. The pH given in the design basis in Table 1 is 3.53, this was too low for effective growth of ammonium oxidising bacteria. According to Van Dongen et al. (2001) ammonium oxidation will not take place at pH levels below 6.5. Caustic soda (NaOH) was used as a pH correction media for this study. Caustic was fed by means of a dosing pump into the incoming raw influent stream prior to entering the SHARON reactor. The key operational parameters for the Sharon reactor is pH (7-8), temperature (30-40°C), dissolved oxygen (1.5 – 2 mg/L) and retention time. For design purposes, it was assumed that pH, temperature and dissolved oxygen in the Sharon reactor would be monitored using in-situ on-line probes. A retention time of 24 hours was used for sizing of the Sharon reactor. A retention time of 24 hours was found by Van Dongen et al. (2001) to provide sufficient time for growth of ammonium oxidizers while still being low enough to ensure sufficient wash out of the nitrite oxidizing organisms. Wash-out of the nitrite oxidizing organisms is important to ensure a sufficiently high concentration of nitrite and to prevent conversion of the nitrites to nitrates. The Sharon reactor was operated as a continuous stirred tank reactor with no sludge retention. The dissolved oxygen in the reactor was to be kept at 1.5 mg/L. The Sharon reactor was to be operated in order to obtain an ammonium to nitrite ratio of 1:1.3, as this is the optimum ratio for the Anammox bioreactor. The nitrite to ammonium ratio in the Sharon reactor can be controlled by adjusting the pH and retention time in the reactor.

For the Anammox reactor, two key design considerations are: (1) mixing ability and (2) suspended solids retention. Efficient mixing is important because the incoming nitrites have to be evenly distributed throughout the reactor to ensure uniform Anammox bioreaction. It is also important that as low solids retention as possible is achieved in the Anammox reactor as even a small retention of solids may lead to strong volume reduction of the Anammox sludge. There are a number of possible reactor configurations that could be used for the Anammox reactor, such as a membrane bio-reactor, packed bed biofilm reactor, fluidized bed reactor and moving bed biofilm reactor to name a few (Paulsrud & Szatkowska, 2014). A moving bed biofilm reactor was found to meet the above two operational criteria well (mixing ability and suspended solids retention) by Van Dongen et al. (2001); and was used for design purposes in this study. A moving bed biofilm reactor (MBBR) consists of a host biofilm media typically fabricated from polyethylene onto which the bacteria grow and create a protected biofilm on the internals of the polyethylene host media. The MBBR biofilm host media is effective in maintaining the Anammox bacteria within the reactor provided that during start-up the Anammox bacteria are given sufficient time to build up their biomass within the host media internals. Then during operation, the host material functions as a shield to the incoming flow and suspended solids; thereby allowing the Anammox bacteria to be maintained in the system. Following the start-up phase, the MBBR is operated with a low residence time so that the incoming flow is sufficient to maintain the biofilm media in suspension. A residence time of between 4-5 hours is suggested for the Anammox reactor by Van Dongen et al. (2001). Key operational parameters for the Anammox reactor are dissolved oxygen, pH, temperature, oxidation reduction potential (ORP), nitrite, nitrate and ammonia concentrations. For design purposes, it was assumed that these operational parameters were monitored by means of in-situ on-line probes in the reactor. The probes may be used for control, by manipulating process variables such caustic dosing, aeration and retention time upstream. For process control a nitrite concentration of 10 mg/L to the Anammox reactor is recommended (van Dongen et al., 2001).

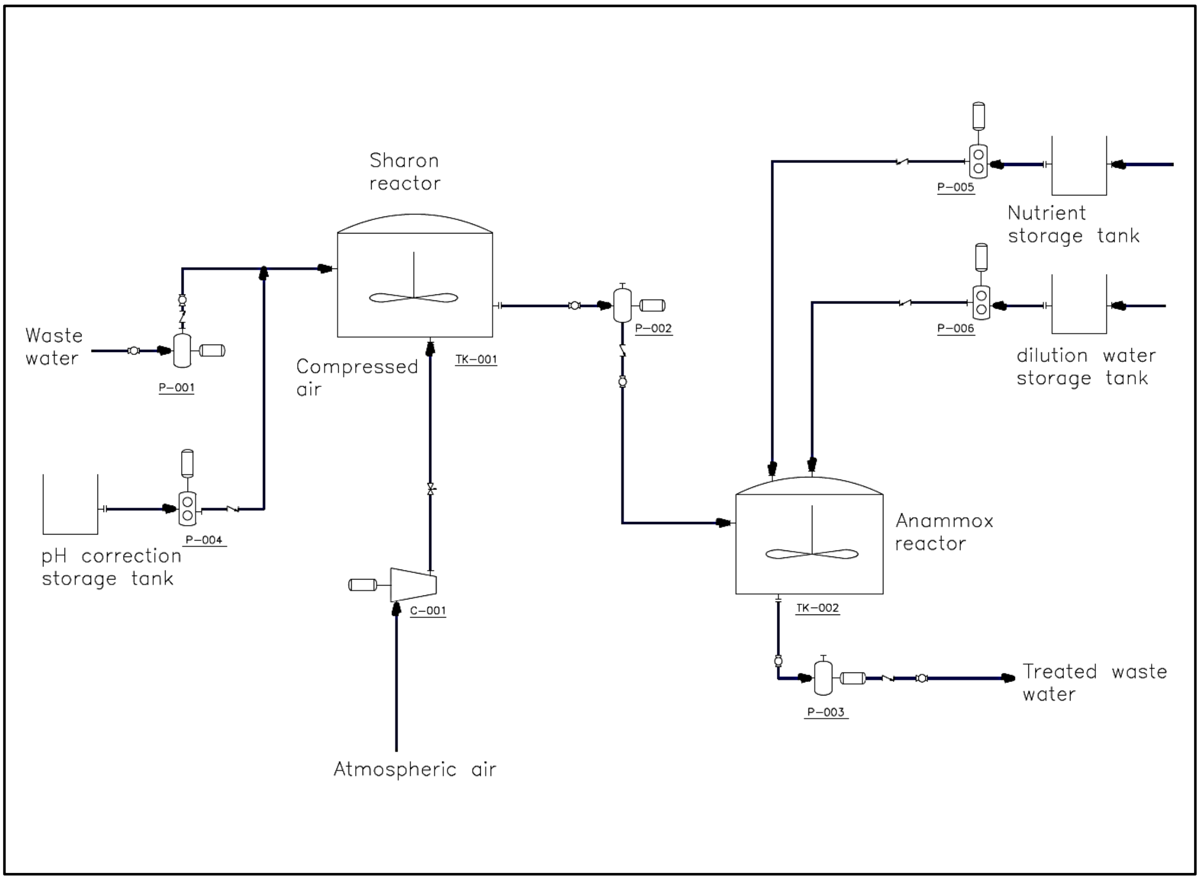


Figure 2: Sharon-Anammox Process Flow Diagram

6.1.4 Technical evaluation

6.1.4.1 Mass and energy flows

The optimum operational temperature for the Sharon reactor is 35°C to favour growth of ammonium oxidisers over the growth of nitrite oxidizers (van Dongen et al., 2001). However, a large portion of the fertilizer effluent plants process water is involved in cooling processes. This water would typically have to undergo cooling prior to being discharged. Thus, it was assumed that the cooling of the effluent water could be limited in order to provide an incoming feed stream to the Sharon reactor that is around 35°C. Therefore, no external heating processes were included in the Sharon reactor or Anammox reactor design. Hence, the main energy users consisted of the various centrifugal pumps used for transfer of the waste water and air supply blowers.

On the actual fertilizer plant the effluent is fed directly to a storage dam. For design purposes it was assumed that centrifugal pumps transferred the effluent from the dam to the Sharon reactor. Sodium hydroxide was fed directly into the pipeline feeding the Sharon reactor using diaphragm type dosing pumps in order to raise the pH to 7.

Furthermore, for design purposes it was assumed that air was supplied to the Sharon reactor by means of a centrifugal air blower in order to maintain the dissolved oxygen concentration at 1.5 mg/L and to provide the oxygen requirements for partial oxidation of ammonia to nitrite at a 1:1.3 ratio.

Partially oxidized effluent was then transferred from the Sharon reactor to the Anammox reactor by means of a centrifugal pump at a rate that coincided with the required residence times indicated in Table 3.

Sodium carbonate or soda ash was fed into the Anammox reactor in order to meet the carbonate requirements for the ammonium to nitrogen gas reaction according to equation 10. A make-up water stream was also included to the Anammox reactor, for chemical dilution or cleaning as required. The additional make-up water was not included in the mass balance as it was not expected to be a continuous operational variable.

Denitrified effluent was transferred from the Anammox reactor for further processing in a conventional waste water treatment plant. The various mass flows outlined here are summarised in Table 4. The detailed mass balance breakdown can be viewed in Appendix A-1.

Table 4: Summary of mass flows

Description	**Unit**	**Inputs**	**Outputs**
Incoming streams			
Influent waste water	kg/d	100 000	
Air for ammonium oxidation	kg/d	782	
Sodium hydroxide Solution	kg/d	2.40	
Sodium Carbonate	kg/d	137	
Exiting Streams			
Effluent	kg/d		100 700
Nitrogen gas	kg/d		339
TOTAL	**kg/d**	**101 000**	**101 000**

*All values rounded off to 3 significant figures.
** Refer to appendix A-1 for mass balance breakdown.

The energy requirements for operating the system is summarized in Table 5. Refer to Appendix B-2 for a breakdown of the various equipment and their related power requirements.

Table 5: Summary of energy demand for option A equipment

Description	**Unit**	**Value**
Feed pump	kW	1.1
Caustic dosing pump	kW	0.2
Transfer pump	kW	1.1
Soda Ash dosing pump	kW	0.2
Treated effluent discharge pump	kW	1.1
Blowers	kW	3
Process instruments and control panel	kW	1
TOTAL	**kW**	**7.7**
TOTAL*	kWh/d	92
TOTAL	kWh/year	33 726

*Electrical consumption based on drawing maximum power requirement for 12 h/d.
** Refer to appendix B-2 for breakdown of the various equipment and their related power requirements.

6.1.4.2 Spatial requirements

Based on the residence times and flows presented in Table 3 and using Equation 19 the required reactor volumes were calculated. The reactor volumes are summarized in Table 6. A reactor depth of 2.5m was used as the air diffuser supplier for the aeration system recommends a back pressure of at least 2mbar (refer to Appendix D-6). Using these reactor volumes with a reactor depth of 2.5m the reactor sizes were determined. The footprint for the reactors and other equipment housing is summarized in Table 7.

Table 6: Reactor sizing

Description	**unit**	**Value**
Sharon reactor volume	m^3	100
Anammox reactor volume	m^3	20.8
Reactor depths	m	2.50

Table 7: Option A spatial requirements

Description	**Footprint (m^2)**
Sharon reactor area	40.0
Anammox reactor area	8.30
Control Room	30.0
Lab/Chemical Storage room	30.0
TOTAL ESTIMATED FOOTPRINT	108

*All values rounded off to 3 significant figures.

Taking into account the surface areas required for the reactors in Table 7 and including an area of 30 m^2 for both the on-site control room and lab facilities it was calculated that 108 m^2 of free space would be required. However, this did not take into account dead spaces, pathways and other miscellaneous spatial requirements, so a safety factor of 2 was used for the overall spatial requirement; resulting in an overall surface area requirement of 216 m^2.

6.1.4.3 Overall performance and operation

The Sharon-Anammox process is an effective ammonia removal treatment process. Based on the mass balances provided in appendix A-1, an ammonia removal of 97% was achieved and an overall Nitrogen removal of 22% was achieved. The overall nitrogen removal of the process was fairly low because the nitrate concentration is higher than the ammonia concentration in the incoming effluent.

Because the Anammox process uses nitrite rather than nitrates for oxidation it is not an ideal process for nitrate removal, thus, the majority of the nitrates pass through the system and significantly decrease the overall nitrogen performance of the process. This illustrates that the Sharon-Anammox process is well suited to effluents containing high concentrations of ammonia and correspondingly low concentration of nitrates, which is typically the case for domestic waste water.

The water quality analyses of the fertilizer effluent presented in this report has a nitrate to ammonia ratio of approximately 1.4 in the incoming effluent as reported in Table 1. It is not clear whether the high concentration of nitrates will significantly affect the activity of the Anammox bacteria. Experiments conducted by Strous et al. (1998) exposed the Anammox bacteria to nitrate concentration as high as 70 mmol/L (which corresponds to 980 mgNO_3-N/L) and experienced no reduction in Specific Anammox Activity (SAA). Nevertheless, the conventional Sharon-Anammox process setup does not make provision for nitrate removal. The primary prerequisites for nitrate removal via a standard denitrification process are an organic carbon source (methanol, ethanol, etc), a heterotrophic biomass, stable pH and an anoxic environment.

A denitrification system can be retrofitted to the standard Sharon-Anammox design presented in this report by operating the Sharon reactor with intermittent aeration. However, this may present potential downstream issues such as methanol toxicity of the Anammox bacteria, sulphide toxicity of Anammox bacteria and loss of Anammox bacteria cell volume in the Anammox reactor. Other potential methods to incorporate denitrification may be to combine the Sharon-Anammox process treating fertilizer effluent with a conventional activated sludge (CAS) plant treating domestic effluent. The nitrate rich stream exiting the Anammox reactor can be sent to the anoxic reactor of a CAS plant and combined with the CAS plant's recycle. Alternatively, the fertilizer stream could be fed to an anoxic reactor in a CAS plant from which the discharge stream could be settled and the settled denitrified effluent could be sent to the Sharon-Anammox reactor. Nevertheless, the Sharon-Anammox treatment system is not ideally suited to treat nitrates in its standard process design. This is a significant negative factor when defining its technical performance.

As stated previously, the Sharon-Anammox process is primarily a nitrogen removing process, as such there is little to no phosphate removal. As can be seen in reaction 10, the Anammox bacteria have no phosphorus in their biomass, thus there is no net removal of dissolved phosphates via cellular assimilation by the Anammox bacteria. The only possible means of phosphate removal in the Sharon-Anammox process is by precipitation with dosed minerals (Sodium, calcium). However, since this would occur as a side reaction, the overall phosphate removal performance of the process was considered low.

Anammox bacteria are a very slow growing bacteria, Strous et al. (1998) indicate that the maximum growth rate of Anammox bacteria is 0.0027 h^{-1} with a doubling time of at least 11 d. Thus, the enrichment time required to reach steady state has been recorded to be as long as 1250 d (van der Star et al., 2007). Some of this time can be saved by seeding the reactor with biomass obtained from an operational facility. Nonetheless, the long start-up times should form part of the performance evaluation of the process.

6.1.5 Economic evaluation

6.1.5.1 Overview of economic evaluation

The options were evaluated in terms of economic potential by considering Capital expenditure (CAPEX), operating costs (OPEX) and revenues from product sales.

CAPEX cost was determined by obtaining actual equipment and construction costs where possible. Where actual costs were not available, best guess estimations obtained from literature sources were used to derive costs. The bill of rates for construction costs as well as supplier quotations for equipment costs can be found in Appendix B-1, B-2 and B-3 of this study.

From a civil engineering and construction perspective, for the purpose of determining the capital costs estimates for the Sharon and Anammox reactors, it was assumed that the reactor structures will be constructed from water retaining concrete. Further detailed assumptions and design considerations taken into account when costing the reactors can be viewed in the bill of quantities provided in Appendix B.

In order to provide capital cost estimates for the mechanical and electro-process installation it was assumed that no provision needed to be made for a heating installation, because the influent water coming from the fertilizer plant would be at the required temperature (35°C) as a result of heating processes in the fertilizer production facility. Furthermore, it was assumed that no dedicated mixing apparatus would be required in the reactors because the fine bubble aeration would provide sufficient mixing in the Sharon reactor; and the short hydraulic retention time and high depth-to-diameter ratio of the Anammox reactor would allow for adequate mixing in this reactor. Further equipment inclusions and design considerations taken into account for the mechanical and electrical costing can be view in the tables and data sheets provided in Appendix B.

6.1.5.2 CAPEX

The best cost estimates for the treatment system was derived by reviewing the various engineering aspects of the system design. Where possible real costs were used for engineering equipment and infrastructure. The reader is advised to review the cost calculations provided in appendix B for a clear breakdown of which costs were obtained from real values derived from actual quotations.

Table 8: Cost summary for option A: Sharon-Anammox

DESCRIPTION	**COST**
Civil works	R9 085 000
Mechanical equipment	R481 000
Electrical equipment	R 789 000
Sub-total Asset Capital Costs	R10 355 000
Consultant Professional Fees % of Capital Costs	22%
Professional Fees Costs	R2 279 000
TOTAL:	**R12 634 000**

*Refer to appendix B for detailed cost breakdown
** Rates as specified by *"Guideline scope of services and tariff of fees for persons registered in terms of engineering profession act, 2000 (Act No. 46 of 2000).*

The professional fees costs were derived from the ECSA guidelines of services and tariff of fees for persons registered in terms of engineering profession act (Act No. 46 of 2000). As per this act a fee of 8% of the overall costs of the works is required for civil engineering consulting on projects involving civil engineering where the costs of the works exceed R 4 300 000. A fee of 7% of the overall costs of the works is required for electrical engineering consulting on projects involving electrical engineering where the costs of the works exceeds R 8 060 000. Furthermore, as per this act a fee of 7% of the overall costs of the works is required for mechanical engineering consulting on projects involving mechanical engineering where the costs of the works exceed R 10 740 000.

6.1.5.3 OPEX and revenues

The estimated primary costs for operation of the Sharon-Anammox plant are summarized in table 9.

Table 9: OPEX costs rates for option A: Sharon-Anammox

Description	Unit	Value	Source
Price of Sodium Hydroxide (caustic soda)	R/kg	7.64	Protea Chemicals
Price of Sodium Carbonate (soda ash)	R/kg	4.35	Alibaba.com
Price of electricity	R/kWh	2.61	City of Cape Town- Electricity generation and distribution tariffs 2017/2018
Operator Wages	R/d	R500	Assumed
Water	R/kL	R57.00	City of Cape Town Water and Sanitation tariffs 2017/2018. (Fixed costs for commercial/industrial tariffs, refer to appendix C-8).
Maintenance costs	% Asset Capital costs per annum	5%	(Sikosana et al., 2015)

There are no readily available resources to recover for revenue generation from the Sharon-Anammox process. Consequently, no revenue projections were included for the Sharon-Anammox process.

The exact municipal water requirements for operation of the plant was not known. However, for purposes of the cost estimation it was estimated that 1 000 L of municipal water would be used per day; which led to an annual usage of 365 kL. Based on this water usage assumption and the data provided in Table 4 and Table 5 the estimated annual operational costs for the Sharon-Anammox process was calculated, the values are given in Table 10.

Table 10: Operational cost calculations for option A: Sharon-Anammox

Description	Annual Quantity	Rate	Total Annual Costs
Sodium Hydroxide	876 kg	7.64 R/kg	R7 000
Sodium Carbonate	50 005 kg	4.35 R/kg	R218 000
Electricity	33 726 kWh*	2.61 R/kWh	R88 000
Operator Wages	365 d	500 R/d	R183 000
Water	365 Kl	57 R/kL	R21 000
maintenance costs		5% of total asset costs per annum	R518 000
Total Annual Operational Costs:			**R1 035 000**

*Electrical consumption based on drawing maximum power requirement for 12 hours a day.

The capital and operational costs represented in this section are further reviewed in section 6.4. The costs are also compared alongside the costs of treatment options B and C to provide a frame of reference for the various costs.

6.1.6 Resource recovery

The Sharon-Anammox treatment system has been primarily developed as an alternative treatment method for nitrogen removal in waste water. Therefore, there is no significant resource recovery built into the conventional Sharon-Anammox process.

The Sharon-Anammox process is rather more focused on more efficient usage of raw materials in effluent treatment compared to conventional nitrogen removal processes. In a steady-state Sharon-Anammox process, no additional chemical substrate (e.g. methanol would usually be required as carbon source for denitrification for facultative heterotrophs) is required for ammonia removal; because the bacteria involved in both the nitrification process (Sharon reactor) and Anammox processes are autotrophic. Furthermore, less oxygen is required in an Anammox reactor to partially oxidize ammonia to nitrites compared to a conventional activated sludge reactor where complete ammonia oxidation to nitrate is required. Lastly, because the Anammox bacteria have such high cell maintenance requirements and low yields, the volumetric requirements for their biomass is a lot less than that required for conventional treatment using heterotrophic bacteria. Thus, the Sharon-Anammox process has a relatively small footprint compared to other sludge based effluent treatment techniques. The smaller footprint of Sharon-Anammox processes may lead to significant cost savings for the reactor infrastructure.

The Anammox process finds particular significance in cases where current effluent treatment plants are not meeting their nitrogen discharge requirements. The Anammox process can be retrofitted to a current effluent plant to receive the treated effluent containing the high nitrogen concentration because it does not require a carbon source and it has a relatively small footprint.

Overall the resource recovery potential of the Sharon-Anammox process is limited. The only potential resource that can be directly recovered from the process is nitrogen gas. However, the financial motivation for nitrogen recovery from the process is questionable as nitrogen gas is readily available from the atmosphere and is more easily harvested directly from air. Thus, no resource recovery has been included in the Sharon-Anammox process and subsequently no revenue product in the cost evaluation.

6.2 Option B: Electrodialysis with struvite precipitation

6.2.1 Overview

Option B proposed using a combined struvite precipitation and electrodialysis process to treat the fertilizer effluent. The combined process utilized electrodialysis to preferentially concentrate the ions required for struvite precipitation, namely magnesium, ammonia and orthophosphate. A struvite precipitation reactor with a slow mixer was utilized to aid struvite formation from a supersaturated solution. Struvite has been found to be an effective slow release fertilizer and strong cases have been presented for sustainable struvite harvesting from waste water for sale as fertilizer (Sengupta et al., 2015).

6.2.2 Design basis

The conceptual design for option B considered treating 100 m^3/d of fertilizer plant effluent as outlined in the methodology section (section 5). Furthermore, option B considered treating all the fertilizer plant effluent through a combined electrodialysis and struvite precipitation process. The key design criteria for option B are summarized in Table 11.

Table 11: Electrodialysis with struvite precipitation key design parameters

Description	Unit	Value	Comment/source
Influent volumetric flow	m^3/d	100	
Influent water quality		As per table 1	
Electrodialysis stack specifications		As per table 12	
Electrodialysis stack Ionic separation efficiency	%	80	(Zhang et al., 2013)
Electrodialysis current density	$A{\cdot}m^{-2}$	31.25	(Zhang et al., 2013)
Electrical potential difference across electrodialysis stack	V	12	(Altmeier, 2018)
Temperature	°C	25	Ambient
Struvite precipitator conversion	%	90	(Sikosana et al., 2015)
Struvite reactor first order reaction constant	hr^{-1}	7.9	(Sikosana et al., 2016)
Struvite reactor pH		8	
Struvite reactor type		Conical Fluidised bed	(Ostara, 2018)
Moisture content of filtered struvite	g/g (water to dry solids)	1.5	(Sikosana et al., 2017)

The electrodialysis membrane stack data sheet and specifications were obtained from the supplier, the key parameters are summarized in Table 12.

Table 12: Electrodialysis Membrane Stack Specifications

Description	Value	Unit
Electrodialysis stack manufacturer and Model	PCCell GMBH – PCCell ED 1000A	
Total Membranes per unit	100	
Total Cells Pairs per Unit	50	
Nominal flow through cell	10	L/hr
Maximum diluate flow per stack	500	L/hr
Total Effective membrane area per unit	10	m^2
Total amount of ED1000A electrodialysis stacks required	9	

6.2.3 Process description

As described in section 2.2.2 and equation 12, struvite precipitation takes place according to the following reaction:

$$Mg^{2+} + NH_4{}^{+} + HPO_4{}^{2-} + 6H_2O \rightleftharpoons MgNH_4PO_4 \cdot 6H_2O \qquad [12]$$

Combining an electrodialysis process with struvite precipitation allows for the preferential selection of ions in the product stream of the electrodialysis stack in order to favour the stoichiometry of reaction 12. An electrodialysis stack typically consists of various alternating charged membranes as demonstrated in Figure 3. For preferential selection of ions for struvite precipitation Zhang et al. (2013) used a stack with repeating membrane clusters containing 4 membranes per cluster in the following order: standard cationic exchange membrane (CM) – standard anion exchange membrane (AM) - monovalent anion exchange membrane (MVA) – standard cationic exchange membrane (CM). However, to avoid unnecessary complexity a standard cationic membrane (CM) and anionic exchange membrane (AM) cell pair was used in this study. Trial tests done by Patrick Altmeir of PCCell GMBH using the ED1000A electrodialysis stack in a continuous desalination process showed ionic separation efficiencies in excess of 80% with sufficient residence times (Altmeier, 2018).

The stack used in the conceptual design of option B is illustrated in Figure 3. The electrodialysis setup allowed for preferential concentration of phosphate, which suits the fertilizer effluent characteristics in this study (reported in Table 1) where the waste water contains a Mg: NH_4^+: HPO_4^{2-} molar ratio of 13: 240: 1. As highlighted in section 2.2.2.1 a Mg: NH_4^+: HPO_4^{2-} molar ratio of 1.2:3:1 will provide optimum struvite precipitation. The electrodialysis stack was used to concentrate the magnesium and ammonium cations in the diluate stream and the phosphate anions in the product stream, resulting in a more desirable molar ratio of Mg: NH_4^+: PO_4^{3-} in the struvite precipitation reactor.

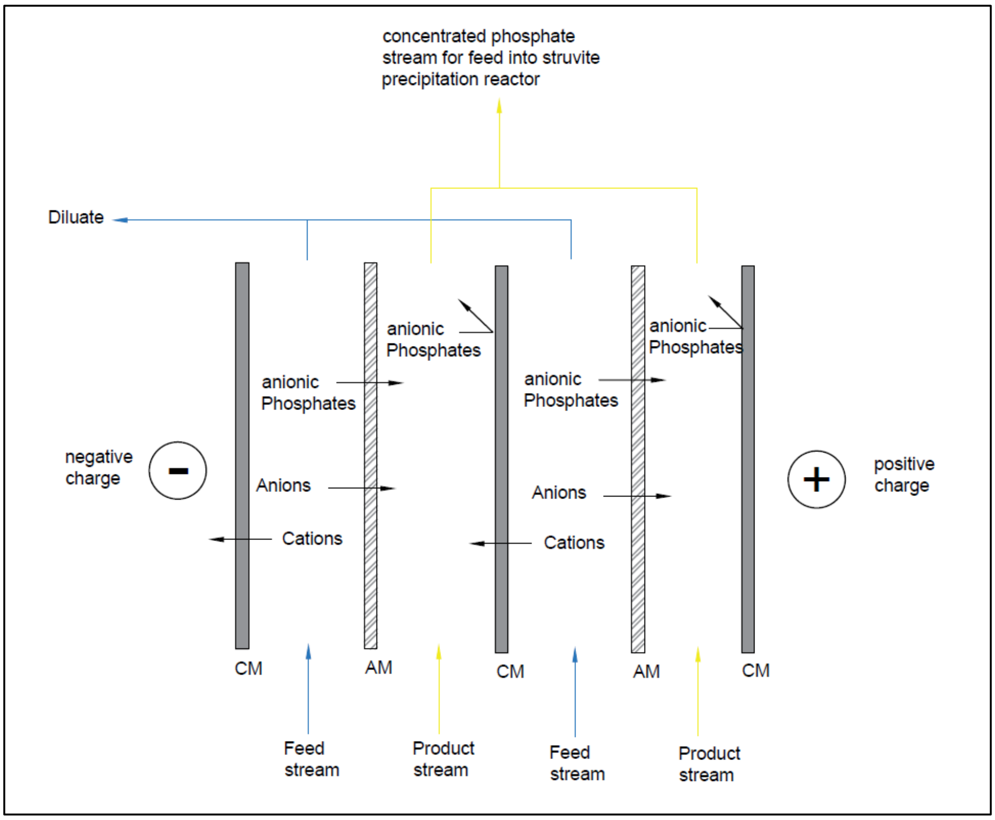


Figure 3: Electrodialysis ion separation schematic (Zhang et al., 2013)

The process setup as presented in Figure 4, demonstrates that the raw waste water was first fed to a pre-treatment setup and storage tank. Pre-treatment consisted of coarse filtration in order to remove suspended solids that may cause the electrodialysis membranes to foul prematurely. The pre-treated feed stream was then fed to the electrodialysis stack between the cation exchange and anionic exchange membranes. Anions, such as phosphates, nitrate, chlorides and sulphates permeate through the anionic exchange membrane into the product stream. The two streams that exit the electrodialysis stack are: 1. the diluted feed stream (or diluate) and 2. The phosphate concentrated product stream (or product stream). The diluted feed stream may then be recycled back to the feed water storage or reused as treated water if sufficiently diluted. An anti-scalant dosing system was included in the feed to the electrodialysis stack in order to prevent scaling of the ion exchange membranes.

Key control variables for the electrodialysis stack were the feed stream flow rates, residence times, pH of the product stream and current density. A nominal flow of 10 L/hr per cell is recommended by PCCell in the technical data sheet provided for the electrodialysis stack (Altmeier, 2018). It was assumed that at a flow rate of 10 L/hr per cell there was sufficient residence time inside the cell to achieve 80% ionic separation provided the current density was adequate. A current is applied across the membranes in order to create ionic flow between the compartments, it is important that the current flow remains constant to ensure efficient ion transportation. A 5V DC power supply was used by Zhang et al. (2013) to apply a constant current density of 31.25 $A{\cdot}m^{-2}$ (0.2A) across the membrane stack, for this experiment they were able to achieve a 95% desalination rate. For this study it was assumed that a current density of 31.25 $A{\cdot}m^{-2}$ with a 5V power supply would also be adequate.

From the electrodialysis stack the phosphate rich product stream was fed to the struvite precipitation reactor. Sodium hydroxide was fed into the product stream prior to feeding it into the struvite reactor to raise the pH above 8, which is favourable for struvite production. No additional magnesium was fed into the product stream as is commonly done in other struvite reactor setups. No additional magnesium was dosed because a mass balance for the process demonstrates that magnesium ions were in excess in the product stream. The struvite precipitator was assumed to be a conical fluidized bed reactor of similar design to the OSTARA Pearl fluidized bed reactor (Ostara, 2018). Struvite granules form inside the fluidized bed and once they have grown to the desired size they are removed. The treated effluent was discharged from the top of the reactor and recycled back as a feed to the electrodialysis stack. The removed struvite granules were then conveyed to a dewatering sieve and dried with air supplied by a blower. The dried granules were then available for packaging as fertilizer. The overall process schematic is provided in Figure 4 below.

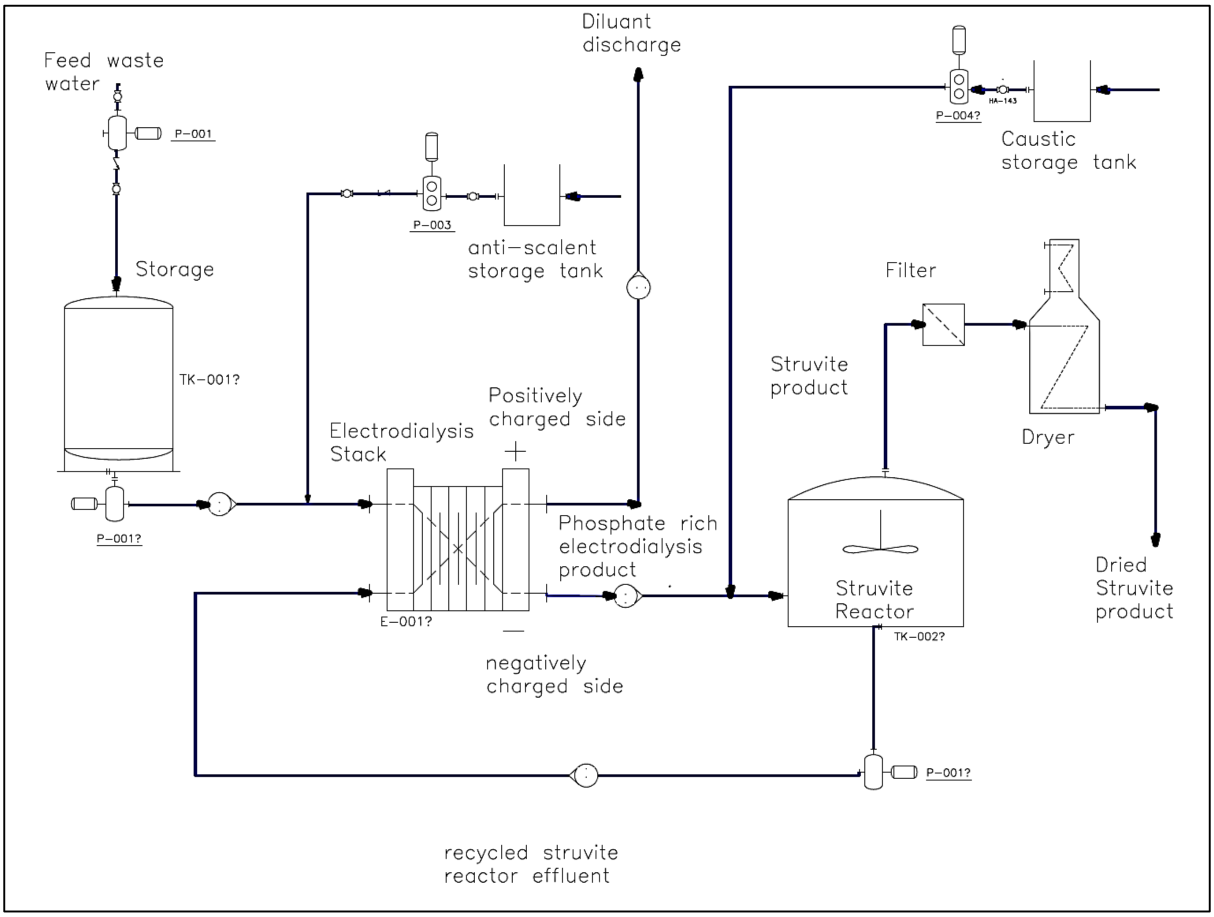


Figure 4: Combined electrodialysis and struvite precipitation process flow diagram

6.2.4 Technical evaluation

6.2.4.1 Mass and energy flows

The primary process streams for the electrodialysis-struvite precipitation system included the raw waste water feed stream, the diluate stream, the phosphate enriched product stream, the struvite recycle stream and the struvite product stream exiting the struvite precipitator. The process streams are displayed in Figure 4. Mass balances for overall flow, phosphate, ammonia and magnesium were performed over all these streams. The summarized overall incoming and exiting streams are given in Table 13.

Additional minor process streams include the anti-scalant dosing and sodium hydroxide dosing. The exact anti-scalant dosing requirement can be determined by a simulation using simulation software obtained from the anti-scalant supplier, however for the purposes of this report a feed stream concentration of 2 ppm has been deemed adequate, as recommended by Genesys anti-scalant manufacturers (Genesys_International, 2018). The exact sodium hydroxide dosing requirements can only be accurately determined with a titration or through modelling. For the purpose of this study, it was assumed that sodium hydroxide had to be dosed to obtain equal molar portions of hydroxide ions to hydrogen ions to reach a pH of 7.

Table 13: Option B: Mass balance summary

Description	Unit	Inputs	Outputs
Primary Incoming streams			
Phosphate mass flow in feed stream	kg/d	11.4	
Ammonia mass flow in feed stream	kg/d	519	
magnesium mass flow in feed stream	kg/d	37.6	
Primary exciting streams			
Phosphate mass flow in diluate stream	kg/d		1.14
Ammonia mass flow in diluate stream	kg/d		517
magnesium mass flow in diluate stream	kg/d		35.7
Phosphate mass flow in product stream	kg/d		10.3
Ammonia mass flow in product stream	kg/d		1.84
magnesium mass flow in product stream	kg/d		2.63
TOTAL	**Kg/d**	**568**	**569**

The mass flows in the feed stream were determined by multiplying the incoming volumetric flow by the feed stream concentrations listed in Table 1. The diluate and product stream concentrations were determined using the ionic separation efficiencies listed in Table 11. The mass separations into the diluate and product streams are demonstrated in Appendix A-2.

Table 14: Option B product recovery and dosing requirements

Description	Unit	Value
Product recovery		
Struvite production wet (40 wt/wt% dry solids)	kg/d	66.3
Packaged struvite (92 wt/wt% dry solids)	kg/d	28.8
Struvite production dry	kg/d	26.5
Chemical Dosing		
Anti-scalant	L/d	2
Sodium hydroxide	L/d	2.4

The 40 wt/wt% struvite production listed in Table 14 was obtained by considering the product mass flows exiting the electrodialysis stack enters the struvite precipitator where a struvite conversion efficiency of 90% has been used (Sikosana et al., 2017). The struvite precipitator mass balance is represented in Appendix A-2.

As displayed in Table 14 the mass flows in the feed stream were determined by multiplying the incoming volumetric flow by the feed stream concentrations listed in Table 1. The diluate and product stream concentrations were determined using the ionic separation efficiencies listed in Table 11. The mass separations into the diluate and product streams are demonstrated in Appendix A-2.

The electrodialysis-struvite precipitation setup produced 66.30 kg/d of wet product which further had to be dried to 92 wt/wt% dry solids to be suitable for sale as high grade commercial product (Sikosana et al., 2016). The wet product was dried using a belt conveyor and air blower arrangement.

Table 15: Option B energy requirements

Description	**Unit**	**Value**
Feed pump	kW	1.1
Transfer pump	kW	1.1
Recycle pump	kW	1.1
Anti-scalant dosing pump	kW	0.2
Caustic dosing pump	kW	0.2
Blowers	kW	3
Heater	kW	4.7
Conveyors	kW	0.2
Electrodialysis battery system	kW	14.1
Process instruments and control panel	kW	1
Estimated Total Power Requirement	**kW**	**26.7**
TOTAL*	kWh/d	319
TOTAL	kWh/year	116 000

*Electrical consumption based on drawing maximum power requirement for 12 hours a day.

The primary energy user for the process was the electrodialysis current system. The electrodialysis current system assumes a current density of 31.25 $A \cdot m^{-2}$ using a 5V power supply to generate sufficient current to drive the ionic separation process.

Another significant energy user was the drying system of the process which consisted of the blower and heater system. A 3 kW blower was selected for the drying system which was capable of delivering 250 m^3/hr of air at 120 mbar. The energy requirements for the heating system was estimated based on the heating requirements presented by Sikosana et al. (2017), in which it was reported that 18 kW of energy was required for the heating system to treat 1170 kg of wet product (40 wt/wt %). The current system treats 66 kg of wet product daily, thus a quarter of the energy requirement compared to the system treating 1170 kg of wet product was deemed to be sufficiently conservative.

Furthermore, there was additional energy requirements for pumping process streams, conveyors, dosing pumps and process controls as summarized in table 15.

6.2.4.2 Spatial requirements

The precipitator reactor was sized using equation 19 with a reaction rate of 7.9 hr^{-1} as described in the methodology section with. A conversion of 90% was assumed and the volumetric flow rate of the process was 4.2 m^3/hr. Using these values according to equation 19 the required precipitator reactor volume was found to be approximately 5 m^3. The Ostara Pearl Modular Precipitator Reactor offering includes the Pearl 500 which has a treatment load capacity of 65 kg PO_4-P/d and reactor volume of approximately 12 m^3 (Ostara, 2018; Sikosana et al., 2016) . This is the smallest unit in the Ostara offering. The Pearl 500 was slightly oversized for the treatment of the design flow and phosphate load in this report. However, it was deemed to be sufficiently near the theoretical volume and load requirements to provide an accurate conceptual design basis. A footprint of 140 m^2 was recommended by Ostara for the Pearl 500 system. This footprint also made provision for space required for product handling (Screen sieves, air dryer and classifying hopper), product storage and a control room (Ostara, 2018).

As reported in Table 12 nine ED-1000A membrane stacks consisting of 50 cell pairs each was required to process 100 m^3/d of feed waste water. The dimensions for each membrane stack was approximately 400mm (L) by 600mm (H) by 150mm (W). For conceptual design purposes, it was assumed that each stack would be placed on a holding frame and would be bunded to contain any leakage. Therefore, the estimated total footprint of each stack was approximately 2 m^2. For nine ED-1000A membrane stacks a footprint of 18m^2 was required; not taking into account pathways, pipe and cable trenches, etc. It was assumed that the electrical control equipment for the electrodialysis stacks would be housed in the same control room as the precipitation system. Therefore, no additional space was included for housing the electrodialysis control equipment.

Furthermore, space was required for raw water storage, sodium hydroxide chemical dosing, and anti-scalant chemical dosing, the estimated footprints for these components are summarized in Table 17. The required storage volumes of the tanks and reactors are provided in Table 16.

Table 16: Option B storage requirements

Description	**Size (m^3)**	**Residence time/storage time**
Precipitator reactor storage volume	12	3 hr
Raw water tank volume	25	6 hr
Effluent Storage volume	25	6 hr
Sodium hydroxide storage volume	0.5	208 d
Anti-scalant storage volume	1	500 d

Table 17: Option B land requirement

Description	**Footprint (m^2)**
Precipitation reactor and associated equipment	140
Electrodialysis Units	18
Raw water tank	13
Effluent Storage tank	13
Sodium hydroxide dosing system	3.5
Anti-scalant dosing system	3.5
TOTAL ESTIMATED FOOTPRINT	**191**

As recorded in Table 17 the estimated footprint for the electrodialysis-struvite precipitation system was 191 m^2. As for option B a safety factor of 2 was applied to account for dead spaces, pathway, pumps and any other unforeseen spatial requirements. With a safety factor of 2 the overall surface area requirement was calculated as 382 m^2. Furthermore, it was assumed that the system would be housed in an enclosed factory building with a ceiling height of 12m to make provision for the 9.5m height of the struvite precipitation reactor.

6.2.4.3 Overall performance and operation

The electrodialysis component of the overall process was included in order to preferentially concentrate the phosphate ions and to create a phosphate rich stream to feed into the struvite precipitation reactor. Given that the magnesium and ammonium concentrations in the feed effluent was in excess stoichiometric proportions to phosphate (refer to Table 1) there was merit in increasing the phosphate ratio, however it was identified that struvite may already precipitate at the feed effluent concentrations and that the primary inhibitor to struvite formation may be the acidic pH. If this hypothesis was true the usefulness of including the electrodialysis setup could be questioned as it added significant capital and operational cost, while also increasing the overall complexity of the process.

Nevertheless, it was identified that the electrodialysis setup was effective at diluting the feed effluent to create a treated stream with an estimated TDS of less than half of the raw water stream. Thus, the desalination capabilities of the electrodialysis setup added merit to inclusion of the electrodialysis setup in the overall process.

The struvite reactor setup was well suited to removing phosphate from the effluent, with removals of upwards of 90% well documented (Sikosana et al., 2017; Zhang et al., 2013), thus from a phosphate removal perspective the technical performance of the process was good.

Aside from phosphate removal, the struvite precipitator also allowed for removal of ammonia. Typically ammonia removal of 20% is accepted when including a struvite precipitator in a municipal sewerage treatment plant process setup (Ostara, 2018). However, given the effluent characteristics of this particular fertilizer effluent, where the NH_4^+:PO_4^{3-} ratio was approximately 45, the ammonia removal capabilities of the struvite reactor became near insignificant as the ammonia removal was limited by the phosphate concentration as can be seen in the mass balance in Appendix A-2.

It was further worth noting, that another major ion in the raw waste water was nitrate. With a concentration of 7254 mg/L (refer to table 1), nitrate was the dissolved ion with the highest concentration in the feed water. The effect of nitrate on the electrodialysis-struvite reactor setup and in particular the effect of the high nitrate concentration on struvite precipitation was not addressed in depth in this study. However, since nitrate is an anion it was assumed it would be concentrated in the product stream of the electrodialysis stack. The separation of nitrate from the feed stream to the product stream was the primary reason for the significantly lowered total dissolved solids concentration in the diluate stream. Given the ionic separation efficiency of 80% in the electrodialysis stack (Altmeier, 2018), the total dissolved solids concentration of the raw waste water was diluted from approximately 16 000 mg/L TDS down to approximately 8 500 mg/L TDS as demonstrated in the mass balance in Appendix A-2. The phosphate component of this TDS decreased from 114 mg/L down to approximately 11 mg/L, thus the diluate water could be recycled back to a conventional waste water treatment plant with a fair amount of certainty that it would not contribute to struvite precipitation in the primary waste water treatment works.

Overall the process demonstrated reasonable desalination capabilities and excellent phosphate removal capabilities. However, the removal of particular key dissolved contaminants, in particular nitrate and ammonia were poor.

6.2.5 Economic evaluation

6.2.5.1 Overview of economic evaluation

The options were evaluated in terms of economic potential by considering capital expenditure (CAPEX), operating costs (OPEX) and revenues from product sales.

The CAPEX cost was determined by obtaining actual equipment and construction costs where possible. Where actual costs were not available, estimations derived from past literature was used. The cost calculations as well as supplier quotations for equipment costs of option B can be found in Appendices B-4, B-5, B-6 and C of this study.

From a civil engineering and construction perspective, for the purpose of determining the capital costs estimates for the electrodialysis-struvite process setup, it was assumed that the entire process would be housed in an industrial factory type building. The building had to provide 382 m^2 of space as described in section 6.2.4. The building would need a ceiling height of approximately 12m to ensure adequate working space above the equipment as the struvite precipitation reactor is 9m tall. The majority of the civil costs were calculated using a rate per m^2 for industrial factory space derived from statistics developed by Stats SA (StatsSA, 2015). Because the factory ceiling height was an unconventional 12 m high, a cost factor of 100% was used on the costs obtained from StatsSA. Further detailed assumptions and design considerations taken into account when determining the civil cost estimates can be viewed in appendix B.

For the mechanicals and electrical costs estimate, equipment requirements were determined by studying the treatment process. The struvite precipitator cost was obtained from Ostara (Napa_Sanitation_District, 2013), this cost included mechanicals for the precipitator reactor, product drying and screening system and product packaging. The cost provided by Ostara also included all electrical components for the above mechanicals, as well as a centralized control panel and control station. Additional mechanical and electrical costs included costs for the electrodialysis membrane stacks, storage tanks, dosing pumps and conveyance pumps. Costs for the electrodialysis stacks were obtained from PCCell GMBH; the remaining equipment costs were obtained by sizing the equipment and obtaining quotations from suppliers. These quotations can be found in the appendices under appendix B.

6.2.5.2 CAPEX

The best cost estimates for the treatment system were derived by reviewing the various engineering aspects of the system design as described above. Where possible real costs were used for engineering equipment and infrastructure. The reader is advised to review Appendix B-4, B-5 and B-6 for a clear breakdown of the cost calculations. Specific equipment quotations can be found in Appendix C.

Table 18: Capital Costs of Option B

DESCRIPTION	COST
Civil works	R642 000
Mechanical equipment	R35 086 000
Electrical equipment	R784 000
Sub-total Asset Capital Costs	R42 295 000
Consultant Professional Fees % of Capital Costs	11%
Professional Fees Costs	R4 653 000
TOTAL:	**R46 948 000**

* Electrical costs are for equipment and instrumentation not included in the Ostara scope of delivery (electricals for precipitator, screening and drying systems, product classification, and control system). Electricals for Ostara system are included in mechanical equipment costs, as this fee was obtained as a lump sum for the entire scope of delivery.
** Rates as specified by *"Guideline scope of services and tariff of fees for persons registered in terms of engineering profession act, 2000 (Act No. 46 of 2000).*

The professional fees in Table 18 were derived from the ECSA guidelines of services and tariff of fees for persons registered in terms of engineering profession act (Act No. 46 of 2000). As per this act, a fee of 5.5% of the overall costs of the works is required for electrical engineering consulting on projects involving electrical engineering, where the costs of the works exceed R34 360 000. Furthermore, as per this act a fee of 5.5% of the overall costs of the works is required for mechanical engineering consulting on projects involving mechanical engineering where the costs of the works exceed R42 940 000.

The capital costs for option B are nearly four times higher than that of option A. The reasons for this will be expanded on in section 6.4. However, it is already worth noting that the struvite precipitator system has been oversized to treat phosphate loadings four times higher than that of the waste water presented in this report and therefore the costs of the mechanicals of the system are inflated. This was unfortunately because the struvite precipitator from Ostara is a pre-fabricated item and the reactor chosen in this study was the smallest size available. Significant costs savings may be realized if a smaller system sized closer to the actual phosphate loading presented in the study is found. Furthermore, if a manufacturer could be found locally to fabricate the precipitator this should also provide further significant cost savings.

The costs for the struvite reactor setup also further increased indirect percentage-based costs, such as the professional fees costs, pipework costs and procurement fee costs. The struvite precipitator also significantly increased the building space requirements. A cost projection without the Ostara reactor was done to further illustrate the cost contribution of just the Ostara reactor. The capital cost summary for option B without the Ostara struvite precipitator is given in Table 19. A cost saving of over R 40 million is achieved when excluding the Ostara reactor and related percentage-based costs.

Table 19: Capital costs of option B without the Ostara Struvite precipitator setup

DESCRIPTION	COST
Civil works	R903 000
Mechanical equipment	R4 013 000
Electrical equipment	R784 000
Sub-total Asset Capital Costs	R5 700 000
Consultant Professional Fees % of Capital Costs	16%
Professional Fees Costs	R912 000
TOTAL:	**R6 612 000**

6.2.5.3 OPEX and revenues

The estimated rates for operation of the electrodialysis-struvite precipitation plant are summarized in Table 20.

Table 20: Operational Cost Rates for Option B

Description	Unit	Value	Source
Price of Sodium Hydroxide (caustic soda)	R/kg	7.64	Protea Chemicals
Price of anti-scalant	R/kg	37.69	Genesys International
Price of electricity	R/kWh	2.61	City of Cape Town- Electricity generation and distribution tariffs 2017/2018
Operator Wages	R/d	R500.00	Assumed
Water	R/kL	R57.00	City of Cape Town Water and Sanitation tariffs 2017/2018
Maintenance costs	% Asset Capital costs per annum	5%	(Sikosana et al., 2015)

The annual operational costs were calculated, taking into account the rates presented in table 20. The summarized annual operational costs are presented in table 21.

Table 21: Summarized annual operational costs for option B

Description	Annual Quantity	Rate	Total Annual Costs
Sodium Hydroxide	876	7.64 R/kg	R7 000
Anti-scalant	73	37.68 R/kg	R3 000
Electricity	117 000	2.61 R/kWh	R305 000
Operator Wages*	365 d	500 R/d	R548 000
Water**	365 kL	57 R/kL	R21 000
maintenance costs		5% of total asset costs per annum	R2 115 000
Total Annual Operational Costs:			**R2 999 000**

*Assumed that three operators per day will be required for the site

**Assumed consumption of 1000L of water per day for day-to-day activities on site such as water requirements in the lab, cleaning, toilets, etc.

The exact municipal water requirements for operation of the plant was assumed, for purposes of the cost estimation it was estimated that 1 000 L of municipal water would be used per day for day-to-day activities such as ablutions and cleaning which led to an annual usage of 365 kL. As option B had a much more significant mechanical component compared to option A, it was assumed that additional operators would be required, furthermore it was assumed an operator would also be required for product handling. For option B, it was assumed that three operators would be required on the plant at any given time.

Struvite was a revenue generating product produced by the plant. It was assumed that the plant is capable of producing high grade struvite (92 wt/wt%); that could be directly sold to market at a value of R 15/kg (Randall, 2018). The revenue projection from struvite production and sales is summarized in Table 22. Unfortunately, the low volumes of struvite produced compared to the inlet raw water volume makes the return on investment (ROI) produced from struvite revenue very long. It was identified that If the phosphate concentration could be further concentrated in the feed stream prior to the current process, this could potentially increase the feasibility of struvite production by increasing the phosphate to feed volumetric flow ratio. This could be done through a process such as an anaerobic digestion.

Table 22: Option B revenue calculations

Description	Unit	Quantity
Struvite Dry product (92 wt/w %)		
Daily production	kg/d	29
Annual production	kg/year	10 585
High grade Struvite retail Price	R/kg	15
Annual revenue from struvite sales	R/year	158 775

The capital and operational costs represented in this section are further reviewed in section 6.4. The costs are also compared alongside the costs of treatment options A and C to provide a frame of reference for the various costs

6.2.6 Resource recovery

Struvite is a natural forming complex in waste water streams containing nitrogen, phosphorus and magnesium that has found much success in application as a slow release fertilizer (Duley, 1998). Phosphates used in synthetic fertilizer production are obtained predominantly from igneous and sedimentary rock deposits in the Earth. These deposits are finite and it is expected that the earth's readily exploitable phosphate reserves will be depleted within in the next 45 to 100 years (Duley, 1998). One of the primary off-products from the electrodialysis-struvite reactor is the struvite mineral from the struvite reactor. The electrodialysis setup in the reactor process allows preferential concentration of the key elements in the struvite complex to be fed to the struvite reactor, thereby allowing more efficient struvite production. Zhang et al. (2013) were able to recover 93% of the feed phosphates as struvite from an anaerobic effluent obtained from a potato processing plant utilizing a selective electrodialysis feed to a struvite precipitation reactor. Hutnik et al. (2013) were able to decrease the phosphate concentration of a feed solution obtained from a phosphorus based fertilizer industrial plant from 0.445% by mass to 0.0009% by mass utilizing a continuous tube mixed suspension air sparged crystallizer reactor, where the phosphate was taken off primarily as crystallized struvite.

Thus, the ability of the electrodialysis-Struvite reactor setup to produce a valuable off-product in the form of slow releasing precipitated Struvite fertilizers gives it strong resource recovery credentials when applied as a waste treatment method for fertilizer production plant effluent.

6.3 Option C: Combined forward and reverse osmosis

6.3.1 Overview

Option C proposed using a combined forward osmosis - reverse osmosis (FO-RO) membrane setup to treat the fertilizer waste water. In the first step raw effluent was fed to the FO vessels as the FS with sea water as a DS. The osmotic pressure gradient between the two streams results in natural flow from the fertilizer effluent through the FO membrane into the DS. The diluted draw solution from the FO vessels was then fed with a high-pressure pump to a RO membrane setup. The FO-RO system produced a concentrated fertilizer stream which could be further treated in a crystallizer or precipitator to harvest the nutrients. A clean product water stream was also produced from the RO vessels of the FO-RO system. The clean product water stream could potentially be used for potable purposes following remineralisation.

6.3.2 Design basis

The conceptual design for option C considered treating 100 m^3/d of fertilizer plant effluent as outlined in the methodology section (section 5). For Option C the fertilizer effluent was treated utilizing a combined FO – RO setup. The key design criteria for option C are summarized in Table 23 below.

Table 23: Option C key design parameters

Description	Unit	Value	Comment/source
Influent volumetric flow	m^3/d	100	
Influent water quality		As per table 1	
Forward Osmosis Parameters			
FO membrane manufacturer		FTS	
Membrane model		OsmoF2O FO-8040-CTA-85-SDS	
Total membranes		30	
Membrane active area	m^2	13.5	(Fluid_Technology_Solutions_Inc, 2016)
Draw solution salt rejection	%	99.9	(Fluid_Technology_Solutions_Inc, 2016)
pH operating range of FO membranes		3 to 7	(Fluid_Technology_Solutions_Inc, 2016)
Phosphate separation efficiency	%	90	(Vallidares Linares et al., 2014)
Ammonia/nitrate separation efficiency	%	70	(Vallidares Linares et al., 2014)
Concentration factor (percentage of draw solution)	%	95	(Fluid_Technology_Solutions_Inc, 2016)
Average design flux	$L{\cdot}m^{-2}{\cdot}h^{-1}$	10	(Vallidares Linares et al., 2014)
Reverse Osmosis Parameters			
RO membrane manufacturer		DOW	
Membrane model		SW_{30}XHR-440i	
Total membranes		4	
Membrane active area	m^2	41	(Dow_Filmtec, 2017)
Minimum salt rejection	%	99.7	(Dow_Filmtec, 2017)
pH operating range of FO membranes		2-11	(Dow_Filmtec, 2017)
Maximum membrane recovery	%	15	(Dow_Filmtec, 2017)

6.3.3 Process description

Seawater was used as the DS for the FO-RO system. The total dissolved solids (TDS) concentration of the fertilizer effluent stream was estimated at 16 000 mg/L. Thus an indirect FO-RO process was utilized because the TDS concentration of seawater is approximately 35 000 mg/L (Jakhrani et al., 2012) and the fertilizer TDS concentration was not high enough to drive a direct FO-RO process. In the indirect FO-RO treatment setup, the fertilizer effluent was fed to the FO vessels generating a product stream with concentrated phosphates and nitrates. The sea water draw solution fed to the FO vessels was then diluted by natural osmosis of water from the fertilizer stream side through the membrane to the draw solution side. The diluted sea water product stream from the FO vessels was fed to a storage tank and then pumped with a high-pressure pump into the RO vessels. Using the diluted draw solution as the feed to the RO vessels resulted in pumping cost savings as the osmotic pressure of the draw solution is significantly lower than that of pure sea water. The process flow diagram of the process described is given in Figure 5.

It is noted that the biofouling potential of using sea water and the required pre-treatment to minimize such bio-fouling has not been addressed in this work. It is understood that such omission does not take into account the cycling up of minor elements from recycling and increased potential scaling and fouling of membranes which may have significant design and cost implications. The bio-fouling effects of using sea-water is important and should be taken into account in subsequent studies, where it may be found that utilizing a pure NaCl draw solution is better from a cost and operational perspective. Nevertheless, option C proceeds under the assumption that utilizing sea water as the draw solution is adequate for the purposes of this study.

Proceeding with sea water as the draw solution, the FO unit was sized using an average design flux of 10 $L \cdot m^{-2} \cdot h^{-1}$ (Vallidares Linares et al., 2014) thus at a flow rate of 100 m^3/d and a membrane active area of 13.5 m^2 approximately thirty OsmoF2O FO 8 inch membranes was required. The technical data sheet for the OsmoF2O 8 inch FO membranes can be viewed in Appendix D-16. The raw waste water was fed to the FO system using a standard centrifugal pump. The data sheet of centrifugal pump utilized can be found in Appendix D-4. The sea water draw solution was made up of fresh seawater and concentrated recycled brine from the RO system. The sea water was fed using a titanium grade centrifugal pump in order to avoid corrosion from the high salinity water. The data sheet of the titanium grade centrifugal pump can be viewed in Appendix D-15. The feed streams to the FO unit required minimum external pressure, because the transmembrane pressure of the FO membranes was low. The minimum transmembrane pressure of the OsmoF2O FO 8 inch membranes is 0.35 bar and maximum feed pressure is 5 bar (Fluid_Technology_Solutions_Inc, 2016). For the process mass balance, a salt rejection of 99.9%, phosphate rejection of 90% and nitrate and ammonia rejection of 70% was used for the FO system (Fluid_Technology_Solutions_Inc, 2016; Vallidares Linares et al., 2014). Processing of the concentrated fertilizer stream was not dealt with in this report, however the concentrated stream could be further treated in a precipitator type reactor similar to that of option B for recovery of a crystallized struvite product.

Key control variables for the FO system were the feed stream flow rates, pH of the feed streams and TDS concentrations of the feed streams. Unlike RO systems, the FO membranes can handle high concentrations of suspended solids and turbidity, with a maximum operational feed turbidity of 1000 NTU (FTS, 2016). Thus, the only pre-treatment required for the FO system was 100 µm microfiltration.

From the FO system the diluted DS was fed to a storage tank where it was pumped using a high-pressure RO feed pump. The TDS of the draw solution was diluted by 35% in the FO system, allowing for a pressure pump with a significantly lower supply pressure requirement when feeding into the RO pressure vessels. DOW's ROSA reverse osmosis simulation software was used to run a simulation on the RO system. The simulation was run using the diluted DS as the feed stream to the RO system. The simulation indicated a feed pressure of 42 bar was required for the system (the simulation results can be viewed in Appendix A-4). A high pressure permanent magnet driven motor pump was required to deliver the water at this pressure to the RO system. The pump internals were constructed from high grade stainless steels (AISI 904L) and tungsten carbides. The data sheet of the high-pressure RO pump can be viewed in appendix D-14. The RO system was operated at a flux of 14.68 $L.m^{-2}.h^{-1}$ using four DOW 8 inch sea water membranes with a minimum salt rejection of 99.7%. The data sheet of the DOW 8 inch sea water membrane can be viewed in Appendix D-17. The RO system was configured to have one pass with the four membranes in series resulting in an overall recovery of 40%. With this configuration, the simulation indicated that the RO system produced 2.4 m^3/hr of product water containing a TDS of 132.66 mg/L. This product stream could be sent for remineralisation after which it could potentially be used directly as a potable water source.

Key control variables for the RO system were the feed stream flow rate and pressure, pH of the feed streams and TDS concentrations of the feed streams. The RO feed stream was first filtered to 1 micron as a pre-treatment prior to being fed to the RO membranes. Anti-scalant was dosed into the RO feed stream after pre-treatment to prevent scale forming on the RO membranes in order to extend their operational life.

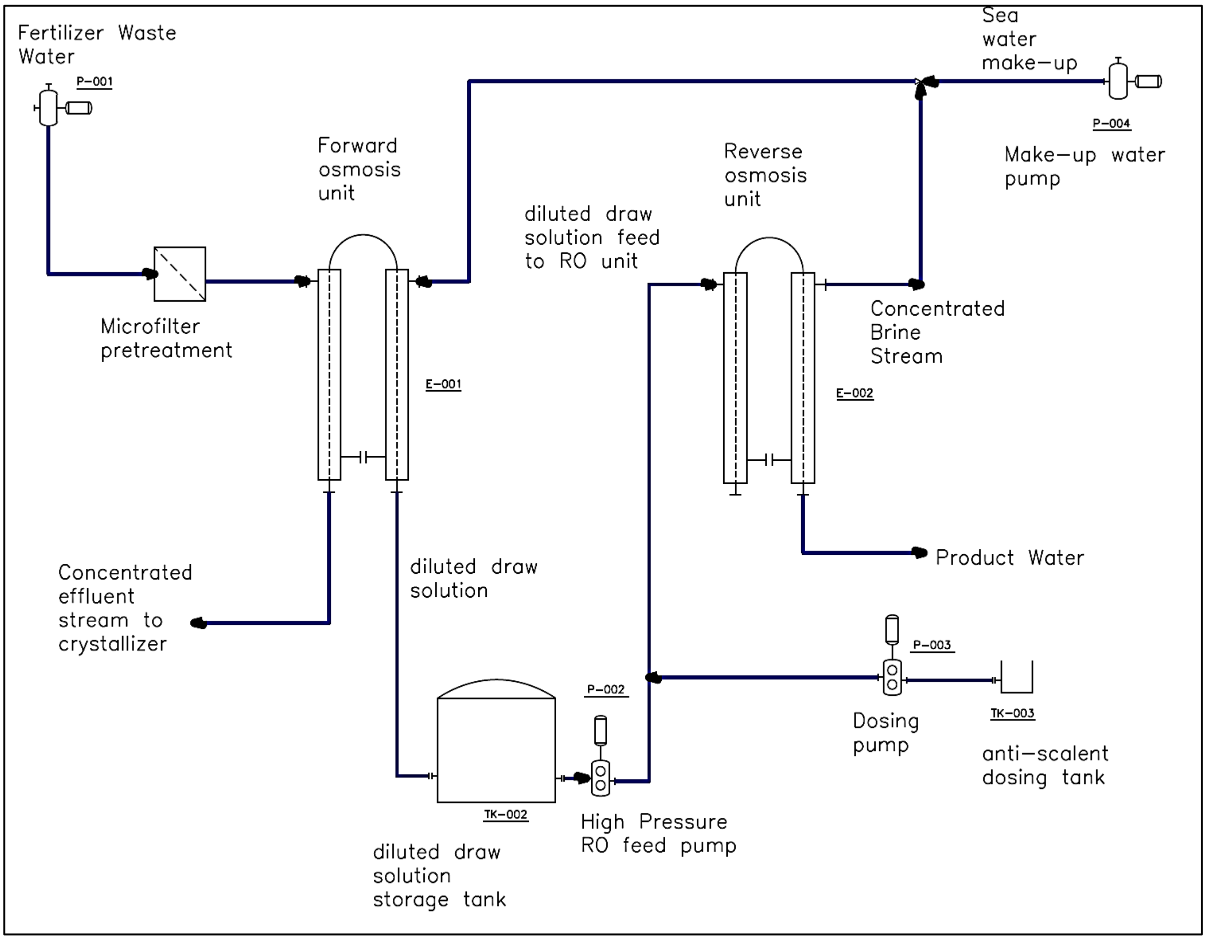


Figure 5: Combined Forward and Reverse Osmosis Process Flow Diagram

6.3.4 Technical evaluation

6.3.4.1 Mass and energy flows

The primary process streams for the FO-RO process included the raw effluent and sea water DS feed streams to the FO vessels, the concentrated effluent product and diluted DS streams exiting the FO vessels, the brine stream exiting the RO vessels and the permeate product stream exiting the RO vessels. The process streams are displayed in Figure 5. Mass balances for the TDS, phosphates, nitrates and ammonia of the overall incoming and exiting streams are given in Table 24. The volumetric flows of the overall incoming and exiting streams are also given in Table 24.

From the mass balance summary given in Table 23, it can be seen that 60 m^3/d of a treated potable water was produced from 100m^3/d of fertilizer effluent and 100 m^3/d of sea water. Furthermore, it can be seen that the nutrients in the waste water stream exiting the FO system was concentrated to as much as 50%.

Additional minor process streams in the FO-RO system include the anti-scalant dosing. The exact anti-scalant dosing requirement can be determined by a simulation using simulation software obtained from the anti-scalant supplier, however for the purposes of this report a feed stream concentration of 2ppm has been deemed adequate, as recommended by Genesys anti-scalant manufacturers (Genesys_International, 2018).

Table 24: Option C mass balance

Description	Unit	Inputs	Outputs
PHOSPHATE MASS BALANCE			
Primary Incoming streams			
Raw Waste Water: Phosphate	kg/d	11.4	
Sea Water: Phosphate	kg/d	0.1	
Primary exciting streams			
Concentrated Waste Water: Phosphate	kg/d		10.3
Product Water: Phosphate	kg/d		0.0
Brine Stream: Phosphate	kg/d		1.2
TOTAL	**kg/d**	**11.5**	**11.5**
AMMONIA MASS BALANCE			
Primary Incoming streams			
Raw Waste Water: Ammonia	kg/d	519	
Sea Water: Ammonia	kg/d	0	
Primary exciting streams			
Concentrated Waste Water: Ammonia	kg/d		360
Product Water: Ammonia	kg/d		3
Brine Stream: Ammonia	kg/d		158
TOTAL	**kg/d**	**519**	**520**
NITRATE MASS BALANCE			
Primary Incoming streams			
Raw Waste Water: Nitrate	kg/d	725	
Sea Water: Nitrate	kg/d	0	
Primary exciting streams			
Concentrated Waste Water: Nitrate	kg/d		505
Product Water: Nitrate	kg/d		2
Brine Stream: Nitrate	kg/d		217
TOTAL	**kg/d**	**726**	**725**

TDS MASS BALANCE			
Primary Incoming streams			
Raw waste water: TDS	kg/d	1626	
Sea water: TDS	kg/d	3500	
Primary exciting streams			
Concentrated waste water: TDS	kg/d		1579
Product Water: TDS	kg/d		9
Brine Stream: TDS	kg/d		3539
TOTAL	**kg/d**	**5126**	**5127**
FLOW BALANCE			
Primary Incoming streams			
Raw waste water: Flow	m^3/d	100.0	
Sea water: Flow	m^3/d	100.0	
Primary exciting streams			
Concentrated waste water: Flow	m^3/d		47.5
Product Water: Flow	m^3/d		61.0
Brine Stream: Flow	m^3/d		91.5
TOTAL	**m^3/d**	**200.0**	**200.0**

The energy requirements to operate the equipment for the FO-RO process is presented in Table 25. The primary energy user for the process was the high-pressure feed pump to the RO system. The RO high pressure pump had to deliver approximately 6.25 m^3/hr of diluted draw solution at a pressure of 42 bar. To obtain this high pressure the pump motors required large amounts of energy. The data sheet for the high-pressure RO pump can be found in Appendix D-14. The pump drew approximately 22.5 kW of power at a duty point of 6.25 m^3/hr and 42 bar.

Table 25: Option C estimated energy requirements

Description	Unit	Value
Feed pump – Appendix D-4	kW	1.1
High Pressure Feed pump – Appendix D-14	kW	22.5
Sea Water make-up pump – Appendix D-15	kW	0.55
Anti-scalant dosing pump – Appendix D-11	kW	0.2
Process instruments and control panel	kW	1
Estimated Total Power Requirement	**kW**	**25.35**
TOTAL*	kWh/d	608
TOTAL	kWh/year	222 000

*Electrical consumption based on drawing maximum power requirement for 24 hours a day.

6.3.4.2 Spatial requirements

As reported in Table 23, thirty OsmoF2O FO-8040-CTA-85 Forward Osmosis membranes and four DOW $SW_{30}XHR$-440i Reverse Osmosis membranes were used for the process. The dimensions of the membranes are given in Table 26 below.

Table 26: Membrane specifications

Description	Length (m)	Outer Diameter (mm)	Inner feed aperture diameter (mm)
OsmoF2O FO-8040-CTA-85	1 016	201	73
SW_{30}XHR-440i	1 029	201	29

The FO and RO membranes were housed in two element pressure vessels with approximate overall length of 2.2 m. For the RO system it was assumed that two pressure vessels would be stacked one on top of the other to house all four of the membranes. For the FO system it was assumed that five pressure vessels would be stacked on top of each other. In total 15 pressure vessels were required for the FO system, thus the FO system consisted of three rows of five stacked pressure vessels. It was further assumed that both the RO and FO system would be built onto stainless steel frames. Taking into account piping and various fittings, the estimated footprint of the RO system was 2.5 m by 2 m and the estimated footprint of the FO system was 2.5 m by 6 m. This resulted in a footprint of 5 m^2 and 15 m^2 for the RO system and FO system, respectively.

Additional equipment that had significant spatial requirements included the high-pressure RO pump, feed pumps, storage tanks, anti-scalant dosing system, pre-treatment setup and control rooms. The residence times and volumes of the storage tanks are summarized in Table 27. The estimated footprints of all the major equipment and equipment housing is summarized in Table 28.

Table 27: Option C storage requirements

Description	Size (m^3)	Residence time/storage time
FS tank volume	25	6 hr
Diluted DS storage volume	25	4 hr
Anti-scalant storage volume	10	10 d

Table 28: Option C estimated footprints

Description	Footprint (m^2)
FO and RO system	20
Pre-treatment systems	20
High Pressure RO feed pumps	5
Raw water and sea water feed pumps	5
Control room	20
FS tank	13
Diluted DS tank	13
Anti-scalant dosing system	5
TOTAL ESTIMATED FOOTPRINT	**101**

As displayed in Table 28 the estimated footprint for the overall FO-RO system was 101 m^2. A safety factor of 2 was applied to account for dead spaces, pathways, piping, other miscellaneous equipment and any other unforeseen spatial requirements. With a safety factor of 2, the overall surface area requirement was 202 m^2. It was assumed that the system would be housed in an enclosed factory building with a standard ceiling height of 6m to allow for entry of trucks.

6.3.4.3 Overall performance and operation

Unlike option A and option B, the FO-RO process was a completely physical separation process involving no chemical reaction. Since it was a physical separation process there was no net removal of any of the nutrients or contaminants by converting them into more benign compounds or resources; such as nitrogen gas or struvite fertilizer for option A and option B, respectively. For option C there was only an exchange of nutrients from one stream to another via membrane separation.

The physical separation processes in option C were effective in concentrating or removing contaminants and nutrients from the desired streams as required. This is demonstrated by an overall clean water recovery of 30% from the 100 m^3/d of raw effluent and 100 m^3/d of sea water. Furthermore, the treatment setup effectively concentrated the nutrients in the effluent stream by 50%. Hence, the performance of the plant from a physical separation perspective was good.

However, the resource recovery performance of option C was poor as no precipitator or crystallizer is included in the process. It was determined that if the concentrated effluent stream was not used for resource recovery the discharged effluent would have a worse impact on the environment, because the concentrated waste stream had a higher eutrophication potential than the raw waste water. Thus, it was concluded that it would be important to include some resource recovery mechanism to future fertilizer FO-RO designs even if economically unfeasible.

The clean water product stream produced from the process did give some merit to option C's resource recovery capabilities. The clean water stream had a very low total dissolved solids concentration and could be used for potable use with very little additional treatment. Although an overall recovery of 30% clean water is low from a conventional RO treatment perspective, where recoveries of 50%-80% are common, this recovery was evaluated from the perspective that one of the primary objectives of the process was to concentrate the contaminants in the effluent stream in order to produce an effluent stream more favourable for resource recovery.

Overall the process demonstrated very good desalination and nutrient removal capabilities. However, it was deduced that unless the concentrated nutrient stream was used for beneficial reuse via the appropriate resource recovery processes the negative impact on the environment caused by the concentrated waste stream would outweigh the positives of the process.

6.3.5 Economic evaluation

6.3.5.1 Overview of economic evaluation

Option C was evaluated in terms of economic potential by considering capital expenditure (CAPEX), operating costs (OPEX) and any potential revenues from product sales.

CAPEX cost was determined by obtaining actual equipment and construction costs were possible. Where actual costs were unavailable, estimations derived from past literature were used. The cost calculations as well as supplier quotations for equipment costs can be found in Appendix B-7, B-8, B-9 and C of this study.

From a civil engineering and construction perspective, for the purpose of determining the capital costs estimate for the FO-RO osmosis setup, it was assumed that the entire process would be housed in an industrial factory type building. The building had to provide 202 m^2 of space as described in section 6.3.4. A standard 6m high ceiling factory type was assumed for costing the factory space. The majority of the civil costs were calculated using a rate per m^2 for industrial factory space derived from statistics developed by Stats SA (StatsSA, 2015). Further detailed assumptions and design considerations taken into account when determining the civil cost estimates can be viewed in appendix B-7.

For the mechanicals and electrical cost estimates, equipment requirements were determined by reviewing the treatment process and drawing up a high-level equipment list. Where possible, equipment costs were obtained from relevant suppliers as indicated in the equipment lists (Appendix B-8 and B-9). The major mechanical cost contributors were the membrane vessels and membranes for the forward and reverse osmosis processes, the high-pressure pump for the RO treatment and the pre-treatment systems for the FO and RO membranes. The quotations for the various mechanical equipment can be viewed in Appendix C. Costs for the high-pressure RO pumps were obtained from Grundfos South Africa, the data sheet and cost quotation for the high-pressure pump is provided in Appendix D-14 and C-5, respectively. Costs for the FO membranes and pressure vessels were obtained from fluid technology solutions, Inc. The costs quotation for the FO membranes can be viewed in Appendix C-4. Costs for the RO membranes and pressures vessels were obtained from a South African DOW-Filmtec supplier. As mentioned the costs for these various components along with data sheets can be found in the appendices C and D, respectively. Other significant mechanical equipment costs include feed pumps and storage tanks. Furthermore, connecting pipework and valves costs were accounted for by using a factor of 10% of all the mechanical equipment listed in the materials list.

The major electrical costs contributors were the process instrumentation and control panel. The process instrumentation included magnetic flow meters for all primary process streams, electrical conductivity sensors for all primary process streams and low level and high-level pressures sensors for pumps. The control panel includes a programmable logic controller (PLC) with sufficient inputs and outputs for all the instrument sensors, a human-machine-interface (HMI) for local control of the panel and variable frequency drives for control of the pumps. The costs for the various process instrumentation and electrical equipment can be viewed in Appendix B-9 and C-3.

A procurement factor of 25% was used for both the mechanical and electrical equipment costs in order to take into account costs associated with the labour and logistics of procuring the equipment. The detailed mechanical and electrical equipment costing along with the various supplier quotations used in the costing can be found in appendices B and C.

6.3.5.2 CAPEX

The best cost estimates for the treatment system was derived by reviewing the various engineering aspects of the system design as described above. Where possible real costs have been used for engineering equipment and infrastructure. The reader is advised to review appendix B-7, B-8 and B-9 for a clear breakdown of the cost calculations.

Table 29: CAPEX costs for option C

DESCRIPTION	COST
Civil works	R1 823 000
Mechanical equipment	R5 495 000
Electrical equipment	R1 586000
Sub-total Asset Capital Costs	R 8 904 000
Consultant Professional Fees % of Capital Costs	16%
Professional Fees Costs	R1 425 000
TOTAL:	**R10 329 000**

* Rates as specified by *"Guideline scope of services and tariff of fees for persons registered in terms of engineering profession act, 2000 (Act No. 46 of 2000).*

As can be seen from Table 29, the biggest capital cost for option C was the mechanical equipment costs. This was expected as the process was primarily a mechanically driven process and the civil costs were limited to the factory space required to house the process. No portion of the civil costs were actually related to the direct operation of the process; unlike option A where the process tanks were constructed from concrete and were part of the direct operation of the process. Pipework costs were also included in the mechanical costs rather than the civil costs as the pipework was fairly process specific and not just bulk conveyance pipework.

The professional fees were derived from the ECSA guidelines of services and tariff of fees for persons registered in terms of engineering profession act (Act No. 46 of 2000). As per this act, a fee of 8% of the overall costs of the works is required for electrical engineering consulting on projects involving electrical engineering where the costs of the works exceed R 2 690 000. Furthermore, as per this act a fee of 8% of the overall costs of the works is required for mechanical engineering consulting on projects involving mechanical engineering where the costs of the works exceed R 3 760 000.

The capital costs of the various options are further compared in section 6.4.

6.3.5.3 OPEX and revenues

The estimated rates for operation of the combined FO-RO plant are summarized in Table 30.

Table 30: Option C operational rates

Description	Unit	Value	Source
Price of anti-scalant	R/kg	37.69	Genesys International
Price of electricity	R/kWh	2.61	City of Cape Town- Electricity generation and distribution tariffs 2017/2018
Operator Wages	R/d	R500.00	Assumed
Water	R/kl	R57.00	City of Cape Town Water and Sanitation tariffs 2017/2018
Maintenance costs	% Asset capital costs per annum	5%	(Sikosana et al., 2015)

The annual operational costs were calculated taking into account the rates presented in table 30. The summarized annual operational costs are presented in table 31. Where the anti-scalant requirements are described in Section 6.3.4.1 and energy requirements are described in Table 25.

Table 31: Option C annual operating costs

Description	Annual Quantity	Rate	Total Annual Costs
Anti-scalant	111	37.68 R/kg	R5 000
Electricity	222 000	2.61 R/kWh	R580 000
Operator Wages*	365 d	500 R/d	R183 000
Water**	365 Kl	57 R/kl	R21 000
maintenance costs		5% of total asset costs per annum	R366 000
Total Annual Operational Costs:			**R1 155 000**

*Assumed that one operator per day will be required for the site
**Assumed consumption of 1000L of water per day for day-to-day activities on site such as water requirements in the lab, cleaning, toilets, etc.

The exact municipal water requirements for operation of the plant has to be assumed as there was no clear method of calculation. For the purposes of cost estimation it was estimated that 1 000L of municipal water would be used per day for day-to-day activities such as ablutions and cleaning which led to an annual usage of 365 kL. From the process description and maintenances costs, it was deduced that option C would have a similar operator requirement to option A. Thus, it was assumed that one operator would be required per day to meet the operational requirements for option C. The electricity costs were the biggest operational cost for option C, this was expected because the high-pressure RO pump was energy intensive.

6.3.6 Resource recovery

The FO setup in the overall FO-RO process resulted in a concentrated effluent stream with an increase of 50% in phosphate concentrations and 35% for ammonia and nitrate concentrations. This concentrated effluent stream could have been processed in a precipitator setup similar to option B. However this additional treatment was not considered in this study. Nevertheless, the potential for further processing of the concentrated effluent stream for resource recovery through mineral precipitation such as struvite is realistic.

Additionally, the FO-RO setup was able to recover 61 m^3/d of clean water from 100 m^3/d of raw effluent and 100 m^3/d of seawater. This water could potentially have been utilized for potable purposes with minimal further treatment.

Although the current process presented in this report for the FO-RO setup has not been geared toward resource recovery, the concentrated product waste water stream and treated permeate stream could have been processed with relative ease by including tertiary treatment processes already presented for option B in this report.

6.4 Option comparisons

6.4.1 Technical performance comparison

From a treatment performance and environmental impact perspective the various options were primarily evaluated based on the ammonia, phosphate, nitrate and TDS removal efficiencies of the processes. The removal of these components was selected as they were deemed to be most representative of the eutrophication potential and environmental degradation potential of the specific fertilizer waste water. The concentrations of these various components in the treated water streams for the various options is presented in Table 32 below.

Table 32: Treated Water concentrations for the various options

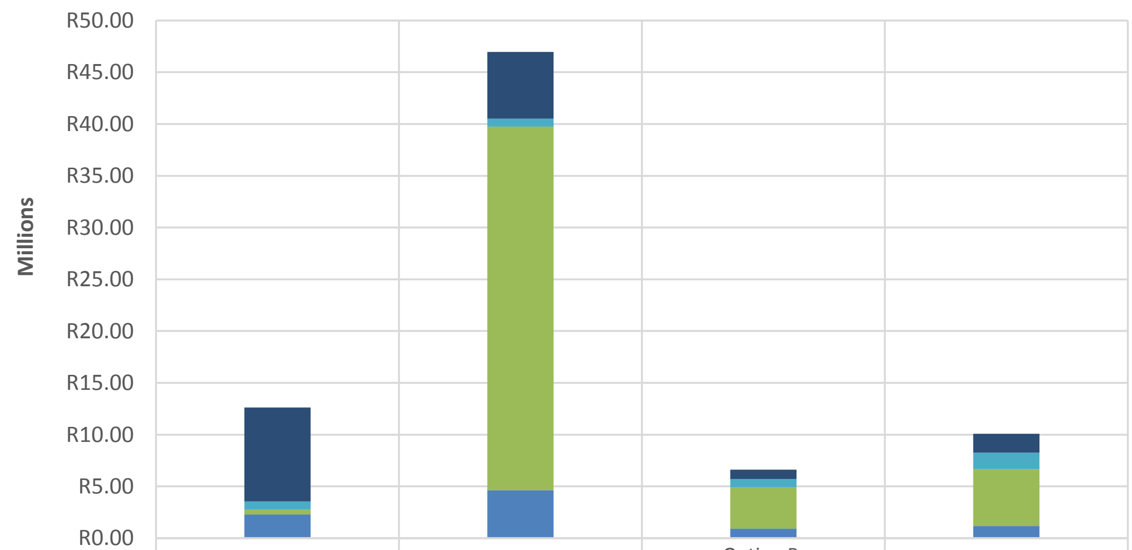
Description	**TREATED WATER CONCENTRATIONS**			
	Nitrates (mg N/L)	**Ammonia (mg N/L)**	**Phosphates (mg P/L)**	**TDS (mg/L)**
Option A: Sharon-Anammox	9 180	112	114	13 500
Option B: struvite-electrodialysis	1 450	5 170	11	8 560
Option C: FO-RO - concentrated FS	10 641	7 580	216	33 300
Option C: FO-RO - product stream	39	42	0	154

As can be seen in Table 32 option A was effective at removing ammonia but shows negligible phosphate removal and minimal TDS removal. Furthermore, option A increased the nitrate concentration rather than removing it. This can be explained by considering that option A was a biological nutrient removal process focusing on removing ammonia; it had no significant application for removal of nitrates, phosphates and TDS. Option C produced two product streams, a concentrated feed stream and permeate stream. The removal efficiency of nutrients in the permeate stream was high and the water was close to drinkable standards. However, the removal efficiency of the RO process is offset by the water quality of the concentrated waste stream. The process setup of the FO process was designed to concentrate the nutrients in the effluent rather than remove it. This is in order to make it suitable for enhanced nutrient recovery downstream. However, since there was no resource recovery step included in the process presented for option C, the FO process created, however the design of option C had a massive shortcoming in that a nutrient recovery step was not included for the concentrate stream. It can be observed that If the permeate stream and concentrated waste stream were blended for option C the effluent water quality would be better than option A and B, however this would be counter-productive as the process was designed to initially concentrate the incoming fertilizer stream.

6.4.1 Cost economic comparisons

6.4.2.1 Capital cost comparisons

The capital costs for all three options are displayed graphically in Figure 6 below. The overall capital cost were broadly divided into four major cost contributors: professional fees, electricals, mechanicals and civils. The portions of the overall capital costs divided into these four contributors are also displayed graphically in Figure 6 below. The costs for option B without the struvite precipitator has also been included in Figure 6 to demonstrate how much of the capital costs of option B can be attributed to the struvite precipitator.



	Option A: Sharon-Anammox	Option B: Struvite+Electrodialysis	Option B: Electrodialysis without precipitator	Option C: FO-RO
Civils	R9,085,000.00	R6,425,000.00	R903,000.00	R1,823,000.00
Electricals	R789,000.00	R784,000.00	R784,000.00	R1,586,000.00
Mechanicals	R481,000.00	R35,086,000.00	R4,013,000.00	R5,495,000.00
Professional Fees	R2,279,000.00	R4,653,000.00	R912,000.00	R1,171,000.00

Figure 6: Capital Costs of the various process options

From Figure 6 we can see that the capital costs of option B with the struvite precipitator is significantly more than both option A and C. Option B is 370% and 460% more expensive than options A and C, respectively. However, this cost difference is misleading as the major portion of the costs for option B was derived from the capital costs for the Ostara struvite precipitator setup. The capital cost contribution of the struvite precipitator can be seen in the third cost column of Figure 6: "Option B: Electrodialysis without precipitator". The direct costs for the struvite reactor setup is approximately R 24 850 000 which accounted for close to half of the overall capital costs of option B. The costs for the struvite reactor setup also further increased indirect percentage-based costs, such as the professional fees costs, pipework costs and procurement fee costs. Furthermore, the struvite precipitator reactor significantly increased the building space requirements. Excluding the Ostara struvite precipitator setup resulted in an overall cost saving of nearly R 40 000 000; the electrodialysis setup then becomes the cheapest treatment option of the three options.

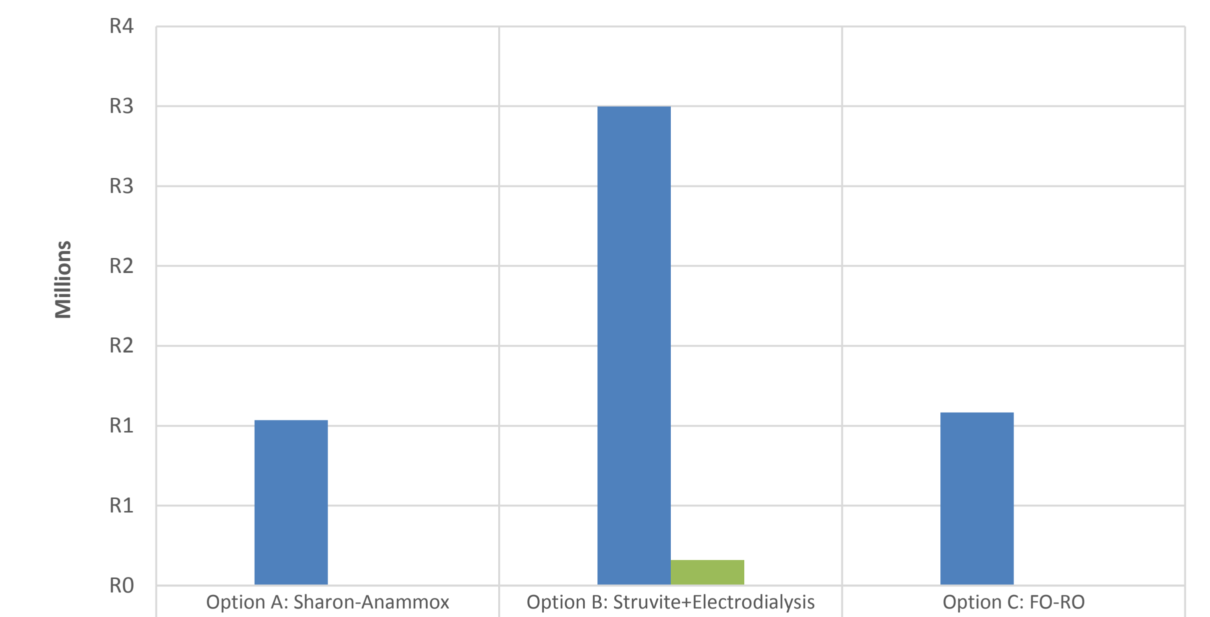
However, removing the struvite precipitator from option B significantly impacts the technical performance of the process. The struvite precipitator contributes significantly to the phosphate and ammonia removal performance of the overall process for option B. Furthermore, the struvite precipitator also sets option B apart from option A and option C in that it was the only process presented that could effectively recover resources within the battery limits of the presented processes.

The considerable costs of the struvite precipitator highlight that significant cost savings could have potentially been achieved if the struvite precipitator could have been manufactured locally. The Ostara reactor is an imported item from Canada. The specific Ostara reactor chosen for this study was the smallest one in the Ostara range (The Ostara 500) but was still oversized for the specific design basis of this report. Thus, a locally manufactured precipitator that is sized specifically for the flow requirements of the design basis may potentially lead to significant cost savings. It is advised that this be investigated in future work.

As mentioned, option B is the only process with significant resource recovery capabilities. The lack of resource recovery capabilities for option C results in a significant drop in its technical performance. The concentrated effluent stream from the FO vessels in option C required further treatment to remove the nutrients and make option B feasible from a technical perspective. Thus, a similar process to the Ostara setup should have been included in option C to increase its technical feasibility. This would have resulted in capital costs for option C like that presented for option B in Figure 6, if the same Ostara reactor setup was used.

6.4.2.2 Revenue and Operational cost comparisons

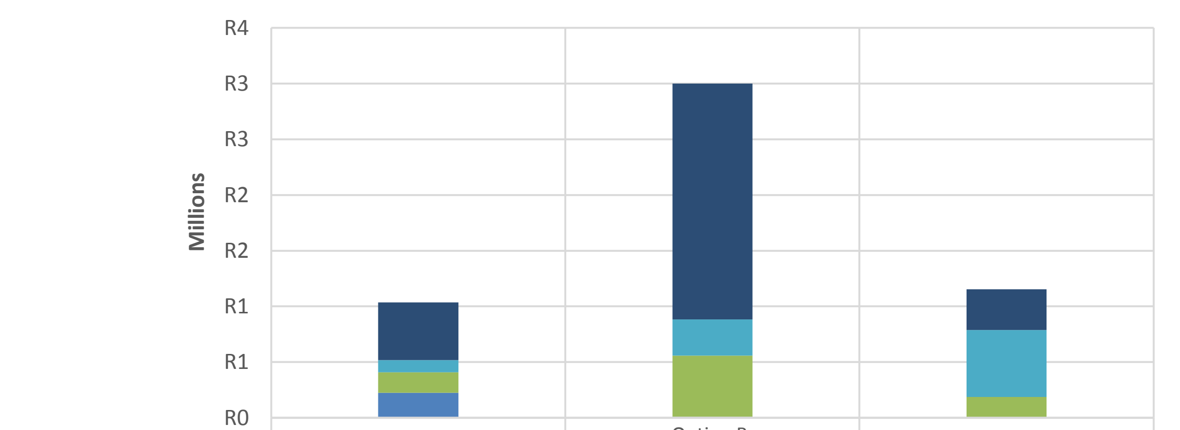
The electrodialysis-struvite reactor setup was the only process presented in this report that had any significant revenue generation. However, at an estimated production rate of only 10.5 tons of struvite annually, the revenue generation from struvite sales was small compared to the overall operational costs of operating the struvite precipitator. The estimated operational costs for running the struvite precipitator component of the overall process amounted to approximately R 2 200 000 annually which was more than 70% of the overall operational costs of option B. It was estimated that R 160 000 could be generated from struvite sales annually at the current market value of struvite as displayed in Figure 7. Hence, for the sale of the struvite to be financially feasible, the costs of struvite needed to increase by more than 1000%. Alternatively, the process had to be altered to produce struvite more efficiently by decreasing the cost of operation and maintenance.



	Option A: Sharon-Anammox	Option B: Struvite+Electrodialysis	Option C: FO-RO
OPEX	R1,035,000	R2,999,000	R1,083,000
REVENUE	R0	R158,775	R0

Figure 7: OPEX and Revenue comparison of the various process options

The operational cost breakdown of the various processes was broken down into four primary contributing factors: maintenance, utilities, labour and chemical consumables as displayed in Figure 8. Similar to the capital costs presented in Figure 6, the operational costs of option B were somewhat skewed by the inclusion of the Ostara struvite precipitator setup. This could be seen by the maintenance portion of the operational costs that are a direct percentage of the mechanical and civil capital costs components of the various options. As mentioned previously, the struvite precipitation reactor setup contributed approximately R 2 200 000 to the overall annual operational costs of option B. Without the inclusion of the struvite precipitator reactor setup, the operational costs of all three options would have been similar at approximately R 1 000 000 annually.



	Option A: Sharon-Anammox	Option B: Struvite+Electrodialysis	Option C: FO-RO
Maintenance	R518,000	R2,115,000	R366,000
Utilities	R109,000	R326,000	R601,000
Labour	R183,000	R548,000	R183,000
Chemical consumables	R225,000	R10,000	R5,000

Figure 8: OPEX costs breakdown of the various process options

6.4.3 Resource recovery comparisons

Option B was the only process presented in this report with significant resource recovery capabilities. Once again, if a struvite precipitator was included in the process for option C, then option C would also have had resource recovery capabilities. However given the current battery limits of the processes presented in this study, option B was the only process with resource recovery capabilities and by default was the best process overall from a resource recovery perspective.

7. CONCLUSION

7.1 Study objectives

This study set out to provide potential treatment solutions for effluent derived from a fertilizer production plant containing high nitrate, ammonia, phosphate and overall high total dissolved solids concentrations. Three novel treatment options were presented as potential solutions; these included the Sharon-Anammox bioreaction, the electrodialysis with a struvite precipitator and a combined forward and reverse osmosis process.

The primary objective of this study was to provide a critical evaluation of these three novel treatment techniques by mainly reviewing the following criteria:

- Overall technical feasibility
- Economic feasibility
- Resource recovery ability

7.2 Selection of best option

For the economic feasibility criteria, option C was found to be best from a capital as well as an operational expenditure perspective with capital costs and annual operational costs of R10 million and R1 million, respectively. The approximate capital cost and annual operational costs of options A were R13 million and R1 million, respectively and the costs for option B were R47 million and R3 million, respectively. However, it was identified that the addition of the Ostara struvite precipitator setup significantly inflated the costs of option B. The actual capital costs for option B without the Ostara reactor was closer to R7 million. It was hypothesized that significant cost savings could be achieved if a local manufacturer could be found who could build the reactor for the specific treatment volume required.

From a revenue perspective, option B was the only process capable of significantly recovering resources for revenue production because of the addition of the struvite reactor. However, the revenue produced from struvite compared to the costs of the struvite precipitation infrastructure costs was too small to make the process profitable.

From a technical feasibility perspective, all options were found to perform poorly as demonstrated by the negative effluent quality Index (EQI) values of the processes. Option A performed poorly because it was unable to significantly remove nitrates, phosphate and total dissolved solids; although it was effective at removing up to 90% of the influent ammonia concentration. Option B was moderately effective at removing nitrates, phosphates and total dissolved solids, but performed poorly overall because it was ineffective at removing ammonia. Option C performed poorly because the forward osmosis step concentrated the nutrients in the FS rather than removing it; this was done with the intention of allowing easier resource recovery of the nutrients downstream. However, since a resource recovery step was not included in the initial process design, the concentration of the nutrients resulted in a concentrated effluent stream with a higher potential for eutrophication. If option C was coupled with a resource recovery step, it would most likely have outperformed both option A and B from a technical feasibility perspective. However, in general all the presented processes performed inadequately from a treated water quality perspective and cannot be considered for effective treatment with the current presented designs.

Taking into account the findings of the performance criteria evaluation presented thus far, option C was found to perform best overall. Although, it is recommended that a resource recovery step, such as a struvite precipitation reactor, be included to the combined FO-RO process to make it technically feasible from an effluent quality perspective and to give it resource recovery capabilities. It is further recommended that a local manufacturer be found to build the precipitation reactor for the required volumes to save on costs and increase the economic feasibility of the process.

7.3 Recommendations for future research

Based on the findings of this report, the following recommendations are made for future research:

- **Feasibility of locally produced struvite precipitation reactor:**
  The capital costs of the struvite precipitation reactor presented in this study largely resulted in the process being economically unfeasible. However, the costs of the struvite precipitation reactor presented in this work was obtained from a Canadian supplier. Significant cost saving could potentially be achieved if the reactor was manufactured locally and according to the specific size requirements. It is recommended that further research be done in determining a potential local manufacturer for an industrial scale struvite precipitation reactor. It is further recommended that a feasibility study be done on a utilizing a locally manufactured precipitation reactor for resource recovery from fertilizer effluent.
- **Fertilizer effluent water quality:**
  A further contributing factor to the economic unfeasibility of the resource recovery option was that the overall phosphate loading was too low from a resource recovery perspective. However, the fertilizer production facility presented in this study was a nitrogen-based fertilizer producer. Thus, it is recommended that further feasibility studies be completed on fertilizer effluent obtained from a phosphate-based fertilizer production plant. Theoretically, fertilizer production facilities should be ideal for a struvite recovery process as the water usually contains excess magnesium, unlike other struvite recovery facilities where magnesium concentrations are low and have to be added to the process. The dosing of magnesium results in additional operational costs and potential chemical by-products, which is often a major cost factor in struvite production facilities.
- **Combined FO-RO with locally manufactured precipitation reactor:**
  The overall performance of the combined forward-reverse osmosis process was good in this study. The primary concern with the process was that there was no resource recovery step included to treat the concentrated effluent. Therefore, it is recommended that further research be done in determining the economic feasibility of the combined FO-RO process with a locally manufactured struvite precipitation reactor included in the process.
- **Experimental validation:**
  The mass balances completed in this study were all based on data obtained from literature, it is recommended that lab-scale, and potentially pilot-scale, experiments be conducted to verify the results obtained from this study.

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9. APPENDICES

APPENDIX A: TECHNICAL ASSESSMENT DATA

A-1: Sharon-Anammox Mass balance

REACTION IN SHARON REACTOR:

	NH_4^+	O_2		H^+	NO_2	H_2O
Mass	519 000 g/d	782 000 g/d	→	32 900 g/d	750 000 g/d	293 000 g/d
Mol	28 800 mol/d	24 400 mol/d		32 600 mol/d	16 300 mol/d	16 300 mol/d

RAW WATER

INFLUENT FLOW RATE:

100 m^3/d

INFLUENT CONCENTRATION:

Parameter	Concentration	Unit
EC	17 887	uS/cm
Estimated TDS (based on EC)	16 261	mg/L
Ammonium	5 190	mg/L
Nitrate	7 250	mg/L
Nitrite	0.49	mg/L
Phosphate	114	mg/L
Potassium	325	mg/L
Zinc	7.95	mg/L
pH	3.53	
COD	193	mg/L
Ca	936	mg/L
Mg	376	mg/L
HCO_3^-	25	mg/L
SO_4	98	mg/L

SODIUM HYDROXIDE

0.1	L/hr
50	wt/wt%

SHARON REACTOR

Reactions:

1. $NH_4^+ + 1.5O_2 \leftrightharpoons 2H^+ + NO_2^- + H_2O$

Assumptions:

1. Assumed sufficient ammonium oxidising bacteria.
2. No further side reactions

AIR SUPPLY

Required O2:	782	Kg O_2/d
Air density at STP:	1.225	kg/m^3
O2 fraction of air:	0.21	
Oxygen transfer factor	0.25	
Required Air supply at STP:	12 100	m^3/d
	507	m^3/hr

INTERMITTENT STREAM CONCENTRATION:

Parameter	Concentration	Unit
Ammonium	2 250	mg/L
Nitrate	7 250	mg/L
Nitrite	7 500	mg/L

INTERMITTENT STREAM FLOW:

100 293.6 kg/d

REACTION IN ANNAMOX REACTOR:

	NH_4^+	NO_2^-	HCO_3^-	H^+	→	N_2	NO_3	$CH_2O_{0.5}N_{0.15}$	H_2O
Mass	225 000 g/d	758 000 g/d	50 300 g/d	1 620 g/d		339 000 g/d	194 000 g/d	18 900 g/d	433 000 g/d
Mol	12 500 mol/d	16 500 mol/d	824 mol/d	1 620 mol/d		12 100 mol/d	3 080 mol/d	783 mol/d	24 100 mol/d

NITROGEN GAS

339 kg/d

INTERMITTENT STREAM CONCENTRATION:

Parameter	**Concentration**	**Unit**
Ammonium	2 250	mg/L
Nitrate	7 250	mg/L
Nitrite	7 500	mg/L

INTERMITTENT STREAM FLOW:

100 300 kg/d

ANNAMOX REACTOR

Reactions:

1. $NH_4^+ + 1.32NO_2^- + 0.066HCO_3^- + 0.13H^+ \leftrightharpoons 1.02N_2 + 0.26NO_3^- + 0.066CH_2O_{0.5}N_{0.15} + 2.03H_2O$

Assumptions:

1. No reaction inhibitors
2. Steady State operation
3. No further side reactions.
4. 95% conversion

TREATED WATER

EFFLUENT FLOW RATE:

101 m3/d

EFFLUENT CONCENTRATION:

Parameter	**Concentration**	**Unit**
EC	1 480	mS/m
Ammonium	113	mg/L
Nitrate	9 180	mg/L
Nitrite	376	mg/L
Phosphate	114	mg/L
HCO_3^-	24.95	mg/L

SODA ASH DOSING

137 L/d

35 wt/wt%

SODIUM HYDROXIDE DOSING CALCULATION:
Assumptions:
1. Pure solutions
pH = 3.53 = -log10[H^+]
pOH = -log10[OH^+]
pH + pOH = 14 -> pOH = 14 - 3.53 = 10.47
Flow Rate = 100 m^3/d = 4.2 m^3/hr = 4200 L/hr
[H^+] = 1/($10^{3.53}$) = 2.95×10^{-4} mol/L
H^+ molar flow Rate = 4200 × 2.95×10^{-4} = 1.239 mol/hr
Therefore, to bring pH to 7 we require a molar flow rate of NaOH = 1.239 mol/hr
NaOH solution concentration ≈ 50 wt/wt%NaOH
= 50g NaOH/100ml solution
Mw_NaOH = 39.997 g/mol
Molar concentration NaOH solution = 50/39.997
= 1.25 mol/100ml
= 0.0125 mol/ml
= 12.5 mol/l
Therefore, flowrate NaOH = (X)×12.5
= 1.239 mol/hr
Solving for X we can find that the required NaOH solution flow rate = 0.1 L/hr

REQUIRED AIR SUPPLY CALCULATION:
Assumptions:
1. Oxygen demand based on achieving an ammonium to nitrite molar ratio of 1:1.3
2. Oxygen transfer factor of 25%
$NH4_4$-N loading = 5191.35 mg/L × 100 m^3/d
= 519 kgNH_4-N/d
For a NH_4^+:NO_2 molar ratio of 1:1.3 [NH_4^+]/[NO_2] = 0.769
if the initial NH_4^+ and NO_2 concentrations are 5191.35 and 0 mg/L, respectively, then to obtain a NH_4^+:N02 molar ratio of 0.769 a certain amount of NH_4^+must be converted to NO_2 according to reaction 1 (where NH_4^+ is converted to NO_2 on a 1:1 molar ratio). The extent of conversion is arbitrarily name Cf. Then if the initial molar concentration of NH4 = Xi, we can obtain Cf as follows:

$$0.769 = \frac{Xi \times Cf}{Xi \times (1 - Cf)}$$

Solving for Cf, we find that Cf = 0.434. Thus to obtain NH_4^+:NO_2 with a molar ratio of 0.769, (1-0.434) NH_4^+ needs to be converted NO_2.
We can now determine the oxygen requirements for this conversion according to reaction 1 as follows:
519 (1-0.434) × (32 g/mol/ 18 g/mol) × 1.5 = 783 kgO/d
Where, 32 g/mol is the molecular weight of oxygen and 1.5 is the molar ratio of O_2: NH_4^+required to convert one mole of NH_4^+ according to reaction 1.

SODA ASH DOSING CALCULATION:

Assumptions:

1. 35 wt/wt% Soda Ash (Na_2CO_3) solution
2. Carbonate concentration entering reactor ≈ 25 mg/L HCO_3^-

incoming water HCO_3^- loading = 25 * 100 = 2 500 g/d

Required HCO_3^- loading = 50 542 g/d

Therefore, required make-up HCO = 50 542 - 2 500 =

= 48 042 g/d

for a 35 wt/wt% Soda ash solution there is 35g Na_2CO_3 per 100ml of solution. Therefore 48 042/35 = 1372.6 100ml/d

= 137 260 ml/d

= 137, 26 L/d

Therefore 137 L/d 35 wt/wt% soda ash solution required.

A-2: Electrodialysis with Struvite mass balance

RAW WATER

INFLUENT FLOW RATE:		
	100	m^3/d
INFLUENT CONCENTRATION:		
Parameter	**Concentration**	**Unit**
EC	17 900	uS/cm
Estimated TDS (based on EC)	16 200	mg/L
Ammonium	5 190	mg/L
Nitrate	7 250	mg/L
Nitrite	0.49	mg/L
Phosphate	114	mg/L
Potassium	325	mg/L
Zinc	7.95	mg/L
pH	3.53	
COD	193	mg/L
Ca	936	mg/L
Mg	376	mg/L
[OH]		mmol/L
Alkalinity	50	mg/L
HCO3-	25	mg/L
SO4	98	mg/L
estimated Na	650	mg/L
estimated Cl	1350	mg/L

FEED MASS FLOWS

Parameter	Concentration	Unit
Ammonium	519	kg/d
Phosphate	11.4	kg/d
Mg	37.6	kg/d

Anti-scalant Conc.	100 000	mg/L
required concentration in process stream	2	mg/L
Anti-scalant dosing rate	2	L/d

ANTI-SCALANT

ELECTRODIALYSIS STACK

ED1000A-electrodialysis stack		
Cell Pairs	50	
Total membranes	100	
Total membrane surface area	10	m^2
Maximum Flow through cell	10	L/h
Max allowable flow	500	L/h
ED-1000A-stacks required	9	
Approximate anionic separation efficiency	80%	(Zhang et al.)
Approximate cationic separation efficiency	80%	(Zhang et al.)

DILUANT

DILUANT FLOW RATE		
	99.96	m^3/d
DILUANT CONCENTRATIONS		
Parameter	**Concentration**	**Unit**
Approximate TDS	8 560	mg/L
EC	941	mS/m
Ammonium	5 180	mg/L
Nitrate	1 450	mg/L
Na	520	mg/L
Cl	270	mg/L
Phosphate	11.4	mg/L
HCO3-	25	mg/L
Ca	749	mg/L
Mg	358	mg/L

DILUATE MASS FLOWS

Parameter	Concentration	Unit
Ammonium	517	kg/d
Phosphate	1.14	kg/d
Mg	35.8	kg/d

PRECIPITATOR FEED STREAM

RECYCLED STRUVITE SOLUTION

0.1	L/hr
50	wt/wt%

CAUSTIC DOSING

PRECIPITATOR FEED STREAM

FLOW	100	m^3/d
STRUVITE FEED MASS FLOWS		
Parameter	**Concentration**	**Unit**
Magnesium	8.94	kg/d
Ammonium	129	kg/d
phosphate	11.4	kg/d
PRODUCT CONCENTRATIONS		
Magnesium	89.4	mg/L
Ammonium	1 290	mg/L
phosphate	114	mg/L

STRUVITE PRECIPITATOR		
P-Conversion	90%	Sikosana et al. 2014

STRUVITE PRECIPITATOR

$$Mg^{2+} + NH_4^+ + HPO_4^{2-} + 6H_2O \rightleftharpoons MgNH_4PO_4.6H_2O$$

STUVITE PRODUCT

FLOW	0.04	m^3/d
PRODUCT MASS FLOWS		
Parameter	**Concentration**	**Unit**
Magnesium	2.63	kg/d
Ammonium	1.84	kg/d
phosphate	10.3	kg/d
PRODUCT CONCENTRATIONS		
Magnesium	66010	mg/L
Ammonium	46300	mg/L
phosphate	258000	mg/L
Struvite MW	245	g/mol
Mol Struvite produced	108.	mol/d
Struvite mass flow	26.5	kg/d
Moisture content	1.5	g/g
Water mass flow	39.8	kg/d

RECYCLED STRUVITE SOLUTION

RECYCLE MASS FLOWS		
Parameter	**Concentration**	**Unit**
Magnesium	6.31	kg/d
Ammonium	128	kg/d
phosphate	1.14	kg/d
PRODUCT CONCENTRATIONS		
Magnesium	63.2	mg/L
Ammonium	1280	mg/L
phosphate	11.4	mg/L
Recycle rate	1	
Recycle flow	100	m3/d

SODIUM HYDROXIDE DOSING CALCULATION:

Assumptions:

1. Pure solutions

pH = 3.53 = -log10[H^+]

pOH = -log10[OH^+]

pH + pOH = 14 -> pOH = 14 - 3.53 = 10.47

Flow Rate = 100 m^3/d = 4.2 m^3/hr = 4200 L/hr

[H^+] = 1/($10^{3.53}$) = 2.95×10^{-4} mol/L

H^+ molar flow Rate = 4200 × 2.95×10^{-4} = 1.239 mol/hr

If we require to bring pH to 7 we require a molar flow rate of NaOH = 1.239 mol/hr

NaOH solution concentration ≈ 50% NaOH
= 766.5 g/L NaOH

Mw_NaOH = 39.997 g/mol

Molar concentration NaOH solution = 766.5/39.997
= 19.16 mol/1 L

Therefore, flowrate NaOH = (X)×19.16 = 1.239 mol/hr

Solving for X we can find that the required NaOH solution flow rate ≈ 0.1 L/hr

A-3: Combined forward-reverse osmosis mass balance

RAW WATER

INFLUENT FLOW RATE:		
	100	m^3/d
INFLUENT CONCENTRATION:		
Parameter	**Concentration**	**Unit**
EC	17 900	uS/cm
Estimated TDS (based on EC)	16 300	mg/L
Ammonium	5 190	mg/L
Nitrate	7 250	mg/L
Nitrite	0.49	mg/L
Phosphate	114	mg/L
Potassium	325	mg/L
Zinc	7.95	mg/L
pH	3.53	
COD	193	mg/L
Ca	936	mg/L
Mg	376	mg/L
[OH]		mmol/L
Alkalinity	50	mg/L
HCO3-	25	mg/L
SO4	98	mg/L
estimated Na	650	mg/L
estimated Cl	1 350	mg/L

ESTIMATED OSMOTIC PRESSURE		
Parameter	**Concentration**	**Unit**
Estimated osmotic pressure	12.59	bar
MASS FLOWS		
Parameter	**Concentration**	**Unit**
Ammonium	519	kg/d
Nitrate	725	kg/d
Phosphate	11.4	kg/d

FORWARD OSMOSIS UNIT

FORWARD OSMOSIS SYSTEM		
Description	**Value**	**reference**
Concentration factor (Percentage of draw solution)	95%	Fluid Technology Solutions
Phosphate separation efficiency	90%	Valladares Linares Et al.
Ammonia separation efficiency	70%	Valladares Linares Et al.
Nitrate separation efficiency	70%	Valladares Linares Et al.

Concentrated fertilizer stream

FLOWS		
Flow rate	47.5	m^3/d
CONCENTRATIONS		
Parameter	**Concentration**	**Unit**
TDS	33 300	mg/L
EC	3 660	mS/m
Ammonium	7 580	mg/L
Nitrate	10 600	mg/L
Phosphate	216	mg/L
MASS FLOWS		
Parameter	**Concentration**	**Unit**
Ammonium	360	kg/d
Nitrate	505	kg/d
Phosphate	10.3	kg/d

Draw solution: Sea Water

FLOWS		
Flow rate	100	m^3/d
CONCENTRATIONS		
Parameter	**Concentration**	**Unit**
TDS	35 000	mg/L
Ammonium	1.00	mg/L
Nitrate	1.00	mg/L
Phosphate	1.00	mg/L
ESTIMATED OSMOTIC PRESSURE		
Parameter	**Concentration**	**Unit**
Estimated osmotic pressure	27	bar
MASS FLOWS		
Parameter	**Concentration**	**Unit**
Ammonium	0.1	kg/d
Nitrate	0.10	kg/d
Phosphate	0.1	kg/d

DILUTED SEAWATER STREAM

Anti-scalant Conc.	100 000	mg/L
required concentration in process stream	2	mg/L
Anti-scalant dosing rate	3.05	L/d

ANTI-SCALANT

Brine Stream

FLOWS		
Flow rate	91.5	m^3/d
CONCENTRATIONS		
Parameter	**Concentration**	**Unit**
TDS	39000	mg/L
Ammonium	1730	mg/L
Nitrate	2380	mg/L
Phosphate	13.6	mg/L
MASS FLOWS		
Parameter	**Concentration**	**Unit**
Ammonium	158	kg/d
Nitrate	217	kg/d
Phosphate	1.24	kg/d

DILUTED SEAWATER STREAM

FLOWS		
Flow rate	153	m^3/d
CONCENTRATIONS		
Parameter	**Concentration**	**Unit**
TDS	23 300	mg/L
Ammonium	1050	mg/L
Nitrate	1440	mg/L
Phosphate	8.13	mg/L
ESTIMATED OSMOTIC PRESSURE		
Parameter	**Concentration**	**Unit**
Estimated osmotic pressure	18	bar
MASS FLOWS		
Parameter	**Concentration**	**Unit**
Ammonium	159	kg/d
Nitrate	220	kg/d
Phosphate	1.24	kg/d

REVERSE OSMOSIS SYSTEM

REVERSE OSMOSIS SYSTEM		
Description	**Value**	**reference**
Recovery	40%	Rosa simulation

PERMEATE STREAM

FLOWS		
Flow rate	61	m^3/d
CONCENTRATIONS		
Parameter	**Concentration**	**Unit**
TDS	154	mg/L
EC	16.94	mS/m
Ammonium	42.00	mg/L
Nitrate	39.00	mg/L
Phosphate	0.00	mg/L
MASS FLOWS		
Parameter	**Concentration**	**Unit**
Ammonium	2.56	kg/d
Nitrate	2.38	kg/d
Phosphate	0	kg/d

A-4: Reverse Osmosis Simulation

Project Information:

System Details

Feed Flow to Stage 1	6.25 m^3/h	Pass 1 Permeate Flow	2.50 m^3/h	Osmotic Pressure:	
Raw Water Flow to System	6.25 m^3/h	Pass 1 Recovery	39.99 %	Feed	17.90 bar
Feed Pressure	43.80 bar	Feed Temperature	25.0 C	Concentrate	30.07 bar
Flow Factor	0.85	Feed TDS	23504.84 mg/l	Average	23.98 bar
Chem. Dose (100% H_2SO_4)	0.00 mg/l	Number of Elements	4	Average NDP	19.36 bar
Total Active Area	163.50 M^2	Average Pass 1 Flux	15.29 lmh	Power	9.51 kW
Water Classification: Seawater with DOW Ultrafiltration, SDI < 2.5				Specific Energy	3.80 kWh/m^3

Stage	Element	#PV	#Ele	Feed Flow (m^3/h)	Feed Press (bar)	Recirc Flow (m^3/h)	Conc Flow (m^3/h)	Conc Press (bar)	Perm Flow (m^3/h)	Avg Flux (lmh)	Perm Press (bar)	Boost Press (bar)	Perm TDS (mg/l)
1	SW30XHR-440i	1	4	6.25	43.46	0.00	3.75	42.99	2.50	15.29	0.00	0.00	153.94

Pass Streams (mg/l as Ion)

Name	Feed	Adjusted Feed	Concentrate	Permeate	
			Stage 1	Stage 1	Total
$NH_4^+ + NH_3$	1008.25	1031.00	1728.19	41.84	41.84
K	0.00	0.00	0.00	0.00	0.00
Na	7700.00	7700.00	12811.03	31.65	31.65
Mg	0.00	0.00	0.00	0.00	0.00
Ca	61.29	61.29	102.08	0.07	0.07
Sr	0.00	0.00	0.00	0.00	0.00
Ba	0.00	0.00	0.00	0.00	0.00
CO_3	0.00	0.00	0.00	0.00	0.00
HCO_3	0.00	0.00	0.00	0.00	0.00
NO_3	1441.00	1441.00	2375.55	38.85	38.85
Cl	12600.00	12940.55	21522.71	64.26	64.26
F	0.00	0.00	0.00	0.00	0.00
SO_4	331.00	331.00	551.61	0.00	0.00
SiO_2	0.00	0.00	0.00	0.00	0.00
Boron	0.00	0.00	0.00	0.00	0.00
CO_2	0.00	0.00	0.00	0.00	0.00
TDS	23141.55	23504.84	39068.43	153.94	153.94
pH	7.60	7.60	7.60	7.60	7.60

Design Warnings

-None-

Solubility Warnings

-None-

Stage Details

Stage 1 Element	Recovery	Perm Flow (m^3/h)	Perm TDS (mg/l)	Feed Flow (m^3/h)	Feed TDS (mg/l)	Feed Press (bar)
1	0.13	0.81	100.31	6.25	23504.84	43.46
2	0.13	0.69	132.72	5.44	26973.34	43.31
3	0.12	0.56	180.28	4.76	30848.23	43.19
4	0.11	0.44	250.96	4.19	34965.81	43.08

Scaling Calculations

	Raw Water	Adjusted Feed	Concentrate
pH	7.60	7.60	7.60
Langelier Saturation Index	-7.30	-7.30	-6.88
Stiff & Davis Stability Index	-8.06	-8.07	-7.85
Ionic Strength (Molal)	0.40	0.41	0.69
TDS (mg/l)	23141.55	23504.84	39068.43
HCO3	0.00	0.00	0.00
CO2	0.00	0.00	0.00
CO3	0.00	0.00	0.00
CaSO4 (% Saturation)	0.58	0.58	1.04
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	0.00	0.00
Mg(OH)2 (% Saturation)	0.00	0.00	0.00

To balance: 340.55 mg/l Cl added to feed.

A-5: Water Quality Results

Fertilizer plant effluent water quality data obtained from the UCT water quality lab

REPORT: TEST RESULTS

WATER QUALITY LAB

STAFF/STUDENT DETAILS

Surname	Ikumi	Staff/ Student Number	
First Name	David	Tittle	Dr
Research Group/Course/ Department	WQ	Cell Number	

Email address	

TEST RESULTS

	DAM 7
Parameter	
pH	4.12
EC (mS/cm)	21.1
Ortho-Phosphate	85.36
Total Phosphate	114
NO2	0.49
TON	2116
NO3	9373
NH3	0
SO4	98
Mg	376
Ca	935.8
TKN	2320.5
COD	193
VFA	

Fertilizer plant effluent water quality data

ID	Date	NH_4-N(ppm)	NO_3-N(ppm)	TN(ppm)	pH	Cond(uS/cm)	P(ppm)	K(ppm)
Dam 7	15/1/2015	6740	7060	13800	4.53	1712	267	345
Dam 7	21/1/2015	5600	5378	10978	4.81	23510	295	365
Dam 7	28/1/2015	5865	4990	10855	5.08	17340	261	297
Dam 7	3/2/2015	4653	5729	10382	3.24	21890	301	230
Dam 7	9/2/2015	4606	7462	12067	3.17	21560	376	326
Dam 7	16/2/2015	5243	6065	11308	3.34	19160	426	349
Dam 7	3/3/2015	4126	5431	9557	3.32	17410	439	348
Dam 7	12/3/2015	5165	5934	11099	3.31	17510	430	348
Dam 7	17/3/2015	5249	5734	10983	3.47	18690	400	335
Dam 7	31/3/2015	5053	5767	10820	3.79	16820	445	322
Dam 7	7/4/2015	5237	6006	11243	4.01	20890	440	314
Dam 7	22/4/2015	6369	7952	14321	2.2	23900	215	
Dam 7	4/5/2015	4591	5639	10230	2.94	7252	127	
Dam 7	26/5/2015	4856	11468	16324	3.2	18290	585	
Dam 7	26/5/2015	4997	10587	15584	3.2	19310	584	
Dam 7	26/5/2015	4939	10821	15760	3.2	19670	586	
Dam 7	26/5/2015	4964	11299	16263	3.19	19170	580	

APPENDIX B: ECONOMIC ASSESSMENT DATA

B-1: Civil BOQ for option A

ITEM NO.	DESCRIPTION	UNIT	QUANTITY	RATE	AMOUNT
	ASSUMED CONCRETE WASTEWATER TREATMENT STRUCTURE Volume Sharon Reactor = 100 m^3 Volume Anammox reactor = 20.8 m^3 Reactor depths = 2.5m Sharon Reactor Area = 40m^2 Anammox Reactor Area = 8.3 m^2			**ZAR**	**ZAR**
1	100KL Sharon reactor concrete tank	R/kL	100	R40,359	R4 035 900
2	20KL Anammox reactor concrete tank	R/kL	20	R40,359	R807 180
3	Control Room	R/m^2	30	8741.816	R262 254
4	Lab/Chemical Storage room	R/m^2	30	8741.816	R262 254
5	fencing	R/m	59	2000	R118 000
6	Miscellaneous Civil costs including hand railings, pathways, paving, piping, etc	%	20%		R1 097 118
	Sub-total Civil Works				R6 582 707
7	P&Gs	%	20%		R1 316 541.35
	Total including profits and contingencies				R7 899 248.10
8	TAX	%	15%		R1 184 887.22
TOTAL FOR CIVIL WORKS OF SHARON-ANAMMOX WWTP					**R9 084 135.32**

* Rates used for cost estimates of civil works on similar projects

B-2: Mechanical costs for option A

DESCRIPTION	TAG	MODEL/TYPE	QUANTITY	UNIT	UNIT COST	TOTAL COST	REFERENCE
MECHANICALS							
PUMPS							
Feed pump	P-001	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
Transfer pump	P-002	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
Treated water pump	P-003	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
pH correction dosing pump and dosing station	P-004	Grundfos DDC with 500L DTS dosing station	2	No.	21639	R43,278	Data Sheet 011_DTS Dosing station
Nutrient dosing pump and dosing station	P-005	Grundfos DDC with 500L DTS dosing station	2	No.	21639	R43,278	Data Sheet 011_DTS Dosing station
Make-up water pump	P-006	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
BLOWERS/COMPRESSORS							
Aeration blower	C-001	BUSCH SB 200 D2 3.0 kW	2	No.	17157.86	R34,316	Data Sheet 005_Air Supply Blower
MISCELLANEOUS							
Biofilm carrier material			20	m3.	2500	R50,000	
Fine bubble air diffusers			35	No.	337.5	R11,813	Data Sheet 006_EDI disk diffusers
Stainless Steel off gas venting system			1	No.	100000	R100,000	
PIPING							
10% of mechanical equipment costs			10%	sum	R349,524	R34,952	
FRAMEWORK							
10% of mechanical equipment costs			10%	sum	R349,524	R34,952	
SUB-TOTAL:						**R384,477**	
PROCUREMENT FACTOR						25%	
TOTAL:						**R480,596**	

B-3: Electrical costs for option A

ELECTRICALS							
PROCESS INSTRUMENTS							
pH sensor for Sharon reactor		Endress+Hauser Orbisint CPS11D	1	No.	R23,866	R23,866	Data Sheet 001_pH sensor
DO sensor for Sharon reactor			1	No.	R50,000	R50,000	
Flowmeter for influent water line to Sharon reactor		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
Flowmeter for transfer water line from Sharon reactor to Anammox reactor		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
Flowmeter for blower air supply			1	No.	R35,000	R35,000	
pH sensor for Anammox reactor		Endress+Hauser Orbisint CPS11D	1	No.	R23,866	R23,866	Data Sheet 001_pH sensor
DO sensor for Anammox reactor			1	No.	R50,000	R50,000	
ORP sensor for Anammox reactor			1	No.	R25,000	R25,000	
Nitrogen Sensor for Anammox reactor			1	No.	R50,000	R50,000	
Flowmeter for treated water line exiting anammox reactor		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
Level Sensor for Sharon Reactor		Endress+Hauser Micropilot FMR10	1	No.	R7,205	R7,205	Data Sheet 003_tank level sensor
Level Sensor for Anammox Reactor		Endress+Hauser Micropilot FMR10	1	No.	R7,205	R7,205	Data Sheet 003_tank level sensor
CONTROL PANEL							
Motor variable frequency drive for air blower			2	No.	7130	14260	Data Sheet 007_Blower VFD
Motor variable frequency drive for centrifugal feed pumps			2	No.	4860	9720	Data Sheet 008_Pump VFD
Motor variable frequency drive for centrifugal transfer pump			2	No.	4860	9720	Data Sheet 008_Pump VFD
Motor variable frequency drive for centrifugal treated water pump			2	No.	4860	9720	Data Sheet 008_Pump VFD
Soft Starter for make-up water pump			2	No.	2000	4000	
Programmable Logic controller (PLC)- (with additional I/O card and power supply)			1	No.	47890	47890	Data Sheet 009_PLC
Human-Machine Interface (HMI)			1	No.	26670	26670	Data Sheet 010_HMI
Control panel Housing and small items (relays, voltage protectors, transformers, wiring, etc)			1	Sum.	50%	60990	
MISCELLANEOUS							
in-field cable trays and wiring (instrument cables and power cables to be armoured)			1	Sum.	15%	R82,896.56	
SUB-TOTAL:						**R630,799**	
PROCUREMENT FACTOR						25%	
TOTAL:						**R788,498**	

B-4: Civil costs for option B

DESCRIPTION	UNIT	QUANTITY	RATE	AMOUNT
ASSUMED 12m HIGH FACTORY BUILDING Factory area required = 382 m^2			**ZAR**	**ZAR**
382 m^2_ 12m high factory building	R/m^2	356	R 10,897.18	R 3,879,394.76
Miscellaneous Costs for retrofitting and upgrading current factory building to fit process requirements	%	20%		R 775,878.95
Sub-total Civil Requirements				R 4,655,273.71
P&Gs	%	20%		R 931,054.74
Total including profits and contingencies				R 5,586,328.46
TAX	%	15%		R 837,949.27
				R 6,424,277.73

* Recommended R 4700 for brickwork to roof height then cladding and roof sheeting above in 2013 (https://www.gdpindustrialproperty.co.za/what-does-it-cost-to-build-a-factory-or-warehouse-in-cape-town/).
** Costs adjusted for 2018 at 3% inflation per year.
*** A cost increase factor of 2 is further applied to the building costs to allow for un-standard roof height of 12m.

B-5: Mechanical costs for option B

DESCRIPTION	TAG	MODEL/TYPE	QUANTITY	UNIT	UNIT COST	TOTAL COST	REFERENCE
MECHANICALS							
PUMPS							
Feed pump	P-001	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
Transfer pump	P-002	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
Recycle pump	P-003	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
pH correction dosing pump and dosing station	P-004	Grundfos DDC with 500L DTS dosing station	2	No.	21639	R43,278	Data Sheet 011_DTS Dosing station
Anti-scalant dosing pump and dosing station	P-005	Grundfos DDC	2	No.	21639	R43,278	Data Sheet 011_DTS Dosing station
ELECTRODIALYSIS STACK							
ED1000A			9	No.	167717	R1,509,453	Data Sheet 012_Electrodialysis stack
STRUVITE PRECIPITATOR							
Complete-Ostara Pearl 500 unit (including classifying screen, dewatering screen, drier unit and bagging system).			1	No.	24858400	R24,858,400	Data Sheet 013_Ostara reactor
MISCELLANEOUS							
Pre-treatment system			1	sum	50%	R823,069.50	
25 000L Raw water storage tank - Galvanized Steel Panel tank		SBS tanks	1	No.	102279	R102,279	
25 000L effluent storage tank - Galvanized Steel Panel tank		SBS tanks	1	No.	102279	R102,279	
1000L Anti-scalant storage tank - HDPE tank		Jojo Tank	1	No.	1500	R1,500	
PIPING							
10% of mechanical equipment costs			10%	m.	R2,675,267	R267,527	
FRAMEWORK							
10% of mechanical equipment costs			10%	m.	R2,675,267	R267,527	
SUB-TOTAL:						R28,068,720	
PROCUREMENT FACTOR						25%	
TOTAL:						**R35,085,900**	

B-6: Electrical costs for option B

ELECTRICALS							
ELECTRICALS FOR COMPLETE STRUVITE PRECIPITATOR SYSTEM INCLUDED IN MECHANICAL COSTS							
ELECTRICALS FOR ELECTODIALYSIS SYSTEM							
PROCESS INSTRUMENTS							
Flowmeter for influent water into Electrodialysis stack		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
Flow meter for diluate stream into Electrodialysis stack		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
Flowmeter for recycle stream into Electrodialysis stack		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
pH sensor for electrodialysis product stream		Endress+Hauser Orbisint CPS11D	1	No.	R23,866	R23,866	Data Sheet 001_pH sensor
CONTROL PANEL							
Motor variable frequency drive for feed pumps			2	No.	4860	9720	Data Sheet 008_Pump VFD
Motor variable frequency drive for transfer pumps			2	No.	4860	9720	Data Sheet 008_Pump VFD
Motor variable frequency drive for recycle water pump			2	No.	4860	9720	Data Sheet 008_Pump VFD
Programmable Logic controller (PLC)- (with additional I/O card and power supply)			1	No.	27890	27890	Data Sheet 009_PLC
Human-Machine Interface (HMI)			1	No.	16670	16670	Data Sheet 010_HMI
Battery System for electrodialysis stack			9	No.	10000	90000	
Control panel Housing and small items (relays, voltage protectors, transformers, wiring, etc)			1	Sum.	50%	81860	percentage of control panel main components
MISCELLANEOUS							
in-field cable trays and wiring (instrument cables and power cables to be armoured)			1	Sum.	15%	R264,419.25	Percentage of Mechanical and Instrument costs
SUB-TOTAL:						**R626,655**	
PROCUREMENT FACTOR						25%	
TOTAL:						**R783,319**	

B-7: Civil costs for option C

DESCRIPTION	UNIT	QUANTITY	RATE	AMOUNT
STANDARDS 6m HIGH FACTORY BUILDING Factory area required = 202 m^2			**ZAR**	**ZAR**
202 m^2_ standard 6m high factory building	R/m^2	202	R 5,448.59	R 1,100,614.81
Miscellaneous Costs for retrofitting and upgrading current factory building to fit process requirements	%	20%		R 220,122.96
Sub-total Civil Requirements				R 1,320,737.77
P&Gs	%	20%		R 264,147.55
Total including profits and contingencies				R 1,584,885.32
TAX	%	15%		R 237,732.80
				R 1,822,618.12

* Recommended R 4700 for brickwork to roof height then cladding and roof sheeting above in 2013 (https://www.gdpindustrialproperty.co.za/what-does-it-cost-to-build-a-factory-or-warehouse-in-cape-town/).

** Costs adjusted for 2018 at 3% inflation per year.

B-8: Mechanical costs for option C

DESCRIPTION	TAG	MODEL/TYPE	QUANTITY	UNIT	UNIT COST	TOTAL COST	REFERENCE
MECHANICALS							
PUMPS							
fertilizer Feed pump	P-001	WILO MHI 405 1.1 kW	2	No.	8355	R16,710	Data Sheet 004_MHI 405
High pressure feed pump (42bar - 6 m3/hr) (high grade SS)	P-002	Grundfos BMS 17-22 HS-E-C-P-A HS-C 22.5 kW	2	No.	608152	R1,216,304	Data Sheet 014_High pressure feed pump
Sea water make-up pump (high grade SS)	P-003	Grundfos CRT 2-3 A 0.55 kW	2	No.	35000	R70,000	Data Sheet 015_Sea water feed pump
Anti-scalant dosing pump and dosing station	P-004	Grundfos DDC with 500L DTS dosing station	2	No.	21639	R43,278	Data Sheet 011_DTS Dosing station
MEMBRANE SYSTEMS							
FO-OsmoF2O FO-8040-CTA-85-SDS membranes			30	No.	18000	R540,000	Data Sheet 016_ FO membrane
FO-vessels (3 element vessel)			10	No.	30000	R300,000	
RO-membrane SW30XHR-440i			4	No.	12938.8752	R51,756	Data Sheet 017_ RO membrane
RO-vessels (single element vessel - 1000PSI)			4	No.	13005.36	R52,021	
MISCELLANEOUS							
FO Pre-treatment system - 100micron ultrafiltration			1	%	50%	R484,994	
RO Pre-treatment System - 1 micron ultrafiltration			1	%	50%	R681,679.47	
25 000L Raw water storage tank - Galvanized Steel Panel tank		SBS tanks	1	No.	102279	R102,279	
25 000L diluted draw water storage tank - Galvanized Steel Panel tank		SBS tanks	1	No.	102279	R102,279	
1000L Anti-scalant storage tank - HDPE tank		Jojo Tank	1	No.	1500	R1,500	
PIPING							
10% of mechanical equipment costs			10%	m.	R3,662,800	R366,280	
FRAMEWORK							
10% of mechanical equipment costs			10%	m.	R3,662,800	R366,280	
SUB-TOTAL:						R4,395,360	
PROCUREMENT FACTOR						25%	
TOTAL:						**R5,494,201**	

B-9: Electrical costs for option C

ELECTRICALS							
ELECTRICALS FOR ELECTODIALYSIS SYSTEM							
PROCESS INSTRUMENTS							
Flowmeter for influent water into FO vessels		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
Flowmeter for sea water draw solution into FO vessels		Endress+Hauser Promag 10W80 DN50-seawater	1	No.	R50,930	R50,930	Data Sheet 002_water flow meter
Flowmeter for Feed solution into RO vessels		Endress+Hauser Promag 10W80 DN50-seawater	1	No.	R50,930	R50,930	Data Sheet 002_water flow meter
Flowmeter for Brine solution from RO vessels		Endress+Hauser Promag 10W80 DN50-seawater	1	No.	R50,930	R50,930	Data Sheet 002_water flow meter
Flowmeter for Permeate from RO vessels		Endress+Hauser Promag 10W80 DN50	1	No.	R30,930	R30,930	Data Sheet 002_water flow meter
EC sensor for concentrated waste water stream		Endress+Hauser Condumax CLS21D	1	No.	R56,000	R56,000	
EC sensor for draw water stream		Endress+Hauser Condumax CLS21D	1	No.	R56,000	R56,000	
EC sensor for permeate stream		Endress+Hauser Condumax CLS21D	1	No.	R56,000	R56,000	
EC sensor for brine stream		Endress+Hauser Condumax CLS21D	1	No.	R56,000	R56,000	
pH sensor for RO feed stream		Endress+Hauser Orbisint CPS11D	1	No.	R23,866	R23,866	Data Sheet 001_pH sensor
Pressure sensors for RO High pressure pumps		Wika Food grade pressure sensor	4	No.	R12,000	R48,000	
Pressures sensors for Feed pumps		Wika Food grade pressure sensor	2	No.	R12,000	R24,000	
Pressure sensors for Draw solution pumps		Wika Food grade pressure sensor	2	No.	R12,000	R24,000	
CONTROL PANEL							
Motor variable frequency drive for feed pumps			2	No.	4860	9720	Data Sheet 008_Pump VFD
Motor variable frequency drive for draw solution pumps			2	No.	4860	9720	Data Sheet 008_Pump VFD
Programmable Logic controller (PLC)- (with additional I/O card and power supply)			1	No.	39890	39890	Data Sheet 009_PLC
Human-Machine Interface (HMI)			1	No.	16670	16670	Data Sheet 010_HMI
Sensor transmitters			3	No.	22000	66000	
Control panel Housing and small items (relays, voltage protectors, transformers, wiring, etc)			1	Sum.	50%	71000	percentage of control panel main components
MISCELLANEOUS							
in-field cable trays and wiring (instrument cables and power cables to be armoured)			1	Sum.	15%	R285,721.20	Percentage of Mechanical and Instrument costs
SUB-TOTAL:						**R1,057,237**	
PROCUREMENT FACTOR						50%	
TOTAL:						**R1,585,856**	

B-10: ECSA professional fee rates

(2) The following additional design fee shall be applicable to the value of the reinforced concrete and structural steel portions of the **works,** inclusive of the costs of concrete, reinforcing, formwork, structural steel work and any pro-rata preliminary and general amounts. Where structures of identical design are repeated on the same project, the combined costs shall be cumulated for the determination of the cost of the reinforced concrete and structural steel works. In cases where structures require individual design, a separate fee shall be calculated for each structure based on the cost of the reinforced concrete and/or structural steel work for that particular structure.

Cost of the Works		Basis of Fee Calculation	
For projects up to R 320 000		A Lump Sum or on a Time Basis	
Range of brackets			
Where the **cost of the works** exceeds	But does not exceed	Primary Fee	Percentage
R 320,000	R 2,950,000	R 16,000	5,00%
R 2,950,000	R 8,590,000	R 147,500	4,00%
R 8,590,000	R 29,290,000	R 373,100	2,65%
R 29,290,000		R 921,650	1,65%

3.2.1 Civil and structural engineering services pertaining to engineering projects

(1) The basic fee for **normal services** in the disciplines of civil and structural engineering, pertaining to engineering projects, is determined from the table below. The fee for the cost of the works within each bracket is the sum of the primary fee and the percentage applied to the difference between the cost of the works and the amount in the first column of the table for the relevant bracket.

Cost of the Works		Basis of Fee Calculation	
For projects up to R 320 000		A Lump Sum or on a Time Basis	
Range of brackets			
Where the **cost of the works** exceeds	But does not exceed	Primary Fee	Percentage
R 320,000	R 860,000	R 40,000	12,5%
R 860,000	R 4,300,000	R 107,500	10,0%
R 4,300,000	R 16,100,000	R 451,500	8,0%
R 16,100,000	R 64,400,000	R 1,395,500	6,0%
R 64,400,000	R 264,400,000	R 4,293,500	5,5%
R 264,400,000		R 15,293,500	5,0%

3.2.5 Electrical engineering services pertaining to engineering projects

(1) The basic fee for **normal services** in the discipline of electrical engineering, pertaining to engineering projects, is determined from the table below. The fee for the cost of the works within each bracket is the sum of the primary fee and the percentage applied to the difference between the cost of the works and the amount in the first column of the table for the relevant bracket.

Cost of the Works		Basis of Fee Calculation	
For projects up to R 270 000		A Lump Sum or on a Time Basis	
Range of brackets		Primary Fee	Percentage
Where the cost of the works exceeds	But does not exceed		
R 270,000	R 860,000	R 33,750	12,5%
R 860,000	R 2,690,000	R 107,500	10,0%
R 2,690,000	R 8,060,000	R 290,500	8,0%
R 8,060,000	R 18,760,000	R 720,100	7,0%
R 18,760,000	R 34,360,000	R 1,469,100	6,0%
R 34,360,000		R 2,405,100	5,5%

3.2.4 Mechanical engineering services pertaining to engineering projects

(1) The basic fee for **normal services** in the discipline of mechanical engineering, pertaining to engineering projects, is determined from the table below. The fee for the cost of the works within each bracket is the sum of the primary fee and the percentage applied to the difference between the cost of the works and the amount in the first column of the table for the relevant bracket.

Cost of the Works		Basis of Fee Calculation	
For projects up to R 270 000		A Lump Sum or on a Time Basis	
Range of brackets		Primary Fee	Percentage
Where the cost of the works exceeds	But does not exceed		
R 270,000	R 860,000	R 33,750	12,5%
R 860,000	R 3,760,000	R 107,500	10,0%
R 3,760,000	R 10,740,000	R 397,500	8,0%
R 10,740,000	R 23,640,000	R 955,900	7,0%
R 23,640,000	R 42,940,000	R 1,858,900	6,0%
R 42,940,000		R 3,016,900	5,5%

APPENDIX C: EQUIPMENT QUOTES

C-1: Electrodialysis stack quote

Pos.	*Description*	*Qty*	*Unit price in €*	*Total price*
1	**ED 1000A-050** Electrodialysis cell unit, completely assembled, Cell size 300 x 500 mm, 2 Chamber System, active membrane area: 1000 cm2 per membrane, 50 cell pairs. Assembled with standard cation exchange membrane PC-SK and standard anion exchange membrane PC SA, electrodes (anode: Pt/Ir-MMO coated Ti-stretched metal, cathode: stainless steel).	1	9.765,00	9.765,00
2	Shipment South Africa	1	850,00	850,00
			Total	10.615,00 €

C-2: Chemicals quote

Item Description	Quantity	Unit	Price (Ex)	Total (Excl)
SODIUM METABISULPHITE (SMBS) FOOD (SHANDONG KAILONG) BAG 25KG	25	KG	R 10,48	R 262,00
SODIUM HYDROXIDE (CAUSTIC FLAKES) (EAST) BAG 25KG	25	KG	R 12,95	R 323,75
SODIUM HYDROXIDE (CAUSTIC LYE) SOLUTION DC 45% POLYCAN (RETURNABLE) 36KG	36	KG	R 7,64	R 275,04
POLYCAN PROTEA GENERIC EACH	1	EA	R 70,00	R 70,00
HYDROCHLORIC ACID (HCL) SOLN 30-33% INDUSTRIAL POLYCAN (RETURNABLE) 28.5KG	28.5	KG	R 5,00	R 142,50

C-3: Instrumentation quote

Item	Qty.	Unit	Order Code Description	Price ZAR		Total Net ZAR
0010	3	PC	FMR10-1144/0 FMR10-AAQBMWDEWFE2+R7 **Micropilot FMR10** Level, radar, contactless and maintenance-free. Economic device. Application: water based liquids (DC >4). :: Reliable measuring: for changing medias, pressure, temperatures, gas phases.	7,205.00		21,615.00
0010	1	PC	10W80-UC0A1AA0A4AA **Promag 10W80, DN80 3"** Electromagnetic flowmeter	38,811.44 10.00- 34,930.30	 % ZAR	34,930.30

additional specifications

Language	English	
Assign line 1	Volume flow	
Assign line 2	Totalizer 1	
Assign current output	Volume flow	
Current span	4-20 mA HART NAMUR	
Value 20 mA	750.000	dm3/min
Time constant	1.000	s
Assign pulse output	Volume flow	
Pulse value (per pulse)	5.00000	dm3
Pulse width	100.000	ms

Pos.	Qty.	Ordercode Product	MatNo.	SPK	Total list price List price per unit	%	Total net price
1	4	FMX21-AA211PGD15A+NBPO Waterpilot FMX21 SPK: PEK [AA] Approval: Non-hazardous area [2] Output: 4-20mA HART [1] Probe Tube: 316L, d=22mm/0.87in [1P] Sensor Range: 10bar/1MPa/150psi gauge,100mH2O/333ftH2O/4000inH2O [G] Reference Accuracy: Standard [D] Calibration; Unit: Sensor range; mm/mH2O [15] Probe Connection: 100 m cable, shortable, PE [A] Seal: FKM Viton [NB] >>Accessory Mounted: Temperature sensor Pt100, 4-wire [PO] >>Accessory Enclosed: Suspension clamp, 316L	71082058	PEK	73.231,52 ZAR 18.307,88 ZAR	15 %	62.246,79 ZAR
2	4	CLS21D-C1E1** Condumax CLS21D SPK: CFA [C] Measuring Range, Cell Constant: 0.01-20.0mS/cm; k=1 [1E] Process Connection: thread G 1; PES [1] Approval: non hazardous area	71035430	CFA	65.882,56 ZAR 16.470,64 ZAR	15 %	56.000,18 ZAR
3	4	CPS11D-7BT21* Orbisint CPS11D Memosens SPK: CAE [7] Version: Basic version [BT] Application Range: 0-14pH, 0-135oC, 16 bar; with ion trap [2] Shaft Length: 120mm [1] Approval: Non-hazardous area	51513424	CAE	23.866,84 ZAR 5.966,71 ZAR	15 %	20.286,81 ZAR
4	4	CPA250-A00** Flowfit W CPA250 SPK: CAD [A] Application: 3 electrodes [00] Process Connection; Material: Thread G1, PP; PAL SS316	50050049	CAD	29.822,64 ZAR 7.455,66 ZAR	15 %	25.349,24 ZAR
5	8	CYK10-A101** Meas. cable CYK10 Memosens SPK: CAR [A] Approval: Non-hazardous area [10] Cable Length: 10m [1] Cable Connection: Wire terminals	51513497	CAR	27.609,36 ZAR 3.451,17 ZAR	15 %	23.467,96 ZAR
6	4	CM442-AAM2A2F060A Liquiline CM442 SPK: CIA [AA] Approval: Non-hazardous area [M2] Sensor Input: 2x digital sensor	71094218	CIA	104.475,24 ZAR 26.118,81 ZAR	15 %	88.803,95 ZAR
7	3	CM442R-AAM2A2F06 Liquiline CM442R SPK: CIA [AA] Approval: Non-hazardous area [M2] Sensor Input: 2x digital sensor [A2] Communication: 2x output 0/4...20mA, HART [F0] Additional Features: W/o [6] Power Supply: 24VDC	71184862	CIA	62.984,07 ZAR 20.994,69 ZAR	15 %	53.536,46 ZAR
8	1	CM442R-AAM2A2F06+SA Liquiline CM442R SPK: CIA [AA] Approval: Non-hazardous area [M2] Sensor Input: 2x digital sensor [A2] Communication: 2x output 0/4...20mA, HART [F0] Additional Features: W/o [6] Power Supply: 24VDC [SA] >>Accessory Enclosed: External graphical display, cabinetinstallation,plastic,Navigator+Softkeys,incl. cable, IP66	71184862	CIA	26.008,41 ZAR	15 %	22.107,15 ZAR
9	4	RSG35-C2A+C1 Ecograph T RSG35 SPK: RCU [C] Input Signal: 8x universal [2] Power Supply: 24V (-10%; +15%) AC/DC [A] Communication: Ethernet RJ45 + USB [C1] >Memory Medium: SD card industrial grade, 1GB	71157728	RCU	91.532,00 ZAR 22.883,00 ZAR	12 %	80.548,16 ZAR

C- 4: FO-RO quote

Item	Quantity	Unit Price (USD)	Total Price (USD)
OsmoBC™ System includes system, including pre-filters, FO skids, HBCR skids, CIP Skid	1		$ 705,500
Forward Osmosis Elements	100		
HBCR 1 Elements, Pass 1	60		
HBCR 2 Elements, Pass 2,3,4	40		
HBCR 3 Elements	5		
Price for one set of FO and HBCR Elements			$ 285,000
Total Price			**$ 990,500**

C- 5: Pumps and dosing pumps quote

Pos	Product		Qty	Unit Net Price	Total Net Price
10	DTS-500-T-0-0-1-4-WO-A-0-B-1-G	Est. time to shipping: **(+/-) 6 - 8 weeks from date of order** Dosing tank station complete: Medium: Ammonium and Phosphates. (P/N: To be generated upon order) Grundfos 500L DTS inclusive of the following: - DDC 9-7 AR-PP/V/C-F-31I002FG: 97721432 - 500L PE-transparent tank for hand mixer: 98149266 - Handmixer: 98133793 - Multi-Function Valve-G5/8-10 PP/V U2: 95704585 - Filling amature, PVC,E, DN15 : 96727265 - Drain valve PVC 3/4": 96689132 - Assembly charge (450 ZAR/H) * 2hrs	1	21,639.80	21,639.80

GRUNDFOS

Company name: Alveo Water
Created by:
Phone:
Date: 2018/07/14

Order Data:

Product name: BMS 17-22 HS-E-C-P-A HS-C-Tungsten Carbide/Carbon
Amount: 1
Product No: 98467178
Price: 608.152,00 ZAR

Total: 608.152,00 ZAR

C- 6: Ostara 500 Cost estimate

Ostara Example Cost- Benefit Analysis		
Capital Cost of Equipment	$	2.3M
Annual O&M	$/year	~ 250,000
Phosphorus Recovered	Tons/year	6
Recovered Revenue	$/Year	~ 12,000
Savings From Reduced Recycle Load	$/Year	0
Savings from Reduced Struvite Maint	$/Year	0
Return on Investment	Years	50+

*(Napa_Sanitation_District, 2013)

C-7: 20KL Steel panel tank quote

Quote - 10282

QTY	DESCRIPTION	UNIT PRICE EXCL.	VAT	LINE TOTAL EXCL.
1.0	Water Tank Complete with Body Liner & Roof ST06-02 (4.09m ? x 2.14m) 28kl (Gross Capacity)	41142.00	14.00%	41142.00
1.0	Installation - ST06-02	12951.00	14.00%	12951.00
1.0	Fixed Ladder Assembly 2 Ring 600mm Compl. With Ladders Platform & Access Hatch	11636.00	14.00%	11636.00
1.0	Inlet 80NB Side Mount Int Deflector Ext Flange	1716.00	14.00%	1716.00
1.0	Outlet 80NB Side Mount Int Anti-vortex (Standard) Ext Flange	1934.00	14.00%	1934.00
1.0	Overflow 100NB 2 Ring int Bell Mouth ext Downpipe Grooved	1888.00	14.00%	1888.00
1.0	Dump Drain Side Mount ext Deflector - 80NB	1480.00	14.00%	1480.00
1.0	Dump Drain Butterfly Valve Geared - 080NB	871.00	14.00%	871.00
1.0	Water Level Indicator Mechanical - 2 Ring	2581.00	14.00%	2581.00
1.0	Ventilator Static - 76 mm	782.00	14.00%	782.00
1946.0	Tank Transport - ST06-02 Cost/km	13.00	14.00%	25298.00

Total Excl	102279.00	
Total VAT	14319.06	
Grand Total		116598.06

C-8: Water and sanitation rates

Commercial / Industrial Tariffs		
Water Steps (1kl = 1 000 litres)	**Level 4 (2017/18)** Until 31/1/2018 **Rands (incl VAT)**	**Level 6 (2017/18)** From 1/2/2018 **Rands (incl VAT)**
Water	R27, 97	R57
Sanitation (standard)	R21, 50	R44, 18

APPENDIX D: EQUIPMENT DATA SHEETS

D-1: Data Sheet 001_pH sensor

TI00028C/07/EN/14.15
71299913

Products | Solutions | Services

Technical Information
Orbisint CPS11D and CPS11

pH electrodes, analog or with digital Memosens technology

For standard applications in process and environmental technology, with dirt-repellent PTFE diaphragm, built-in temperature sensor (optional for analog sensor)

Application

- Long-term monitoring and limit value monitoring of processes with stable process conditions
  - Chemical industry: strong acids/bases, plastic, pulp and paper industry
  - Power plants (e.g. flue gas cleaning), oil and gas
  - Incinerator plants
- Water/wastewater treatment
  - Boiler feedwater and cooling water
  - Well water and drinking water
  - All industrial and municipal treatment plants

With ATEX, IECEx, FM, CSA, TIIS and NEPSI approval for use in hazardous areas

Your benefits

- Low-maintenance and robust thanks to large PTFE ring junction
- Can be used at pressures up to 17 bar abs. (246 psi)
- Process glass also for very alkaline applications (BA and BT versions)
- Process glass for applications in media containing hydrofluoric acid (FA version)
- For media with low conductivity (AS version)
- Integrated NTC30K temperature sensor (Memosens) for effective temperature compensation; Pt100 or Pt1000 for analog sensors
- Optional: Poison-resistant reference with ion trap

Other advantages of Memosens technology

- Maximum process safety
- Data security thanks to digital data transmission
- Very easy to use as sensor data saved in the sensor
- Recording of sensor load data in the sensor enables predictive maintenance with the Memobase Plus CYZ71D

Endress+Hauser
People for Process Automation

D-2: Data Sheet 002_water flow meter

TI00093D/06/EN/14.17
71265675

Products | Solutions | Services

Technical Information
Proline Promag 10W
Electromagnetic flowmeter

Sensor with degree of protection IP68 (Type 6P enclosure) with a highly cost-effective transmitter

Application

- The bidirectionally measuring principle is virtually independent of pressure, density, temperature and viscosity
- The specialist in the water and wastewater industry for the most demanding applications

Device properties

- International drinking water approvals
- Degree of protection IP68 (Type 6P enclosure)
- 2-line display with push buttons
- Device in compact or remote version
- HART

Your benefits

- Secure, reliable long-term operation – robust and completely welded sensor
- Energy-saving flow measurement – no pressure loss due to cross-section constriction
- Maintenance-free – no moving parts
- Cost-effective – designed for easy applications and direct integration
- Safe operation – display provides easy readable process information
- Fully industry compliant – IEC/EN/NAMUR

Endress+Hauser
People for Process Automation

D-3: Data Sheet 003_Tank Level Sensor

TI01266F/00/EN/04.16
71346464

Technical Information
Micropilot FMR10

Free space radar

Level measurement for liquids

Application

- Ingress protection: IP66/68 / NEMA 4X/6P
- Measuring range: up to 8 m (26.25 ft)
- Process temperature: –40 to 60 °C (–40 to 140 °F)
- Process pressure: –1 to 3 bar (–14 to 43 psi)
- Accuracy: up to ± 5 mm (0.2 in)

Your benefits

- Level measurement for liquids in storage tanks, open basins, pump shafts and canal systems
- Radar measuring device with *Bluetooth®* wireless technology
- Simple, safe and secure wireless remote access – ideal for installation in areas or places difficult to reach
- Commissioning, operation and maintenance via free iOS / Android app SmartBlue – saves time and reduces costs
- Full PVDF body – for a long sensor lifetime
- Hermetically sealed wiring and fully potted electronics – eliminates water ingress and allows operation under harsh environmental conditions
- Most compact radar due to unique radar chip design – fits in limited space installations
- Best price-performance-ratio radar

Endress+Hauser
People for Process Automation

D-4: Data Sheet 004_Feed pump_MHI 405

Contact	
E-mail	
Phone	

Customer

Contact	
E-mail	
Phone	

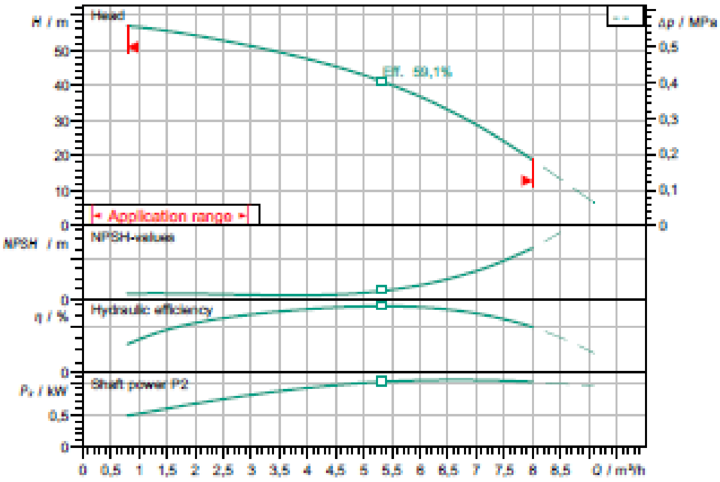
Technical data

High-pressure multistage centrifugal pump
MHI 405-1/E/3-400-50-2

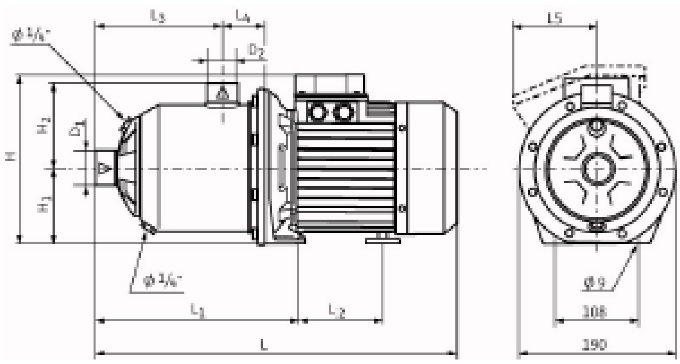
Project name	Untitled project 2018-03-24 10:40:19.327
Project ID	
Installation location	
Customer pos.no	

Date 2018-03-24

Duty chart



Pump curves in accordance with ISO 9906, Appendix A



Requested data

Flow	
Head	
Media	Water 100 %
Fluid temperature	20,00 °C
Density	998,30 kg/m³
Kin. viscosity	1,00 mm²/s

Hydraulic data (Duty point)

Flow	
Head	
Shaft power P2	
Hydraulic efficiency	
NPSH	

Product data

High-pressure multistage centrifugal pump
MHI 405-1/E/3-400-50-2

Max. operating pressure	1 MPa
Inlet pressure max.	6 bar
Fluid temperature	-30 °C ... + 110 °C
Max. ambient temperature	40 °C

Motordata per Motor/Pump

Motor efficiency level	IE3
Mains connection	3~ 400 V / 50 Hz
Permitted voltage tolerance	±10 %
Max. speed	2900 1/min
Rated power P2	1,10 kW
Rated current	2,80 A
Power factor	0,83
Efficiency	
50% / 75% / 100%	81,5/82,7/82,7%
Degree of protection	IP 54
Insulation class	F
Motor protection	no

Fitting dimensions

Pipe connection on the suction side	Rp 1¼, PN 10
Pipe connection on the pressure side	Rp 1, PN 10

Materials

Pump housing	1.4301 [AISI304]
Impeller	1.4301 [AISI304]
Static seal	EPDM
Pump shaft	1.4301 [AISI304]
Mechanical seal	BQ1E3GG

Information for order placements

Weight approx.	15,1 kg
Item number	4210732

D-5: Data Sheet 005_Air Supply blower

Seitenkanalgebläse
Side channel blowers
Soufflantes à canal latéral

BUSCH

Samos
SB 0050 - 1400 D0/D2

Samos SB 0530 D0

Samos SB Seitenkanalgebläse, in einstufiger oder zweistufiger Bauart, sind in allen Bereichen einsetzbar, wo eine pulsationsfreie Förderung des Mediums im Saug- wie im Druckbetrieb gefordert wird.
Einbau in horizontaler und vertikaler Lage möglich. Robuste Bauweise durch Aluminiumdruckgussteile.

Wartungsfrei
durch dauergeschmierte Lager, einen oberflächengekühlten Motor und ein berührungsfrei laufendes Zellenrad.

Umweltfreundlich
durch absolut ölfreie Verdichtung und geräuscharmen Betrieb durch integrierte Schalldämpfer.
Niedriger Energiebedarf.

Side channel blowers Samos SB, single and double stage, are suitable for pressure and vacuum duties and especially suited to applications where a pulsation-free flow is required.
Units can be installed in horizontal and vertical position. Robust construction due to die cast aluminium.

Maintenance-free
Sealed for life bearings, fan-cooled motor and non-contacting impeller ensure maintenance-free equipment.

Environmentally safe
Oil-free compression and low noise level because of internal silencers.
Low power consumption.

Les soufflantes à canal latéral Samos SB, mono et bi-etagées, conviennent pour toutes les applications en vide ou en pression nécessitant un régime non pulsatoire. Montage possible en position verticale ou horizontale.
Construction robuste en fonte d'aluminium coulée sous pression.

Maintenance réduite
Grâce aux roulements graissés à vie, au refroidissement du moteur par ventilation extérieure et à l'absence de friction entre la turbine et son logement.

Respect de l'environnement
Grâce à la compression sans huile et à un niveau sonore réduit dû au silencieux d'échappement intégré.
Consommation électrique réduite.

D-6: Data Sheet 006_Disk Diffusers

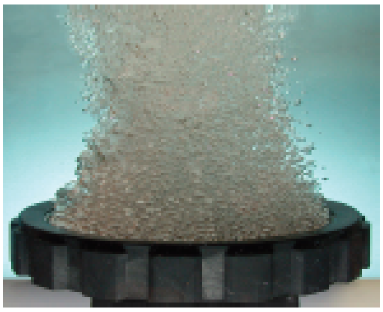
PRODUCT SPECIFICATION SHEET

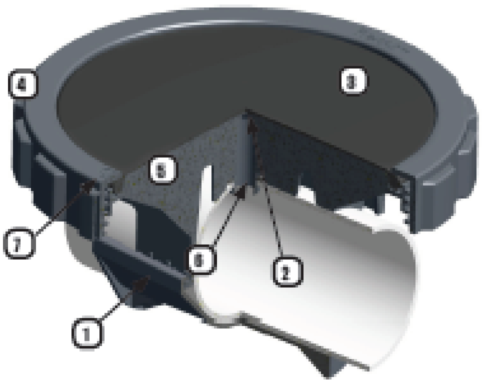
EDI FlexAir® ISM Disc

With Integral Saddle Mount

FlexAir Disc diffuser incorporates EDI's advanced technologies for superior aeration performance, flexibility, and reliability.

- Precision die cut openings for high oxygen transfer, uniform air release, and low operating pressure
- High capacity membrane option available for maximum airflow and low operating pressure
- Advanced technology premium quality membrane materials available in EPDM and special polymer blends
- Full 9 inch (230 mm) and 12 inch (300 mm) of active area
- Triple check valve design prevents entry of liquid/solids into piping. Ideal for on / off applications
- Resistant to fouling and plugging for low maintenance
- Rugged, heavy-duty construction – withstanding over 200 lb (90.7kg) edge load on 9" model
- Glass fiber reinforced polypropylene construction for maximum chemical, temperature, and UV resistance
- Integrated Saddle Mount provides ease of installation and maintenance
- KlicLoc™ retainer for positive mechanical lock
- 9" disc available in *3 inch, 4 inch, 90 mm,* and *110 mm* pipe sizes for design flexibility
- 12" disc available in *3 inch* and *90 mm* pipe sizes
- Mounts on any pipe material (PVC, ABS, CPVC, SS, etc.).
- Standard units IN STOCK for immediate shipment





1. KlicLoc™ Retainer	5. Diffuser Body
2. Primary Check Valve Feature	6. Air Inlet Orifice
3. Flexible Membrane Media	7. EZSeal™
4. Membrane Retainer Ring	

www.wastewater.com

Environmental Dynamics International

aeration for life™

D-7: Data Sheet 007_Blower VFD

SIEMENS

Data sheet for SINAMICS V20

MLFB-Ordering data 6SL3210-5BE23-0CV0

Figure similar

Client order no. :
Order no. :
Offer no. :
Remarks :

Item no. :
Consignment no. :
Project :

Rated data	
Input	
Number of phases	3 AC
Line voltage	380 ... 480 V -15 % +10 %
Line frequency	47 ... 63 Hz
Output	
Number of phases	3 AC
Rated voltage	400 V
Rated power (HO)	3.00 kW / 4.00 hp
Rated power (LO)	3.00 kW / 4.00 hp
Rated current (HO)	7.30 A
Rated current (LO)	7.30 A
Rated current (HO) at 480V	7.30 A
Rated current (LO) at 480V	7.30 A
Pulse frequency	2.00 kHz
Output frequency	0 ... 550 Hz

General tech. specifications	
Power factor λ	0.72
Offset factor cos φ	0.95
Efficiency η	0.98

Ambient conditions	
Cooling	External fan
Installation altitude	1000 m (3281 ft)
Ambient temperature	
Operation	-10 ... 60 °C (14 ... 140 °F)
Storage	-40 ... 70 °C (-40 ... 158 °F)
Relative humidity	
Max. operation	95 %

Communication	
Communication	USS, Modbus RTU

Standards	
Compliance with standards	CE, cULus, C-Tick (RCM), KC
CE marking	EN 61800-5-1 /EN 60204-1 and EN 61800-3

D-8: Data Sheet 008_Pump VFD

SIEMENS

Data sheet for SINAMICS V20

Figure similar

MLFB-Ordering data 6SL3210-5BE21-1CV0

Client order no. :
Order no. :
Offer no. :
Remarks :

Item no. :
Consignment no. :
Project :

Rated data

Input

Number of phases	3 AC
Line voltage	380 ... 480 V -15 % +10 %
Line frequency	47 ... 63 Hz

Output

Number of phases	3 AC
Rated voltage	400 V
Rated power (HO)	1.10 kW / 1.50 hp
Rated power (LO)	1.10 kW / 1.50 hp
Rated current (HO)	3.10 A
Rated current (LO)	3.10 A
Rated current (HO) at 480V	3.10 A
Rated current (LO) at 480V	3.10 A
Pulse frequency	2.00 kHz
Output frequency	0 ... 550 Hz

General tech. specifications

Power factor λ	0.72
Offset factor cos φ	0.95
Efficiency η	0.98

Ambient conditions

Cooling	External fan
Installation altitude	1000 m (3281 ft)

Ambient temperature

Operation	-10 ... 60 °C (14 ... 140 °F)
Storage	-40 ... 70 °C (-40 ... 158 °F)

Relative humidity

Max. operation	95 %

Communication

Communication	USS, Modbus RTU

Standards

Compliance with standards	CE, cULus, C-Tick (RCM), KC
CE marking	EN 61800-5-1 /EN 60204-1 and EN 61800-3

D-9: Data Sheet 009_Programmable Logic Controller (PLC)

SIEMENS

Data sheet 6ES7212-1AE40-0XB0

SIMATIC S7-1200, CPU 1212C, compact CPU, DC/DC/DC, onboard I/O: 8 DI 24 V DC; 6 DO 24 V DC; 2 AI 0-10 V DC, Power supply: DC 20.4-28.8V DC, Program/data memory 75 KB

General information	
Product type designation	CPU 1212C DC/DC/DC
Firmware version	V4.2
Engineering with	
• Programming package	STEP 7 V14 or higher
Supply voltage	
Rated value (DC)	
• 24 V DC	Yes
permissible range, lower limit (DC)	20.4 V
permissible range, upper limit (DC)	28.8 V
Reverse polarity protection	Yes
Load voltage L+	
• Rated value (DC)	24 V
• permissible range, lower limit (DC)	20.4 V
• permissible range, upper limit (DC)	28.8 V
Input current	
Current consumption (rated value)	400 mA; CPU only
Current consumption, max.	1 200 mA; CPU with all expansion modules

D-10: Data Sheet 010_Human Machine Interface (HMI)

SIEMENS

Data sheet 6AV2123-2DB03-0AX0

SIMATIC HMI, KTP400 Basic, Basic Panel, Key/touch operation, 4" TFT display, 65536 colors, PROFINET interface, configurable from WinCC Basic V13/ STEP 7 Basic V13, contains open-source software, which is provided free of charge see enclosed CD

General information	
Product type designation	KTP400 Basic color PN
Display	
Design of display	TFT widescreen display, LED backlighting
Screen diagonal	4.3 in
Display width	95 mm
Display height	53.9 mm
Number of colors	65 536
Resolution (pixels)	
• Horizontal image resolution	480 Pixel
• Vertical image resolution	272 Pixel
Backlighting	
• MTBF backlighting (at 25 °C)	20 000 h
• Backlight dimmable	Yes
Control elements	
Keyboard fonts	
• Function keys	
— Number of function keys	4

D-11: Data Sheet 011_Dosing Stations

DTS - Dosing Tank Stations

GRUNDFOS

General

- Tanks Material PE (LLDPE) UV-stabilised
- Cylindrical Tanks all with threaded M6 inserts for assembly of Smart Digital DDI, DMX221 dosing pumps
- Suction line opening 2" treaded
- Tanks are available „transparent" or black
- Liquid temperature -20°C to 45°C

The Applications

- Disinfection (Chlorine dosing) e.g. Boreholes small water treatment plants swimming pools
- pH Correction (Acid or Alkali) industry and water treatment plants swimming pools
- General chemical dosing

The Benefits

- Fast and easy assembly of tank stations
- Direct assembly of pump & suction line without drilling
- Tidy and protected keeping of level switch signal cable
- Robust and tidy look
- Raised flange for mounting of mixer motor flange
- High wall thickness
- Safe installation on site
- Threaded inserts and bracket kit allow floor mounting
- Matching tank size for many applications.
- (100 / 200 / 500)

D-12: Data Sheet 012_Electrodialysis stack

Electrodialysis

Technical Data of ED Stacks

Stack Type		*Micro-ED*	*ED 64*	*ED 200*	*ED 1000 A*	*ED 1000 H*
Characteristic		*ED*	*ED/EDBM*	*ED*	*cont-ED*	*ED*
Effective Membrane Area	cm^2	8	64	207	950	1050
Membrane Size	cm	6x4	11 x 11	12,5 x 26	30 x 50	30 x 50
Spacer Type	mm	0,45	0,45	0,35/0,45	0,35/0,45	0,35/0,45
Processing Length	cm	2,8	8	20	70	38
Nominal Flowthrough / Cell	l/h	0,5	8	8	10	30
Membranes per Unit	max pcs	25	60	100	200	200
Eff. Membrane Area / Unit	max m^2	0,02	0,38	2,0	20	20

Stack Type		*ED Q380*	*ED Q1600**	*ED 4000 A**	*ED 4000 H**
Characteristic		*ED/EDBM*	*ED/EDBM*	*cont-ED*	*ED*
Effective Membrane Area	cm^2	380	1600	3600	4000
Membrane Size	cm	25 x 25	50 x 50	50 x 100	50 x 100
Spacer Type	mm	0,45	0,45	0,45	0,45
Processing Length	cm	20	40	160	85
Nominal Flowthrough / Cell	l/h	10	25	25	50
Membranes per Unit	max pcs	60	240	400	400
Eff. Membrane Area / Unit	max m^2	2,28	38,4	144	160

Spacer Type		*0,35 mm*	*0,45 mm*	*0,5 mm*
Characteristic		*low process length*	*high chemical stability*	*low pressure drop*
thickness	µm	350	450	500
material	cm	Silicone / Polyester	Silicone / Polypropylene	Silicone / Polypropylene
mesh type		45°	45°	45°

D-13: Data Sheet 013_Ostara Reactor

OSTARA Pearl®

Nutrient Recovery Technology Customized To Meet Your Needs

Converting wastewater treatment plants into resource recovery facilities, Ostara's proprietary Pearl® technology provides cost effective nutrient recovery for reuse.

Pearl® Fluidized Bed Reactor Design

Ostara offers three standard reactor sizes designed for a range of wastewater treatment plants (WWTPs): Pearl® 500, Pearl® 2K, and Pearl® 10K.

Each Pearl reactor is designed based on phosphate mass removal requirements. The Pearl 2K provides nominal orthophosphate (PO_4-P) removal capacity of 550 lbs/day (250 kg/day), which would typically make it suitable for WWTPs ranging in size between 10 to 30 MGD (average dry weather flow). Modular design enables multiple reactors to be installed.

The Pearl reactor has no moving parts and requires minimal maintenance. All other components are industry standard, providing easy access to spares and support.

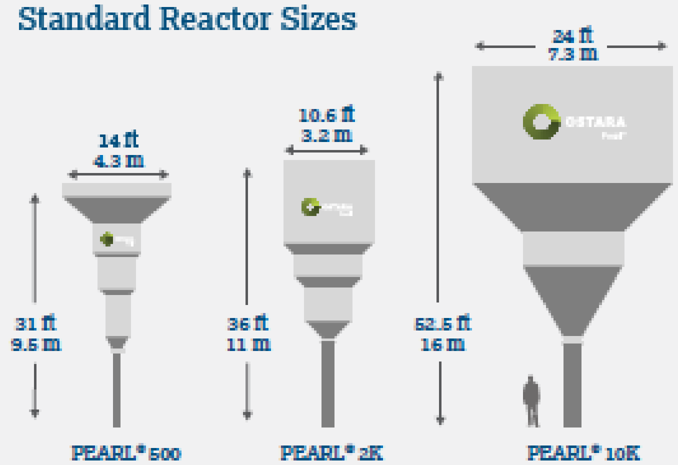
The Pearl® Process

The Pearl process recovers phosphorus from pre- and post-digestion thickening and dewatering liquors through the controlled precipitation of struvite (magnesium ammonium phosphate).

The liquors are fed to the reactor together with magnesium, which is added to maximize struvite precipitation. At the right nutrient concentrations, the crystallization process occurs rapidly.

Like a pearl, the struvite granules grow in diameter resulting in an extremely pure fertilizer marketed and sold as Crystal Green®. Treated effluent is discharged from the top of the reactor and returned to the plant.

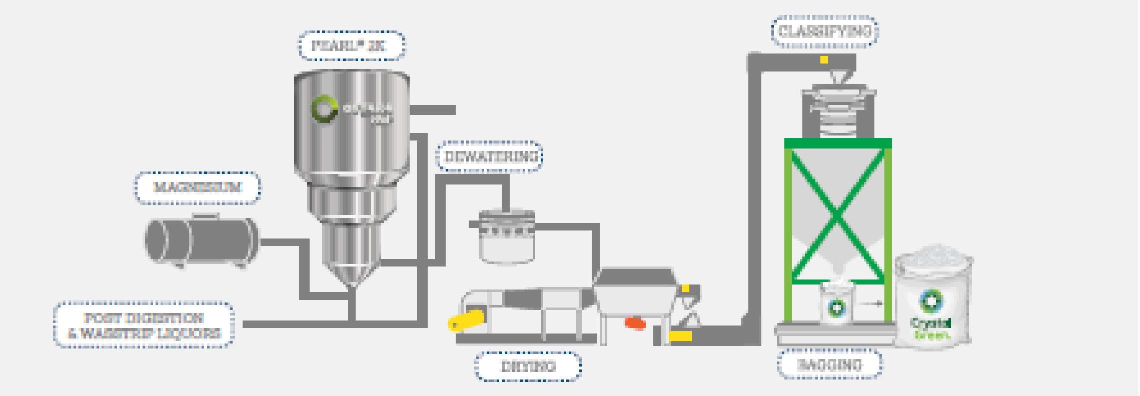
Standard Reactor Sizes



	Pearl® 500	Pearl® 2K	Pearl® 10K
Load Capacity (lbs/kg PO_4-P per day)	145/65	550/250	2,750/1,250
Average Production Capacity (lbs/kg Crystal Green fertilizer per day)	700/325	2,750/1,250	14,000/6,350
Installed Base (2018)	9	17	4

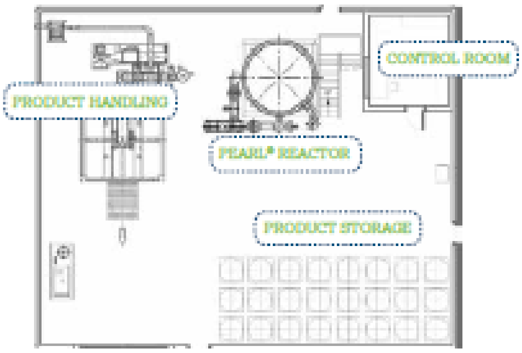
Crystal Green Processing – Simple and Fully Automated

Once granules have reached the desired size, they are removed from the reactor. Finishing occurs automatically in batch-mode, without interrupting the process. Granules are washed as they are conveyed to a dewatering sieve, dried using hot air (which can be recovered from combined heat and power engines or other waste heat sources), then delivered to a classifying screen, before being deposited in silos. The finished product is then ready for bagging in one-ton bulk bags and loading onto trucks for transport to fertilizer customers. As a registered commercial fertilizer, Crystal Green is the only recovered phosphorus ready for reuse as a premium fertilizer directly from WWTPs, with revenue to the plant guaranteed in a long-term offtake agreement.



Flexible, Efficient System Construction

Pearl reactors are provided in standard sizes, utilizing a modular design approach to deliver the required treatment capacity and operational flexibility. System layout is highly flexible, allowing installation into existing buildings where available. System tie-ins include process influent, effluent drainage, power, and non-potable water (e.g. WWTP effluent).



Pearl System Footprint

Reactor Count	Pearl® 500	Pearl® 2K	Pearl® 10K
1	1,500 ft² 140 m²	3,500 ft² 325 m²	6,500 ft² 600 m²
2	-	5,000 ft² 465 m²	8,000 ft² 745 m²
3	-	-	10,000 ft² 930 m²
4	-	-	12,000 ft² 1,115 m²

OSTARA

CrystalGreen.

Ostara Nutrient Recovery Technologies Inc. helps protect precious water resources by changing the way cities around the world manage nutrients in wastewater streams. The company's Pearl® and WASSTRIP® technologies sustainably transform phosphorus and nitrogen recovered from municipal and industrial water treatment facilities into a high-value, eco-friendly fertilizer, sold and marketed by Ostara as Crystal Green®. Crystal Green's unique Root-Activated™ mode-of-action minimizes phosphorus tie-up in the soil, thus enhancing crop yield and performance, and significantly reducing nutrient leaching and runoff. For more information, visit www.ostara.com and www.crystalgreen.com

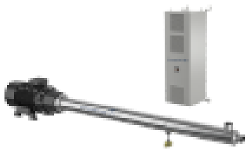
D-14: Data Sheet 014_High Pressure Feed Pump

GRUNDFOS

Company name: Alveo Water
Created by:
Phone:

Date: 2018/07/14

Position	Qty.	Description
	1	**BMS 17-22 HS-E-C-P-A HS-C-Tungsten Carbide/Carbon**



Note! Product picture may differ from actual product

Product No.: 98467178

The BMS hs is designed for use in

- Reverse osmosis systems
- Ultra-filtration systems
- Filtration systems
- Pressure-boosting systems and water supply

A permanent-magnet motor is the secret behind the improved efficiency of the BMS hs range. hs stands for high speed and enables the creation of high pressure of up to 82.7 bar. The BMS hs motor has a speed range of 4000 to 5500 rpm.

The speed of the motor is controlled by a variable frequency drive which features an optional communication module and the possibility of various adaptations to the application. The variable frequency drive is included when ordering the pump.

The high-speed motor also gives the BMS hs range a smaller footprint and drastically reduces the weight of the pump.Max. outlet pressure = 82.7 bar / 1200 psi

Controls:

Frequency converter:	FX75T

Liquid:

Pumped liquid:	Water
Liquid temperature range:	273 .. 313 K
Liquid temperature during operation:	293 K
Density:	998.2 kg/m³

Technical:

Speed for pump data:	5000 rpm
Rated flow:	20 m³/h
Rated head:	651 m
Primary shaft seal:	Tungsten Carbide/Carbon
Approvals on nameplate:	CE, TR
Curve tolerance:	ISO9906:2012 3B

Materials:

Pump housing:	Stainless steel DIN W.-Nr. 1.4539
Impeller:	Stainless steel EN1.4539 AISI 904 L
Motor:	AISI C
Sleeve:	DIN W.-Nr. 1.4462

Installation:

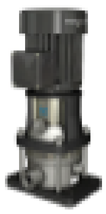
D-15: Data Sheet 015_Sea Water Feed Pump

GRUNDFOS

Company name: Alveo Water
Created by:
Phone:

Date: 2018/07/14

Position	Qty.	Description
	1	**CRT 2-3 A-P-A-V-AUUV**



Product No.: 96100385
Vertical, non-self-priming, multistage, in-line, centrifugal pump for installation in pipe systems and mounting on a foundation.

The pump has the following characteristics:

- Impellers, intermediate chambers and outer sleeve are made of Titanium.
- Pump head cover and base are made of Titanium.
- The shaft seal has assembly length according to EN 12756.
- Power transmission is via cast iron split coupling.
- Pipework connection is via PJE flanges/couplings.

The motor is a 3-phase AC motor.

Liquid:

Pumped liquid:	Water
Liquid temperature range:	253 .. 363 K
Liquid temperature during operation:	293 K
Density:	998.2 kg/m^3

Technical:

Speed for pump data:	3500 rpm
Rated flow:	3 m^3/h
Rated head:	2.594 bar
Primary shaft seal:	AUUV
Approvals on nameplate:	CE,TR
Curve tolerance:	ISO9906:2012 3B

Materials:

Pump housing:	Titanium ASTM B 265
Impeller:	Titanium ASTM B 265
Bush material:	NONE

Installation:

Maximum ambient temperature:	313 K
Max pressure at stated temp:	25 bar / 90 °C 25 bar / -20 °C
Flange standard:	PJE
Pipe connection:	42,4 mm
Flange size for motor:	FT85

Electrical data:

D-16: Data Sheet 016_FO Membrane

Fluid Technology Solutions

OsmoF2O™ FO Industrial Membrane

Model #: **FO-8040-CTA-85-SDS**

OsmoF2O™ FO industrial membrane products utilize low-fouling cellulose membranes. Three types of membrane elements are available:

- One with a spacer for treating low viscous fluids such as clarified juices
- One with a spacer for treating medium viscous fluids such as non-fat milk
- One with an open chevron spacer for treating fouling fluids such as tomato sauce

Each element is designed to be used with a viscous draw solution such as sugars and glycerin. The elements meet sanitary requirements and have a cleanable full-fit outer wrap. The membranes can handle very dirty high-TSS and -TDS wastewaters such as landfill leachate without extensive pretreatment, and require cleaning less frequently. FTS' OsmoF2O™FO industrial membranes are robust and easy-to-clean with high flux recovery over many cleaning cycles, providing long membrane life. The 8040 and 4040 dimensions fit a wide range of conventional membrane module housings.

Features and Benefits

- Fouling-, abrasion-, and chlorine-resistant FO membrane
- Can be used in multiple element housings
- Fouling-resistant feed spacer provide stable FO fluxes
- The SDS standard draw solution spacer (20mil or 0.5mm permeate spacer) for low-viscosity draw solutions

Performance in Food and Beverage Applications

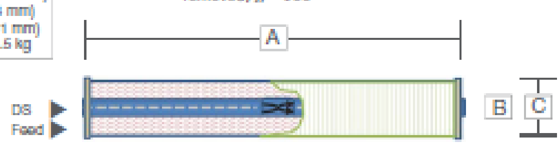
System performance is highly dependent on feed stream characteristics and desired outcomes. FTS works with you to specify the appropriate membrane, system configuration, projected water flux, percentage recovery, and quality of reuse water. Typical applications for the CTA-8040-85 spiral elements include:

- Industrial and process waters and wastewaters
- Municipal and industrial landfill leachate
- Oil and gas wastewaters
- Chemical industry wastewaters high in organic and/or scaling inorganics
- Many other high-fouling wastewaters

Active Area: 139 ft² (13.5 m²)

Draw Solution Salt Rejection: typically, greater than 99.9% rejection of NaCl (draw solution salt) into feed: {1 – [(kg draw transferred to feed) / (kg water removed)]} * 100

A = 40" (1,016 mm)
B = 2.88 (73 mm)
C = 7.9" (201 mm)
Weight – 16.5 kg



OsmoF$_2$O™ FO Industrial Membrane

Model #: FO-8040-CTA-85-SDS

Brief Operating Limits and Guidelines

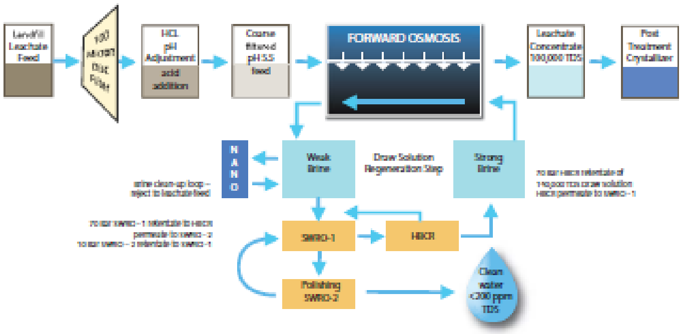
Membrane Requirements	Membrane must be kept moist at all times (do not allow to freeze)
Membrane Type	Cellulose Triacetate (CTA)
Max. Operating Temperature	120°F (50°C)
Max. Side-Port Pressure	75 psi (0.5 MPa)
Minimum Transmembrane Pressure (*)	5 psi (35 kPa)
pH Operating Range	3 to 7
Maximum Chlorine	2 ppm
Maximum Silt Density Index	Very high, can concentrate landfill leachate
Maximum NTU	> 1000
Recommended Pre-filtration	100 µm

(*) Failure to maintain higher pressure on the side ports than the end ports can result in element seam failure, which is not covered under warranty after initial start-up.

Membrane Configurations for Different Applications

FO-CTA-8040-85	For treating high-fouling wastewaters (such as landfill leachate)
FO-CTA-8040-45	For treating moderate-fouling wastewaters (such as dirty seawater and other contaminated brine streams)
FO-CTA-8040-31	For treating low-fouling wastewaters (such as clean brines)
FO-CTA-4040	The elements are identical to the 8040 models but instead of an eight-inch (203 mm) diameter, this model has a four-inch diameter (102 mm). Designed for smaller volumes and specialty applications (such as pharmaceuticals).

Landfill Leachate Treatment with FO-CTA-8040-85 Spiral Elements and HBCR Brine Recovery TFC 8040 Elements



Simplified Process Flow Diagram for the Treatment of Landfill Leachate

D-17: Data Sheet 017_RO membrane

Product Data Sheet

DOW FILMTEC™ SW30XHR–440i Element

Seawater Reverse Osmosis Element with *iLEC*™ Interlocking Endcaps

Description

Dow Water & Process Solutions offers various premium seawater reverse osmosis (RO) elements designed to produce high quality water which may reduce capital and operation costs of desalination systems. DOW FILMTEC™ Elements combine excellent membrane quality with automated precision fabrication to take system performance to exceptional levels.

DOW FILMTEC™ SW30XHR–440i Element are the highest rejection seawater RO elements in the DOW FILMTEC element portfolio, enabling stringent water quality requirements to be met reliably with single-pass seawater systems in most situations. In addition, the combination of highest active area and a thick feed spacer results in higher productivity and lower cleaning frequency which enable sustainable lower lifecycle cost. Benefits of the DOW FILMTEC SW30XHR–440i element include:

- Highest NaCl and boron rejection to help meet World Health Organization (WHO) and other drinking water standards more cost effectively.
- The highest guaranteed active area of 440 ft^2 (41 m^2) permits lowest system cost by maximizing productivity and enables accurate and predictable system design and operating flux.
- The combination of highest active area with a thick feed spacer (28 mil) allows low cleaning frequency and high cleaning efficiency.
- Utilization of the distinct *iLEC*™ Interlocking Endcaps helps reduce system operating costs and reduce the risk of O-ring leaks that can cause poor water quality (see Form No. 609-00446 for information on the cost-saving benefits).
- Sustainable high performance over the operating lifetime, because oxidative treatments are not used in membrane production. This is one reason DOW FILMTEC elements are more durable and may be cleaned more effectively over a wider pH range (1 – 13) than most other RO elements, which use oxidative treatments.
- Effective use in permeate staged seawater desalination systems without impairing the performance of the downstream stage.

Product Type

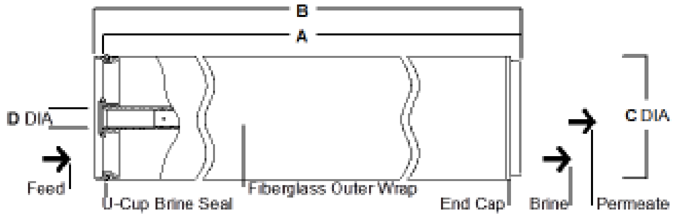
Spiral-wound element with polyamide thin-film composite membrane

Product Specifications

DOW FILMTEC™ Element	Active Area (ft²)	Active Area (m²)	Feed Spacer Thickness (mil)	Permeate Flow Rate (GPD)	Permeate Flow Rate (m³/d)	Stabilized Boron Rejection (%)	Stabilized Salt Rejection (%)
SW30XHR–440i	440	41	28	6,600	25	93	99.82

1. The above benchmark values are based on the following test conditions: 32,000 ppm NaCl, 800 psi (5.5 MPa), 77°F (25°C), pH 8 and 8% recovery.
2. Permeate flows for individual elements may vary ± 20%.
3. Minimum Salt Rejection is 99.7%
4. Stabilized salt rejection is generally achieved within 24 – 48 hours of continuous use; depending upon feedwater characteristics and operating conditions.
5. Product specifications may vary slightly as improvements are implemented.
6. Active area guarenteed ±5 %. Active area as stated by Dow Water & Process Solutions is not comparable to the nominal membrane area figure often stated by some element suppliers. Measurement method described in Form No. 609-00434.

Element Dimensions



DOW FILMTEC™ Element	A (in.)	A (mm)	B (in.)	B (mm)	C (in.)	C (mm)	D (in.)	D (mm)
SW30XHR–440i	40.0	1,016	40.5	1,029	7.9	201	1.125 ID	29 ID

1. Refer to Dow Water & Process Solutions Design Guidelines for multiple-element applications. 1 inch = 25.4 mm
2. Element to fit nominal 8-inch (203-mm) I.D. pressure vessel.
3. Individual elements with *iLEC*™ Interlocking Endcaps measure 40.5 inches (1,029 mm) in length (B). The net length (A) of the elements when connected is 40.0 inches (1,016 mm).

Operating and Cleaning Limits

Maximum Operating Temperature [a b]	113°F (45°C)
Maximum Operating Pressure [b]	1,200 psig (83 bar)
Maximum Element Pressure Drop	15 psig (1 bar)
pH Range, Continuous Operation [a]	2 – 11
pH Range, Short-Term Cleaning (30 min.) [c]	1 – 13
Maximum Feed Silt Density Index (SDI)	SDI 5
Free Chlorine Tolerance [d]	< 0.1 ppm

[a] Maximum temperature for continuous operation above pH 10 is 95°F (35°C).

[b] Operation at pressures up to 1,200 psig (83 bar) is allowable under certain conditions. Consult your Dow representative for advice on applications above 1,000 psig (69 bar) and/or above 95°F (35°C).

[c] Refer to guidelines in "Cleaning Procedures" for more information.

[d] Under certain conditions, the presence of free chlorine and other oxidizing agents will cause premature membrane failure. Since oxidation damage is not covered under warranty, Dow Water & Process Solutions recommends removing residual free chlorine by pretreatment prior to membrane exposure. Please refer to technical bulletin "Dechlorinating Feedwater" for more information.