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The Economic and Technical Feasibility of the Application of Partial Nitritation Anammox Technology Over Conventional Nitrification Denitrification for the Treatment of Sidestream Liquor at Cape Flats WWTP

Master of Engineering in Civil Engineering
Research Dissertation

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ABSTRACT

The objective of this study was to provide a comparative analysis for treating sidestream liquor from a future regional anaerobic digestion (AD) facility (Thermal Hydrolysis Process (THP)+AD) at Cape Flats WWTP in South Africa. The study focused on comparing a conventional nitrification denitrification (Modified Ludzack-Ettinger (MLE) configuration) with a novel partial nitritation/ anammox (PN/A) process. The sidestream liquor was characterised by mass balance over the AD process. Steady-state models were used to predict equipment sizes and determine the process efficiency for each treatment technology. The models were evaluated using an effluent quality index (EQI) and operational cost index (OCI). The capital cost of each treatment solution was calculated and used to perform an economic lifecycle cost analysis (LCCA).

The study concluded that the sidestream liquor (untreated) would recycle 3 415 kg TKN/d back to the mainstream process which represented 29% of the design capacity of the plant. The MLE process required a 53% smaller reactor volume but consumed 93% more energy and produced 64% more sludge which resulted in the MLE OCI being 10.2 times higher than the PN/A process. The total capital cost of the PN/A and MLE processes were calculated to be R139 537 000 and R117 420 000 respectively. Although the MLE process costs 16% less to implement initially, the LCCA over a 20-year operational period indicated that the net present value of the MLE process is 2.8 times higher than the PN/A process.

EXECUTIVE SUMMARY

The City of Cape Town in the Western Cape of South Africa, intends to create a regional digestion facility at Cape Flats wastewater treatment plant (WWTP) to anaerobically digest primary sludge (PS) and waste activated sludge (WAS), pre-treated with the thermal hydrolysis process (THP), imported from various wastewater treatment sites in the Western Cape. Anaerobic digestion of municipal sludge releases nutrients in the form of phosphorus and nitrogen into the soluble phase which, if untreated, are then recycled back to the mainstream WWTP via the sidestream liquor which can lead to overloading of the existing infrastructure. This can lead to poor effluent quality being released and with that, negative social, environmental, and economic impacts to the surrounding areas. This thesis addresses the comparison of technologies to remove nitrogen from sidestream liquor at Cape Flats WWTP.

There are various technologies used to remove nitrogen from sidestream liquor, however at the scale being applied at Cape Flats, biological processes are more feasible. Conventional nitrification denitrifcation (ND), such as a Modified Luzack-Ettinger (MLE) process, is a robust process using activated sludge which is a widely applied and understood technology. In contrast, Partial Nitritation Anammox (PN/A) technology, which has yet to be applied on the African continent, represents a novel technology which has developed in recent years and offers significant operational savings as it short circuits the Nitrification Denitrification process.

A desktop study was conducted to assess the economic and technical feasibility of applying PN/A technology over conventional MLE within a South African context. Based on the anticipated sludge throughput to the future digestion facility and the known unit operations, the anticipated quality of the sidestream liquor was estimated. Steady state models were developed for each process to calculate major equipment sizes, energy consumption and effluent quality. The results of the models were evaluated using an operational cost index (OCI) and an effluent quality index (EQI). The capital and operational cost of each treatment process was then calculated in South Africa Rand and finally a lifecycle cost assessment (LCCA) was conducted.

The nitrogen load in the sidestream liquor was calculated to be 3 415 kg TKN/d. Although the sidestream liquor represents approximately 1-2% of the hydraulic load to the mainstream works, it accounts for 29% of the design total Kjeldahl nitrogen (TKN) treatment capacity. Due to THP being applied to the digestion process, 45% of the soluble chemical oxygen demand (COD) in the sidestream liquor was expected to be biodegradable. The MLE process required a reactor volume 53% smaller than the PN/A process, however consumed 92% more energy, required 5.7 m^3/d of methanol to be dosed for denitrification and produced 64% more sludge. This resulted in the OCI for the MLE process being 10.2 times larger than the PN/A process with 80% of the OCI attributed to organic dosing. Both processes had a similar EQI with MLE being 1% less than the PN/A process. The higher ammonia released by the PN/A process being largely offset by higher COD and TSS from the MLE process. Considering the treated effluent from each process is recycled back to the mainstream plant, the MLE effluent resulted in a 23% and 40% higher aeration energy and sludge production, respectively, when compared to the PN/A process.

The total capital cost of the PN/A and MLE processes were calculated to be R139 537 000 and R117 420 000 respectively. Although the MLE process costs 16% less to implement initially, the LCCA over a 20 year operational period indicated that the net present value of the MLE process is 2.8 times higher than the PN/A process.

Applying the PN/A process at Cape Flats was found to be substantially more cost effective which meetings the challenges currently faced in South Africa. However, significant risks include the availability of seed sludge and the technical expertise available in South Africa to operate the process. These risks can be managed by importing seed sludge from Europe/USA and by applying carefully constructed operation and maintenance contracts with the private sector to enable the skills development within South Africa.

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LIST OF ABBREVIATIONS

Abbreviation	**Units**	**Meaning**
AD	-	Anaerobic Digestion
AAD	-	Advanced Anaerobic Digestion
ANO	-	Autotrophic Nitrifying Organisms
AOR	kg O_2/h	Actual Oxygen Requirement
BPO	mg/l	Biodegradable Particulate Organics
BSO	mg/l	Biodegradable Soluble Organics
CHP	-	Combined Heat and Power
COD	mgCOD/l	Chemical Oxygen Demand
DO	mg/l	Dissolved Oxygen
EQI	-	Effluent Quality Index
FSA	mg N-NH_3/l	Free and saline ammonia
LCCA	-	Life-Cycle Cost Analysis
MLE	-	Modified Luzack-Ettinger
MLSS	kg/m^3	Mixed Liquor Suspended Solids
OCI	-	Operational Cost Index
OP	mg P-PO_4^{3-}/l	Orthophosphate
PAO	-	Polyphosphate Accumulating Organism
PWSSMBM	-	Plant Wide Steady State Mass Balance Model
RBCOD	mg/l	Readily Biodegradable COD
THP	-	Thermal Hydrolysis Process
TIN	mgN/l	Total Inorganic Nitrogen
VSR	mg/l	Volatile Solids Reduction
SOR	kg O_2/h	Standard Oxygen Requirement
SRL	-	Sidestream Return Liquor
TOD	-	Total Oxygen Demand
TSS	mg/l	Total Suspended Solids
UPO	mg/l	Unbiodegradable Particulate Organics
USO	mg/l	Unbiodegradable Soluble Organics
VSS	mg/l	Volatile Suspended Solids
WWTP	-	Wastewater Treatment Plant

LIST OF SYMBOLS

Symbol	Units	Meaning
A	m^2	Area
b_{AT}	mg/mg/d	Specific endogenous mass loss rate of nitrifying bacteria
b_H	/d	Endogenous respiration rate
C_L	mg/l	DO concentration in aeration basin
$C_{p, SRL}$	kg/(KJ °C)	Heat capacity of sidestream liquor
C_s	mg/l	DO oxygen saturation concentration
$D_{p1,RBCOD}$	mgN-NO3/l	Denitrification potential using readily biodegradable COD
$D_{p1,SBCOD}$	mgN-NO3/l	Denitrification potential using slowly biodegradable COD
E_{shaft}	kW	Aeration energy – shaft power
E_{total}	kW	Aeration energy – total
f_{cv}	mgCOD/mgVSS	Ratio of COD to VSS
f_H	-	Endogenous residue fraction
f_n	-	Fraction of nitrogen in VSS
f_p	-	Fraction of nitrogen in VSS
f_{PAO}	-	Fraction of PAOs in WAS
$f_{s,up}$	-	Fraction of unbiodegradable particulate COD within the total COD
$f_{s,up,part}$	-	Fraction of unbiodegradable particulate COD within the total particulate COD
$f_{s,us}$	-	Fraction of unbiodegradable soluble COD within the total COD
$f_{s,us,sol}$	-	Fraction of unbiodegradable soluble COD within the total soluble COD
f_{VSR}	-	% volatile solids reduction
f_x	-	Unaerated mass fraction
FN_t	tN/d	Mass flux of nitrogen
FO_c	kg O_2/d	Carbonaceous oxygen flux
FO_d	kg O_2/d	Nitrification Oxygen flux
FO_n	kg O_2/d	Denitrification oxygen credit
FO_t	kg O_2/d	Total oxygen flux
FP_t	tP/d	Mass flux of phosphorus
FX_{IO}	tDS/d	Mass flow rate of sidestream return liquor
FX_t	tDS/d	Mass flux of sludge

FX_v	tDS/d	Mass flux of volatile solids in sludge
h_c	W/m^2 °C	Convective heat transfer co-efficient
k_n	W/m °C	Thermal conductivity
K_{nT}	mg N-NH_3/l	Nitrification half saturation constant
MX_t	kg TSS	Total sludge mass
MX_{VA}	kg VSS	Total autotrophic granular sludge mass
MX_{VH}	kg VSS	Total heterotrophic sludge mass
N_a	mg N-NH_3/l	Total FSA concentration
N_c	mg N-NO_3/l	Nitrate concentration in the aerobic zone
N_{obpi}	mg N/l	Particulate biodegradable organic nitrogen in the influent
N_{obsi}	mg N/l	Soluble biodegradable organic nitrogen in the influent
N_{org}	mg N /l	Total organic nitrogen concentration
N_{oupi}	mg N/l	Particulate unbiodegradable organic nitrogen in the influent
N_{ouse}	mg N/l	Soluble unbiodegradable organic nitrogen in the effluent
N_s	mg N/l	Nitrogen concentration in the sludge
N_t	mg N/l	Total TKN concentration
P_{ai}	mg P-PO_4^{3-}/l	Total orthophosphate concentration
P_{atm}	mm Hg	atmospheric pressure
P_{org}	mg P/l	Total organic phosphorus concentration
P_t	mg P/l	Total phosphorus concentration
Q	Ml/d	Volumetric flowrate
$Q_{heating}$	kW	Total thermal heat energy
Q_{lost}	kW	Heat lost to the environment
$Q_{sensible}$	kW	Sensible heat
R_{hn}	days	Hydraulic retention time
R_s	days	Sludge age
S_{bi}	mgCOD/l	Biodegradable COD in the influent
S_{bp}	mg COD/l	Biodegradable particulate COD concentration
S_{bs}	mg COD/l	Biodegradable soluble COD concentration
S_f	-	Safety factor for nitrification
S_n	m	Insulation thickness
S_{ti}	mg COD/l	Total COD concentration in the influent
S_{tp}	mg COD/l	Total Particulate COD concentration
S_{up}	mg COD/l	Unbiodegradable particulate COD concentration
S_{us}	mg COD/l	unbiodegradable soluble COD
T	°C	Temperature

$U_{overall}$	W/m^2 °C	Overall heat transfer
V_p	m^3	Reactor process volume
X_{BH}	mgVSS/l	Active OHO concentration
X_{EH}	mgVSS/l	OHO Endogenous residue concentration
X_{Ii}	mg/l	Inert VSS concentration in the influent
X_{IOi}	mg/l	Inorganic concentration in the influent
X_t	mg/l	Total suspended solids concentration
X_v	mgVSS/l	VSS concentration
Y_h	mgVSS/mgCOD	OHO growth yield
α	-	AOR/SOR alpha correction factor
β	-	AOR/SOR beta correction factor
μ_{Am}	/d	Maximum specific growth rate of nitrifiers

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1 INTRODUCTION

1.1 Background

The City of Cape Town intends on creating a regional digestion facility at Cape Flats wastewater treatment plant (WWTP) to anaerobically digest primary sludge (PS) and waste activated sludge (WAS) from various wastewater treatment sites in the Western Cape to align with changes in legislation (Department of Environmental Affairs, 2013) and improve land use. The proposed facility will make use of an advanced anaerobic digestion (AAD) method that includes the thermal hydrolysis process (THP), to pre-treat the sludge which allows digestion to occur at higher percentage solids and higher loading rates thereby increasing the capacity of the existing infrastructure(Higgins *et al.*, 2017).

Anaerobic digestion of municipal sludge releases nutrients in the form of phosphorus and nitrogen into the soluble phase which are then recycled back to the mainstream WWTP via the sidestream liquor. Sidestream liquors have a very low Carbon/Nitrogen (C:N) ratio and the recycled nitrogen load contributes a substantial percentage of the nitrogen load in the incoming wastewater to the WWTP. Jardin *et al*. (2014) indicates that the percentage of nitrogen in the sidestream liquor can be 15 to 20% of the incoming nitrogen load to the WWTP. In the case of a regional digestion facility, whereby sludge and nutrients from surrounding WWTPs are imported to the site, the recycled nitrogen load to the wastewater plant is even higher. In a similar regional digestion facility at Tilburg WWTP, at maximum digestion capacity, the nitrogen load recycled back to the WWTP was 45% of the incoming nitrogen load to the plant (Driessen *et al.*, 2020). In situations where AAD processes are applied, higher sludge throughputs and nitrogen concentrations are achieved when compared with conventional AD (Bungay, 2012), which subsequently, need to be addressed affectively using sidestream treatment technologies.

Cape Flats WWTP currently has eight biological reactors. Each reactor operates as a 5 Stage Bardenpho process that removes nitrogen from wastewater through conventional nitrification denitrification (ND) using activated sludge. Without any additional treatment, the nitrogen load in the sidestream liquor from the proposed THP-AD process recycled back to the existing biological reactors may exceed the treatment capacity of the reactors which will lead to non-compliant effluent. The uncontrolled release of nitrogen (in the form of ammonia and/or nitrates) will lead to the pollution of waterbodies surrounding Cape Flats WWTP which borders on the False Bay Nature Reserve. False Bay Nature Reserve is a sensitive ecological area which is home to a variety of bird and fish species. In addition to this, if the additional nitrogen load is not treated, the regional sludge digestion would need to cut back or completely stop sludge processing. This would result in more sludge being sent to landfill or sacrificial land which puts more pressure on a sludge management system that is already overburdened. Sludge being sent to landfill is then left to anaerobically digest releasing methane into the environment creating an even larger greenhouse gas impact. Therefore, the proposed sidestream technology is crucial as a holistic treatment method for increasing the capacity of the existing treatment facilities.

Environmental damage to the False Bay Nature Reserve will have knock-on social effects by making that area unpleasant or hazardous to use. The nature reserve is used by various bird watching, fishing and water sport societies which will all be under threat. The reserve is also a designated public gathering area, used for family recreational activities. From a political perspective, the release of non-compliant effluent from the biggest wastewater treatment plant in the Western Cape could ruin the City of Cape Town's reputation of clean governance. Further to this, legislation dictates untreated sludge may not be disposed to landfill (Department of Environmental Affairs, 2013). If the regional sludge digestion facility had to cut back on or stop processing sludge due to the additional nitrogen load being placed on the works, this would result in the City of Cape Town being non-compliant to waste management legislation.

Due to the scale of the nitrogen load requiring treatment, the selection of treatment technology is crucial to ensure that the solution is economically feasible. The capital and future operational costs will be financed by the South African taxpayer. In addition to this, if sludge needs to be diverted away from the digestion facility, the City of Cape Town would need to pay more for sludge disposal and possibly incur additional fines. From an economic impact perspective, the stakes could not be higher.

One of the potential technologies for removing nitrogen from the sidestream wastewater is the partial nitritation/anammox (PN/A) process. The PN/A process is a biological process that short circuits the ND process by using autotrophic anaerobic ammonia oxidising (anammox) organisms and ammonia oxidising bacteria (AOB), rather than the conventional autotrophic nitrifying organisms (ANO), and facultative ordinary heterotrophic organisms (OHO). The PN/A process is particularly suited to treating sidestream wastewater with low C:N ratios and has been successfully applied at an industrial scale with more than 100 installations worldwide (Lackner *et al.*, 2014). Short circuiting the ND process and taking advantage of the low growth yield of anammox sludge, results in smaller reactor volumes and lower aeration requirements when compared to conventional ND processes (Gonzalez-Martinez *et al.*, 2018). This results in significant savings in capital and operational expenses. However, Lackner *et al*. (2014) has shown that there are significant challenges to implementing the technology which need to be carefully considered based on site specific limitations. Some of these major challenges relate to a lack of process robustness, complex control strategies and long start up times due to slow anammox growth rates. In to this, the technology has not been implemented on the African continent and, to date, has only been applied in the developed world. This creates additional challenges in terms of access to experience and technical expertise to design and manage the process in South Africa.

Another potential treatment solution is applying a conventional ND process such as a Modified Luzack-Ettinger (MLE) reactor configuration. This solution takes advantage of the robust activated sludge process. However the conventional ND process requires a larger reactor, higher aeration rates, generates more sludge and requires an additional carbon source to be dosed to ensure low nitrogen concentrations in the effluent.. This corresponds to larger capital and operational expenses (Bungay, 2012). From the above, one can appreciate that there are advantages and disadvantages of each process option.

Mathematical modelling of wastewater processes is a powerful tool that can be used to assess the feasibility of various process solutions. There is a vast history of dynamic modelling in the wastewater industry (Gernaey *et al.*, 2004) in which standard performance indices can be introduced to evaluate the models and assist with decision making. Standard performance indices, such as the operational cost index (OCI) and the effluent quality index (EQI), have been developed and successfully applied (Jeppsson *et al.*, 2007; De Ketele, Davister and Ikumi, 2018). Steady state models derived from dynamic model equations have evolved to provide a simpler yet powerful tool for plant design and evaluation making use of user-friendly spreadsheets over advanced modelling software (Ekama, 2011b; Wu, 2015).

The objective of this research is to outline the design of an ND and a PN/A process treating sidestream liquor from a regional digestion facility at Cape Flats WWTP. The research will then make use of steady state models to calculate equipment sizes and effluent qualities. The results of the models will be used to justify the ideal solution when considering the effluent quality returned to the mainstream works. Further to this, the research will compare the total economic life-cycle cost and the technical risks associated for each process operating at Cape Flats WWTP and within a South African environment. The aim of the research is to provide a comprehensive motivation for the most appropriate treatment solution and at the same time, provide a method for the assessment of future sidestream treatment applications. This study represents the first comprehensive assessment of the application of PN/A technology within a South African context.

1.2 Research Questions and Hypothesis

This section lists some of the key the research questions relevant to this study.

- Based on the unit operations (namely THP-AD, struvite precipitation and sludge dewatering), sludge throughput and combination of PS and WAS to the regional digestion facility, what is the

expected sidestream liquor quality in terms of COD, nutrient load, TSS and ISS?
- What are the available technologies to treat sidestream liquor on an industrial scale?
- Narrowing down the technology selection to a biological ND process and a novel PN/A process, what are the existing commercial technologies available and what are the major differences in the process designs?
- Looking at the case specific limitations of Cape Flats WWTP and operating within a South African environment, what is the most appropriate technology to treat side-stream liquor from the regional digestion facility?

By answering the above questions in this research, the following hypothesis was tested:
A PN/A process, although completely novel to South African domestic wastewater treatment, will be a preferred treatment process over a conventional MLE configuration from a process performance and economic perspective.

1.3 Research Objectives

The primary objective of the research was to determine the optimal process (either ND or PN/A) for treating a nitrogen rich digestate stream from a proposed regional digestion facility at Cape Flats WWTP. In determining the optimal process, an economic and technical comparison between a conventional ND process and a novel PN/A process needed to be conducted. The economic evaluation needed to assess life-cycle costing involving capital and operational expenses while the technical evaluation looked at process performance, site specific limitations and limitations in terms of operating within a South African environment.
In doing so, the following specific objectives were addressed:

- Quantify the quality of the sidestream liquor (namely the concentrations of the constituents present) on a desktop study basis by conducting a mass balance over the unit operations for the regional digestion facility.
- Conduct a literature review of commercial PN/A and ND processes and develop a conceptual plant design for both treatment technologies. Development of a steady state process model for each technology to calculate equipment sizes and effluent quality.
- Evaluate the PN/A and MLE processes by comparing the OCI. Determine a capital cost estimate and in conjunction with the OCI use this to calculate the overall economic life-cycle costs for both scenarios to justify the choice from an economic perspective.
- Evaluate the proposed PN/A and MLE processes by comparing the EQI and the effect on the mainstream treatment works.
- Compile a comparative review of the technical aspects and challenges of implementing an MLE or novel PN/A process at Cape Flats WWTP and within the South African environment.

1.4 Research Assumptions and Limitations

This research only focusses on the removal of nitrogen from sidestream liquor. Although high concentrations of phosphorus are generated from AAD processes, Cape Flats WWTP will have a struvite precipitation process where it is assumed that 90% of the OP in the digestate is removed (Zhou *et al.*, 2015).

Another key limitation to this research is that it will only focus on the PN/A and ND (MLE) processes. It was shown in the literature review of Chapter 2 that there are many technologies that can be used to treat sidestream liquor, however, in order for the research to have practical significance, it is simply not possible to research all the processes, within a 60 credit thesis, at the level of detail required. From the literature review it is evident that biological processes are more abundantly applied to treat sidestream effluent of this nature and scale when compared to physical processes. When considering which of the biological processes to focus on for this thesis, the approach was taken to look at the most novel process and compare it against the most conventional process as two extreme cases.

The research is limited to the conditions and processes at Cape Flats WWTP only. Although the methods and findings can be used to provide input and guidance to similar problems at sites around South Africa, site specific conditions need to be assessed when applying these results elsewhere.

Another crucial limitation lies within the choice of modelling methods used in this research. The research will make use of steady state models derived from complex differential equations used in advanced dynamic simulation models such as WEST®, SIMAB etc. By their nature, steady state models cannot be used to predict the results of transient conditions or process upsets but are useful in the design and sizing of such processes, which can later be optimised for performance using dynamic models. Optimisation of the processes is therefore not part of the research. The primary assumption allowing for the chosen process models to be valid, is that the processes are operated at a steady state (i.e., the processes are subjected to a constant load). Considering that the sidestream treatment occurs outside of the mainstream treatment process, and is therefore not subjected to the same level of diurnal variations, this assumption is valid.

This research was performed as a desktop study in which no lab or pilot testing was conducted. The research relies on the accuracy of the information and assumptions backed up by relevant credible literature.

It is crucial to note that some key assumptions were made in conducting the research:

- A key part of the research lies in quantifying the expected nitrogen load to make the assessment between each of the sidestream technologies. Since the digestion process does not exist yet, no physical sampling can occur and certain critical assumptions had to be made to quantify the constituent concentrations in the sidestream liquor. All the results are therefore limited by the accuracy of these assumptions which are listed in more detail in Section 3.1.
- Energy and chemical cost data were assumed for determining the operational cost and were based on quotations obtained at the time the research was undertaken. These expenses had to be escalated at an assumed rate for the purposes of life-cycle costing. More details on these assumptions are listed in Section 3.5 and 3.6.
- For designing the MLE process, an activated sludge steady state model was used to size major equipment. There were many critical assumptions in this model in terms of kinetic rate constants, growth yields, critical wastewater fractions, sludge age etc. More details on these assumptions are listed in Section 3.2.2
- For designing the PN/A process, the activated sludge model was modified to include anammox and partial nitritation stoichiometry at steady state. An ammonia removal efficiency and a total inorganic nitrogen (TIN) removal efficiency was assumed based on literature values. This was used to size major equipment items. More details on these assumptions are listed in Section 3.2.1
- For both plant designs, it will be assumed that the aeration equipment will consist of fine bubble diffused aeration (FBDA) and centrifugal turbo blowers. In converting from the biological actual oxygen requirement (AOR) into a standard oxygen requirement (SOR), various factors such as alpha (oxygen transfer correction factor) and beta (salinity-surface tension correction factor) parameters were assumed. A standard oxygen transfer efficiency was also assumed. These assumptions were identical for both plant designs. More details on these assumptions are listed in Section 3.2.1 and 3.2.2

1.5 Thesis Overview

This section provides a high level overview of the thesis which is divided into six chapters:

- Introduction
- Literature Review
- Research Methodology
- Results and Discussion
- Conclusions
- Recommendations and Future Research

Chapter 2 provides a detailed literature review. Figure 1 below illustrates a conceptual framework which indicates the major concepts explored in the literature review.

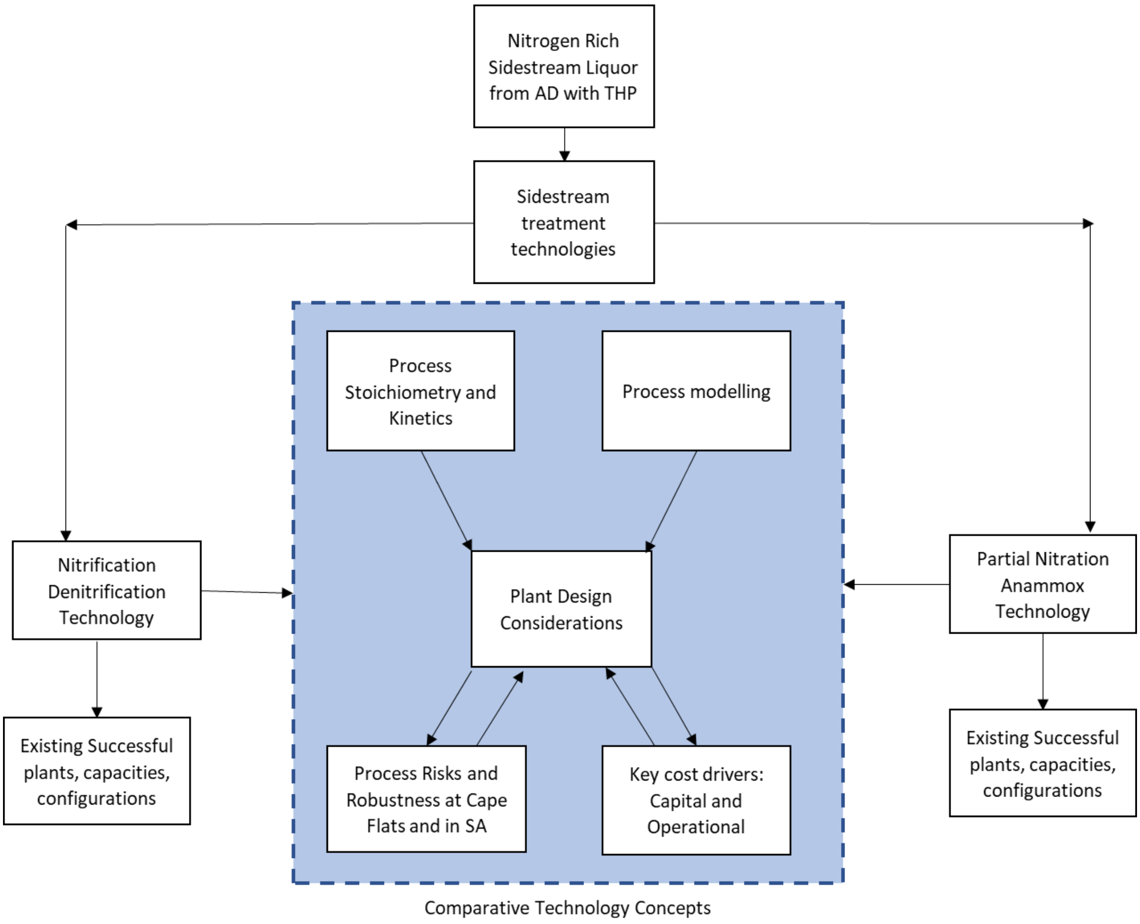

Figure 1.1: A conceptual framework for comparing a PN/A process against a conventional ND process to treat digestate sidestream wastewater.

Figure 1.1 shows that the nutrient rich, sidestream liquor is the key driving force for the research conducted in this thesis. This sidestream liquor will be further explored in terms of defining what typical nutrient loads are expected from AAD with THP and what other critical quality parameters must be considered. This investigation aims to put the scale of the nutrient burden in the sidestream liquor at Cape Flats WWTP into context.

Another important concept will be investigating commercially operated ND and PN/A processes and compare plant configurations and treatment capacities. This will form the baseline for the conceptual designs for either treatment method by building on the existing body of experience and knowledge of these processes.

Figure 1.1 illustrates that five common technological concepts were explored for each treatment process with the aim of determining the most feasible technology to apply at the Cape Flats WWTP when considering the technical, economic and process risk factors. Process stoichiometry, kinetics and modelling formed the basis of plant design and defined the values used in the mass and energy balances which ultimately lead to equipment sizing and selection. Once conceptual plant designs have been determined, the key cost drivers (both capital and operational) are explored.

Chapter 3 presents the detailed methodology used to achieve the research objectives. This section discusses how the design basis for the sidestream liquor was determined which serves as inputs to the process models. It then goes on to discuss the development of the steady-state models and lists all equations and assumptions made. Following on from this, this section discusses the evaluation of the model results using tools such as an operational cost index (OCI), effluent quality index (EQI) and a lifecycle cost analysis (LCCA).

Chapter 4 presents the results and provides discussion around the sidestream liquor calculated at a desktop level. It then goes on to compare the MLE and PN/A model outputs in terms of major equipment sizes, heating and electrical energy requirements, chemical consumption and effluent quality (EQI index). The capital and operational cost estimates are then presented which is together with the results of the LCCA. Lastly, the appropriateness of each process is discussed within a South African context by considering the state of the South African economy, energy and wastewater industry.

Chapter 5 consolidates the findings of Chapter 4 and presents the conclusions of the research. Chapter 6 presents recommendations and opportunities for future research based on the limitations in this research or additional/unanswered questions.

2 LITERATURE REVIEW

2.1 Sidestream liquor from anaerobic digestion

Sidestream liquor is a wastewater stream characterized by a low carbon to nitrogen ratio and a high nitrogen concentration produced from the dewatering of digestate from anaerobic digestion (Eskicioglu, Galvagno and Cimon, 2018). In the anaerobic digestion process, carbon is removed from the digestate (via the breakdown of biodegradable organics, producing methane and carbon dioxide gas) and nutrients (nitrogen and phosphorus) are released into the wastewater (Ikumi *et al.*, 2019).

The application of AAD processes such as THP, allows intensification of existing infrastructure which increases the amount of solids digested per volume of digester capacity (Higgins *et al.*, 2017). This increase in digester capacity further increases the concentration of nutrients in the digestate from AAD processes when compared to conventional AD processes. The aqueous nitrogen concentration in sidestream liquor for conventional AD processes is typically between 500-1000 mg N/l whereas for AAD processes, concentrations range between 1500 – 3500 mg N/l (Bungay, 2012; Eskicioglu, Galvagno and Cimon, 2018). Table 2.1and Table 2.2 show typical values for sidestream liquor from conventional AD and THP AAD, taken from separate studies (Zhang *et al.*, 2016; Gu *et al.*, 2018).

Table 2.1: A comparison of AD and THP-AD sidestream liquor chemical properties.

Parameters	Unit	THP-AD digestate	AD Digestate
pH	-	8.1 ± 0.2	8.2 ± 0.2
NH_4^+	mg N/L	1781 ± 188	1068 ± 138
NO_2^-	mg N/L	< 0.5	< 0.5
NO_3^-	mg N/L	9.0 ± 3.2	1.58 ± 0.5
OP	mg P/L	19.8 ± 10.4	30.0 ± 9.6
COD_{total}	mg COD/L	4242 ± 1048	369 ± 63
$COD_{soluble}$	mg COD/L	3583 ± 1030	300 ± 63
$COD_{particulate}$	mg COD/L	800 ± 508	< 30
TSS	mg/L	283 ± 217	49 ± 20
Alkalinity	mg/L as $CaCO_3$	6278 ± 1119	3852 ± 749
Acetic Acid	mg COD/L	25 - 5000	ND

*Table displayed as per Zhang *et al.* (2016)

Table 2.2: A comparison of conventional AD (regular sludge - RS) and THP-AD (THP sludge – THPS) sidestream liquor wastewater parameters.

Parameters	Unit	THP-AD digestate	AD Digestate
NH_4^+	mg N/L	851.7 ± 95.3	229.4 ± 32.9
NO_2^-	mg N/L	0.1 ± 0.1	0.9 ± 0.1
NO_3^-	mg N/L	0.4 ± 0.2	0.3 ± 0.1
TP	mg P/L	88 ± 9	82 ± 7
COD_{total}	mg COD/L	3290 ± 640	841 ± 186
BOD_5	mg COD/L	834.6 ± 162.5	162.5 ± 78.3
Alkalinity	mg/L as $CaCO_3$	5000 ± 1280	3260 ± 670

*Table displayed as per Gu *et al.* (2018)

Both studies indicate significantly larger nitrogen concentrations for THP before AD. In addition to this, Table 2.1 indicates that THP processes produce higher concentrations of soluble COD and particulate (colloidal) COD making dewatering more challenging which generally results in a higher TSS in the sidestream liquor than conventional AD. Gu *et al.* (2018) found that the THP process increased the biological degradation of the sludge and resulted in a higher fraction of soluble biodegradable COD in the sidestream liquor. This is supported by Figdore *et al.* (2011) who found that the fraction of soluble COD that is biodegradable can be up to a fraction of 0.45 of the total COD. It is also noted that THP produces a larger fraction of soluble unbiodegradable COD than conventional AD. Both Table 2.1 and Table 2.2 indicate substantially high alkalinities in the THP sidestream which are in the order of 3750-7500 mg/l as $CaCO_3$.

Due to the characteristics of sidestream liquor (high nitrogen and phosphorus content), it cannot be released to the environment and is typically recycled back to the head of the WWTP. This practice results in a significant amount of nutrients being recycled back to the existing wastewater works. Jardin *et al.* (2014) and Ikumi *et al.* (2019) indicate that the percentage of nitrogen in the sidestream liquor can be 15 to 20% of the incoming nitrogen load to the WWTP. This is supported by Eskicioglu *et al.* (2018) who indicates that sidestream liquor contains up to 25% of the nitrogen load to the plant while consisting of only 2% of the total flow. The recycling of nutrients can have a significant impact on the mainstream works. In a recent study done by Ikumi *et al.* (2019), the effect of returning sidestream liquor, from the existing digestion process (conventional mesophilic digestion), back to Cape Flats WWTP was assessed. In this study it was determined that returning sidestream liquor would result in an increase of 19.1% and 125% of ammonia and phosphate loads to the mainstream plant, respectively. This was due to the current on site anaerobic digestion of WAS and PS. The effect of returning the sidestream liquor back to the head of the plant resulted in an increase in aeration (5.7%) and a decrease in the nitrate recycle (A-recycle) for the mainstream activated sludge reactors (-76%). Due to the denitrification potential remaining the same, returning sidestream liquor untreated, results in a 5.7% increase in nitrates concentration in the effluent. No effect to COD, orthophosphates or ammonia concentrations in the effluent was noted. In the case of a regional digestion facility implementing AAD whereby sludge is imported to the site, the percentage of the nitrogen load being recycled will be even higher. In a study by Driessen et al. (2020), at a regional digestion facility at Tilburg WWTP, at maximum digestion capacity, the nitrogen load recycled back to the WWTP was 45% of the incoming nitrogen load to the plant. Existing WWTPs with AD processes are often designed to handle the additional nitrogen load and accommodate the nitrification and denitrification process. However, with the goal of intensification of existing digester capacity through AAD, and creating a regional digestion facility, the new nitrogen load on the existing WWTP will far exceed the existing design capacity. If no additional treatment is applied to the sidestream liquor, this can result in an increase in total nitrogen concentration in the mainstream effluent leading to non-compliance and potential damage to the environment.

2.2 Current technologies to remove ammonia from sidestream effluents

There are a multitude of technologies available for removing ammonia from sidestream effluent generated from the anaerobic digestion of municipal sludge. They can broadly be separated into two groups (Eskicioglu, Galvagno and Cimon, 2018):

- Physical and chemical treatment
- Biological treatment

2.2.1 Physical and chemical treatment technologies

Physical and chemical treatment technologies include technologies such as ammonia stripping, struvite precipitation, membrane filtration and ion exchange. Physical and chemical treatment technologies such as ammonia stripping, membrane filtration and ion exchange show promise for the future, particularly when it comes to the recovery (and not just removal) of nutrients from wastewater, however these technologies either require more research, or are typically more expensive than biological treatment when applied commercially (Eskicioglu, Galvagno and Cimon, 2018). Struvite precipitation technology is well established commercially whereby chemical precipitation of the struvite mineral ($MgNH_4PO_4.6H_2O$) removes both phosphorus and nitrogen from wastewater. However, the main objective of this technology is to remove soluble phosphorus and to prevent unwanted precipitation of struvite in piping and mechanical equipment which can lead to blockages and downtime. Due to the costs associated with the high consumption of magnesium and phosphate salts, this technology alone is typically not feasible to reach the ammonia removal target required to release sidestream wastewater back to the mainstream wastewater works (Huang, Song and Xu, 2011).

2.2.2 Biological treatment technologies

Biological treatment includes technologies such as Nitrification Denitrification (ND), Nitritation Denitritation and Partial Nitritation Anammox (PN/A) (Eskicioglu, Galvagno and Cimon, 2018).

2.2.2.1 Nitrification Denitrification

An ND process is a two-step process whereby ammonia is first completely nitrified to form nitrate by autotrophic nitrifying organisms (ANO), consisting of ammonia oxidising bacteria (AOB) and nitrite oxidising bacteria (NOB), followed by nitrate being denitrified to nitrogen gas by facultative ordinary heterotrophic organisms (OHO). The process stoichiometry is as follows (Bungay, 2012)
:

$$NH_4{}^+ + 2O_2 \rightarrow NO_3{}^- + H_2O + 2H^+ \tag{2.1}$$

$$5CH_3OH + 6NO_3 \rightarrow 3N_2 + 5CO_2 + 7H_2O + 6OH^- \tag{2.2}$$

Equation 2.1 is performed by slow growing autotrophic organisms with low growth yields under aerobic conditions. Equation 2.2 is performed by fast growing heterotrophic organisms with high growth yields, under anoxic conditions (low oxygen and high nitrate), and in the presence of COD as an electron donor. In Equation 2.2, the COD source is indicated as methanol but other sources of COD, such as glycerol, can also be used (Bungay, 2012). Glycerol has the advantage over methanol in that it is does not generate explosive atmospheres which presents a much lower operational risk as well as a significantly lower capital investment cost to install a storage and dosing plant as the equipment need not be explosion proof.

Crude glycerol is a byproduct of biodiesel production. With the rapid increase in biodiesel production in the mid 2000's, a surplus of crude glycerol was created. The use of crude glycerol in wastewater treatment to improve dentification is seen as value adding opportunity, however, the quality of crude glycerol can vary substantially depending on the biodiesel production process which could limit its application (Yang, Hanna and Sun, 2012). Crude glycerol from biodiesel waste has already been used successfully for an MLE process treating landfill leachate at Vissershok (Cape Town, South Africa). However, the biodiesel market in South Africa is relatively small and the ability to reliably supply the quantity required for Cape Flats WWTP would need to be verified.

Conventional ND processes are well established commercially and known to be robust. The typical reactor configuration is a Modified Luzack-Ettinger (MLE) process (refer to Figure 2.1) which involves an anoxic reactor upstream of an aerobic reactor. Equation 2.1 occurs in the aerobic reactor to generate nitrates as ANOs are obligatory aerobes. The nitrates are then recycled back to the anoxic reactor whereby Equation 2.2 proceeds. When compared to the other biological treatment processes, ND processes are characterized by higher aeration and energy requirements, greater sludge generation (due to high biomass yields) and require an additional source of COD to ensure low nitrates in the effluent. Lackner et.al. (2014) have indicated that a commercial ND process operating at Ingolstadt (Bavaria, Germany) achieved an energy consumption of 4 kWh/kg N treated. A high-rate MLE process (AMTREAT® Process) operating at elevated temperatures, was developed in the 1990s and has been applied commercially. The benefit of this over conventional ND processes is that the plant is able to achieve the same performance with a much lower plant footprint with the aim of significantly reducing the capital cost of the plant while still achieving the process robustness associated with ND processes (Bungay, 2012).

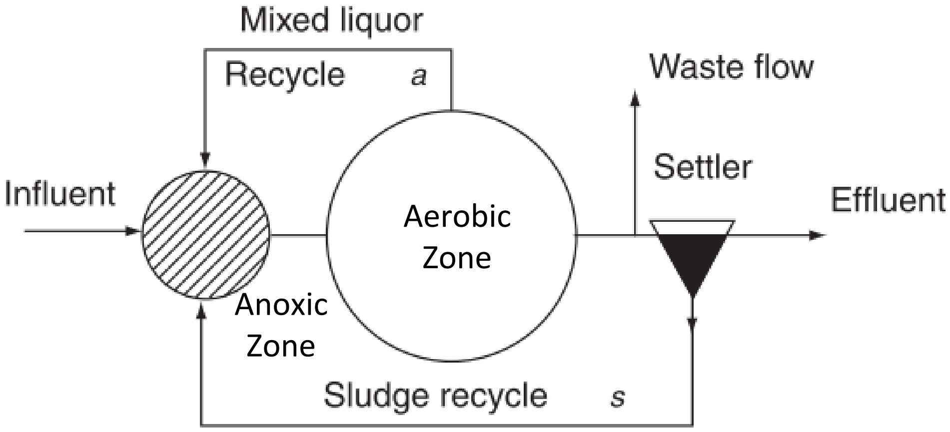

Figure 2.1: Schematic of an MLE reactor configuration (Ekama, 2011b).

2.2.2.2 Nitritation Denitritation

Nitritation is the process whereby ammonia is converted to nitrite by AOBs. Denitritation is similar to denitrification, whereby nitrite is converted to nitrogen gas. The process can be characterized by the following stoichiometric reactions(Bungay, 2012):

$$2NH_4^+ + 3O_2 \rightarrow 2NO_2{}^- + 2H_2O + 4H^+ \qquad (2.3)$$

$$3CH_3OH + 6NO_2^- \rightarrow 3N_2 + 5CO_2 + 3H_2O + 6OH^- \qquad (2.4)$$

Equation 2.3 occurs in a SHARON (Single reactor system for High activity Ammonium Removal Over Nitrite) reactor at elevated temperatures (35°C) and short sludge ages which favours the growth of Ammonia Oxidizing Bacteria (AOB) over Nitrite Oxidizing Bacteria (NOB) (Eskicioglu, Galvagno and Cimon, 2018) (Ikumi *et al.*, 2019). This effectively shortcuts the nitrification process resulting in a significantly lower (25% reduction) aeration requirement when compared to ND processes. Denitritation (indicated in Equation 2.4) does not occur in this process unless a second anoxic tank is included in the design in which case, an additional COD source (such as methanol) is dosed (Bungay, 2012).

2.2.2.3 Partial Nitritation Anammox

A PN/A process involves partial nitritation of a portion of the ammonia to nitrite followed by the anaerobic oxidation of the remaining ammonia with nitrite to from nitrogen gas. The process stoichiometry is as follows (Eskicioglu, Galvagno and Cimon, 2018):

$$NH_4{}^+ + HCO_3{}^- + 0.75O_2 \rightarrow 0.5NH_4{}^+ + 0.5NO_2{}^- + CO_2 + 1.5H_2O \tag{2.5}$$

$$\begin{aligned} &NH_4{}^+ + 1.32NO_2{}^- + 0.066HCO_3{}^- + 0.13H^+ \\ &\qquad \rightarrow 1.02N_2 + 0.26NO_3{}^- + 0.066CH_2O_{0.5}N_{0.15} + 2.03H_2O \end{aligned} \tag{2.6}$$

Similar to an ND process, Equation 2.5 is carried out by NOBs, however, Equation 2.6 is carried out by autotrophic anammox organisms. Both biological reactions are characterized by slow growth rates and low growth yields when compared to the growth rates of heterotrophic organisms. Typical growth yields for nitrifying bacteria are 0.1 g VSS/g NH_3-N (Ekama, 2011b) and for anammox bacteria values of 0.11 VSS/g NH_3-N (Jin *et al.*, 2012) and 0.13 g VSS/g NH_3-N (Van Hulle *et al.*, 2010) have been reported. The PN/A process has been found to be optimal in the temperature range of 30-40°C. The temperature, along with other factors such as the dissolved oxygen (DO) concentration allow for the AOBs to outcompete the NOBs for substrate, subsequently, allowing for the production of nitrite over nitrate (Van Hulle *et al.*, 2010); (Szatkowska and Paulsrud, 2014).

From Equation 2.5 one can see that ammonia is not completely nitrified to nitrate as is the case with Equation 2.1. In addition to this, not all the ammonia is nitrified, as a portion of it is consumed by the anammox organisms in Equation 2.6. When compared to ND processes, a PN/A process requires no additional COD source, can achieve a reduction in sludge production of approximately 90% and requires nominally 60% less aeration (Lackner *et al.*, 2014). With a PN/A process it is possible to achieve a 90% nitrogen removal efficiency of the incoming ammonia/nitrite with 10% of the incoming nitrogen leaving as nitrates in the sidestream effluent (Van Hulle *et al.*, 2010). In a survey of existing full scale PN/A processes conducted by Lackner *et al.* (2014), the range of energy efficiencies of PN/A processes was between 0.8 and 2 kWh/kg N. In addition to this, PN/A processes occur in much smaller reactors leading to reduced capital expenditure.

Initially PN/A processes were developed as a two-step process consisting of a SHARON reactor and an anammox reactor in series (van der Star *et al.*, 2007). Over the years they have developed into single step processes as the understanding of the biological processes have improved and with a drive to lower capital investment costs (Lackner *et al.*, 2014). In addition to this, single step processes have also shown significantly lower emissions of Nitrous Oxide (N_2O) when compared to two step processes (Guo *et al.*, 2019). N_2O is a greenhouse gas which has a global warming impact 265 times greater than carbon dioxide (Fenu *et al.*, 2019). Single step PN/A processes have been developed into a variety of reactor configurations:

- Sequencing Batch Reactor (SBR)
- Moving Bed Biofilm Reactor (MBBR)
- Integrated Fixed film activated sludge (IFAS)
- Granular Sludge

When looking at existing full scale applications, SBR configurations are most abundant (approximately 50% of installations) however granular sludge treat the highest amount of nitrogen worldwide with an average plant load of approximately 3 250 kg N/d (Lackner *et al.*, 2014). Some notable technology names include: ANITAmox™ (Veolia), ANAMMOX® (Paques), DEMON® (University of Innsbruck) and NAS® (Colsen) (Bungay, 2012; Lackner *et al.*, 2014). In a review done by Ali *et al.* (2015), by that time there were more than 114 anammox based technologies worldwide however they are almost exclusively located in the developed world i.e. Europe, USA and China. To date, there are no anammox based technologies in Africa. Figure 2.2 is taken from Ali *et al.* (2015) and depicts the geographical locations of the full scale anammox installations.

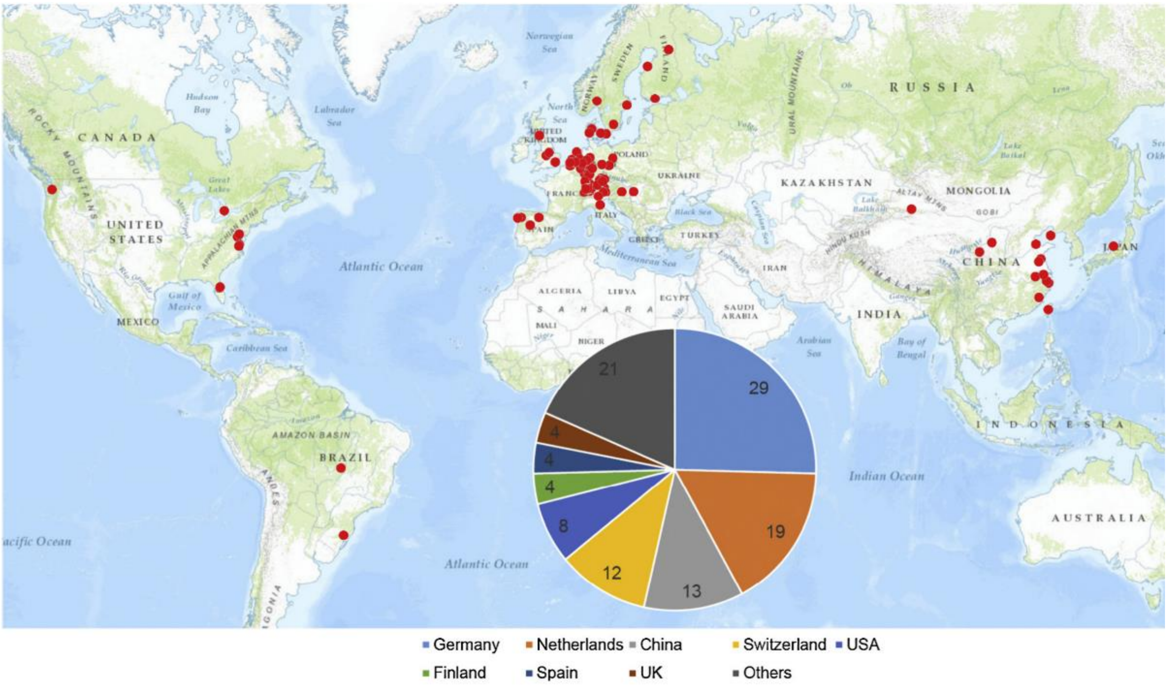

Figure 2.2: A map indicating the geographical distribution of full scale anammox plants around the world.

2.3 General Process risks and robustness

The commercial application of physical and chemical treatment technologies for the treatment of sidestream liquor from municipal digestate is either limited by cost or requires further research to be viable. For this reason, biological processes will be focused on. Further to this, the two extremes in terms of risk and robustness will be discussed namely, ND and PN/A processes.

From section 2.2.2 one can see that there are a variety of biological process technologies that can be applied to treat sidestream liquor. However, each technology has its advantages and disadvantages. No technology can be said to be the best for all applications. It is therefore crucial to consider each design on a case by case basis (Bungay, 2012; Lackner *et al.*, 2014). Important aspects to consider include: The nitrogen load to be treated, variability of sidestream liquor quality, space available at the site, skills available in the area, the available capital and the operational budget are but to name a few.

2.3.1 ND processes

Conventional and high-rate ND processes use common activated sludge bacteria (OHO and ANO) to achieve nitrogen removal of which critical factors such as kinetics, growth yields and inhibitory factors are well known and understood both at lab and full-scale applications. These processes can handle large variations in flow and concentrations of COD, ammonia, phosphorus and suspended solids (Bungay, 2012). ND processes operate on the same philosophy as mainstream wastewater ND processes and can therefore be run and maintained by operators with less operational education and training.

2.3.2 PN/A processes

With the current knowledge of the biological processes and the current commercially applied reactor designs, PN/A processes have a higher degree of risk than ND processes. The successful application of PN/A processes require the careful control of multiple biological processes occurring simultaneously. This results in a high degree of automation (PLC controllers and online sensors) to ensure that all process parameters are tightly controlled (Lackner *et al.*, 2014). Following on from this, it is logical to assume that operators and technicians need to have a higher process operation education and experience level to successfully operate and maintain this process.

There are various conditions that are known to be inhibitory to the biological processes occurring in PN/A reactors. Nitrogen species (ammonia, nitrite and nitrate) need to be carefully monitored and controlled. High ammonium concentrations leading to high free ammonia concentrations (>200 mg/l) are known to be inhibitory for the oxidation of ammonium through AOB's (Lackner *et al.*, 2014). Similarly, high concentrations of nitrite can also inhibit the process. This is particularly challenging during start-up when AOB growth rates are faster than anammox bacteria which can lead to an accumulation of nitrite in the system. Nitrate accumulation is less significant from an inhibitory perspective, however, the concentration still needs to be monitored as this is an indication of how well the partial nitritation process is controlled i.e. the selective growth of AOB's over NOB's. In this regard, the control of DO is critical as over aeration (leading to high DO) leads to the accumulation of NOB's. High DO concentrations are also known to inhibit anammox bacteria (Lackner *et al.*, 2014).

On a physical level, high suspended solids in the feed to a PN/A process results in an increased inert fraction which leads to lower biomass activity (Lackner *et al.*, 2014). This is a particular concern when treating sidestream liquor produced from THP-AD processes that are known to produce colloidal solids which are difficult to separate out in dewatering processes. Colloidal solids have been found to inhibit AOB activity (Zhang *et al.*, 2018). THP-AD processes are also known to produce a higher concentration of organics in the sidestream liquor which causes inhibition in PN/A processes (Gu *et al.*, 2018; Zhang *et al.*, 2018).

Anammox bacteria have a very low growth rate resulting in the slow start-up of commercial PN/A processes which is a risk to its application (Guo *et al.*, 2019). The doubling time of anammox bacteria has been reported to be between 10-12 days (van der Star *et al.*, 2007; Eskicioglu, Galvagno and Cimon, 2018) Since the first commercial reactor start-up in Rotterdam, Netherlands (van der Star *et al.*, 2007), significant improvements in start-up times have occurred especially with the availability of anammox seed sludge (Lackner *et al.*, 2014). Wang *et al.* (2018) showed that increasing the amount of anammox sludge seed material from 0.02 g VSS/l to 0.2 g VSS/l reduced the start-up time from 21 to 5 days thereby confirming that a rapid start-up is possible with access to anammox seed sludge. However, the location of the site and therefore the availability of seed sludge must be considered. There is also a large demand for anammox seed sludge with a limited supply globally. Considering that there are no plants located in Africa (Ali and Okabe, 2015), this makes the risk of obtaining good quality seed material even higher. However, Wang et al. (2018) have shown that with the correct seeding strategy it is possible to reduce the start-up time of anammox processes even with relatively low concentrations of anammox seed material.

2.4 Process Modelling

2.4.1 Dynamic and Steady State Models

Process modelling can be broadly grouped into dynamic and steady state modelling. Dynamic modelling has a long and rich history within the wastewater field with the development of various activated sludge models to cater for carbon and nutrient removal (Gernaey *et al.*, 2004). Dynamic models are beneficial to evaluate complex transient conditions experienced at wastewater works but generally run on sophisticated software packages involving complex mathematical solvers which can make them impractical as a simplified design tool for decision making. Based on the same biological processes used by dynamic models, various explicit steady state equations have been developed (Ekama, 2011b) which allow for the creation of steady state models for various activated sludge configurations on spreadsheet platforms (Wu, 2015). This allows for a simplified development of different plant design scenarios that can be evaluated for decision making purposes.

Table 2.3 below provides a summary of steady state explicit algebraic equations that would be applicable for the design of an MLE activated sludge process (Ekama, 2011b). The equations account for the growth and death of the organisms, the accumulation of unbiodegradable organic and inert particulates, the oxygen required and the predicted effluent concentrations.

Table 2.3: Activated sludge steady state model design equations.

Equation	**Equation Number**
$FS_{ti} = Q_i S_{ti}$	(2.7a)
$FS_{bi} = FS_{ti}(1 - f_{S'us} - f_{S'up})$	(2.7b)
$FX_{Ii} = \frac{FS_{ti} f_{S'up}}{f_{cv}}$	(2.7c)
$FX_{IOi} = Q_i X_{Ioi}$	(2.7d)
$MX_{BH} = FS_{bi} \frac{Y_H R_s}{(1 + b_H R_s)}$	(2.7e)
$MX_{EH} = f_H b_H MX_{BH} R_s$	(2.7f)
$MX_I = R_s \frac{FX_{Ii}}{f_{cv}}$	(2.7g)
$MX_{VH} = MX_{BH} + MX_{EH} + MX_I$	(2.7h)
$MX_{IO} = FX_{IOi} R_s + f_{iOHO} MX_{BH}$	(2.7i)
$MX_t = MX_{VH} + MX_{IO}$	(2.7j)
$FO_c = FS_{bi}[(1 - f_{cv} Y_H) + (1 - f_H) b_H \frac{Y_H f_{cv} R_s}{(1 + b_H R_s)}$	(2.7k)
$FX_t = \frac{MX_t}{R_s}$	(2.7l)

$S_{te} = S_{use} + f_{cv}X_v$	(2.7m)
$N_a = N_{ae} = \frac{K_{nT}(b_{AT} + \frac{1}{R_s})}{\mu_{AmT} - (b_{AT} + \frac{1}{R_s})}$	(2.8a)
$N_{ae} = \frac{K_{nT}(b_{AT} + \frac{1}{R_s})}{\mu_{AmT}(1 - f_{xt}) - (b_{AT} + \frac{1}{R_s})}$	(2.8b)
$N_{ae} = \frac{K_{nT}}{(S_f - 1)}$	(2.8c)
$N_{ae} = N_{ai} + N_{obsi} + N_{obpi} - (N_s - N_{oupi})$	(2.8d)
$N_{te} = N_{ouse} + N_{ae}$	(2.8e)
$N_s = f_n \frac{MX_{VH}}{Q_i R_s}$	(2.8f)
$N_{te} = N_{ti} - N_s$	(2.8g)
$N_{ne} = N_c = N_{ti} - N_{te} - N_s$	(2.8h)
$D_{p1RBCOD} = \frac{f_{Sb's}S_{bi}(1 - f_{cv}Y_H)}{2.86}$	(2.8i)
$D_{p1SBCOD} = \frac{K_2 f_{x1} S_{bi} Y_H R_s}{(1 + b_h R_s)}$	(2.8j)
$D_{p1} = D_{p1RBCOD} + D_{p1SBCOD}$	(2.8k)
$D_{p3SBCOD} = \frac{K_3 f_{x3} S_{bi} Y_H R_s}{(1 + b_h R_s)}$	(2.8l)
$D_{p3} = 0 + D_{p3SBCOD}$	(2.8m)
$FO_n = 4.57 N_c Q_i$	(2.8n)
$FO_d = 2.86(N_c - N_{ne})Q_i$	(2.8o)
$FO_t = FO_c + FO_n - FO_d$	(2.8p)

PN/A processes have been modelled successfully as granular biofilm processes with and without the additional influence of heterotrophic biomass growth due to the presence of biodegradable COD (Volcke *et al.*, 2010; Mozumder *et al.*, 2014). Due to the complex nature of mass transfer limitations within the various biofilm layers and its effect on rate determining species concentrations (ammonia, oxygen, nitrate/nitrite), explicit steady state equations such as the ones in Table 2.3 have not been derived for PN/A processes. Instead, the nitrogen conversion concentrations, as well as the predicted effluent concentrations were calculated using the mass-balanced stoichiometry of Equations 2.5 and 2.6.

2.4.2 Evaluating Process Models

Process modelling is a valuable tool that can be used to evaluate various proposed treatment solutions. Various performances indices such as an effluent cost index (EQI) and operational cost index (OCI) have also been developed to allow the results from process modelling to be objectively evaluated (De Ketele, Davister and Ikumi, 2018; Ikumi *et al.*, 2019). These indices can be adapted taking into account critical variables that are applicable to a specific process or set of processes being compared.

2.5 South African Challenges

2.5.1 The Status of Wastewater Treatment in South Africa

In a study by Ikumi et al. (2019) it was noted that most treatment plants operating in South Africa do not consider the sidestream liquor being recycled back to the mainstream works for further treatment. This results in existing infrastructure being overloaded and application of costly chemical dosing solutions (metal salts i.e. ferric chloride) to bring wastewater back into specification.

South Africa uses the Green Drop Auditing program to assess the status of its wastewater treatment plants which was started in 2009 (Ntombela *et al.*, 2016). A critical review conducted by Ntombela *et al.* (2016) indicated that the performance of mainstream wastewater treatment plants in South Africa is concerning with only 50.4% of plants achieving ≥ 50% in their respective audits in 2012/13, with the national average Green Drop score of only 46.4%. More recently, the South African department of Water and Sanitation released the National Water and Sanitation Master Plan. This document highlights the status quo of the wastewater industry whereby approximately 56% of WWTP's in South Africa are in poor or critical condition and require rehabilitation (South African Department of Water and Sanitation, 2018). When one looks at the provincial level however, the Western Cape (wherein Cape Flats WWTP is situated) achieved a provincial average score of 84.5% which is vastly higher than the national average. Similar provincial scores for Gauteng (82.7%) and KwaZulu Natal (81.5%) were achieved (Ntombela *et al.*, 2016). These three provinces represent the biggest metropolitan areas which indicates that poor wastewater treatment performance is skewed towards the more rural areas within South Africa. Other challenges highlighted by the Green Drop Audits include (Ntombela *et al.*, 2016):

- Lack of financial capacity to prioritize wastewater treatment solutions
- Lack of human resource capacity, particularly in rural areas, to implement wastewater treatment solutions/corrective actions

- Inefficient bureaucratic processes particularly when it comes to handling maintenance of plants and tendering procedures as stipulated by the Municipal Finance Management Act (Act No. 5 of 2003)
- Theft and Vandalism of WWTP infrastructure.

(Ntombela *et al.*, 2016)

2.5.2 Energy Generation and Cost

Approximately 90% of the energy in South Africa is supplied by the State Owned Entity (SOE), Eskom whereby 83% is derived from coal power stations (Ratshomo and Nembahe, 2018). The cost of electricity has outstripped inflation since the electricity crisis in 2008 as indicated by Moolman (2019) in Figure 2.3 .

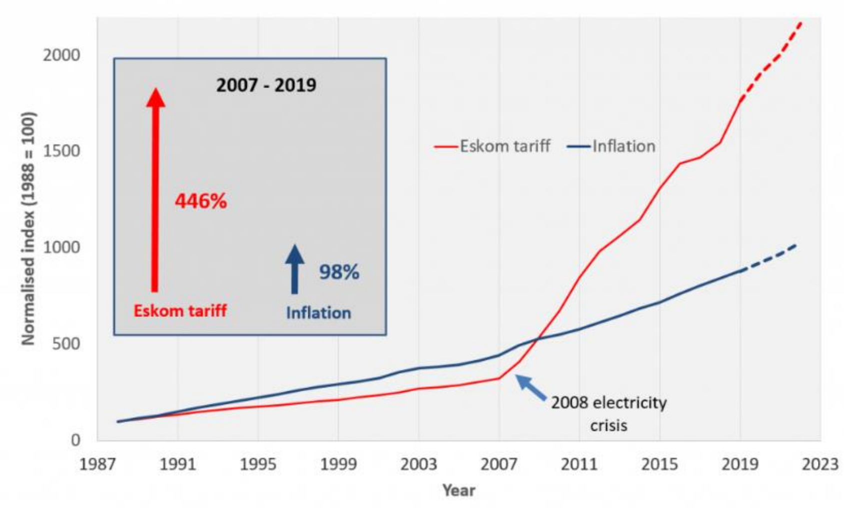

Figure 2.3: Eskom average tariffs versus inflation (CPI) in South Africa.

The electricity crisis in 2008 was a result of a lack of investment into additional generation capacity. Since then a sharp increase in tariffs has been implemented to offset the construction of new generation capacity and for maintenance of existing power plants that are reaching end of life. In addition to this, political interference, mismanagement and corruption has plagued Eskom over the same period which has resulted in Eskom accruing large sums of debt. Due to South Africa’s reliance on Eskom, the utility will not be allowed to fail and therefore above inflation tariffs can be expected into the future as indicated by Moolman (2019) .

2.5.3 The South African Economy

The South African economy has been struggling for many years with real GDP growth being well below 1% since 2018. The low economic growth has resulted in a consistent fiscal deficit and government debt has increased substantially and continues to grow (African Development Bank Group, 2020). The onset of COVID-19 pandemic in 2020 has resulted in a contraction in the economy and has caused government debt to GDP ratio to balloon. An IMF bailout package has also been initiated to assist with funding the current deficit. On the back of this, the government plans to implement structural reforms to limit the peak government debt to GDP ratio at 87% by 2023/24. These structural reforms will come in the form of budget cuts (zero based budgeting exercises) across all sectors of government spending (International Monetary Fund, 2020).

2.5.4 Contextualization of South African challenges

This section has highlighted specific challenges unique to South Africa. Although a PN/A process cannot be seen as completely novel due to it being applied to more than 100 installations worldwide (Lackner *et al.*, 2014), it is regionally novel to South Africa and therefore the appropriateness of the technology needs to be aligned to the surrounding operating conditions.

As communicated by Lackner *et al.* (2014)*,* the application of sidestream treatment technologies needs to be assessed on a case by case basis considering site-specific conditions. Although site specific conditions such as process loads, available utilities, available space, skills and expertise etc. must be considered, assessing broader issues such as macroeconomic factors and energy supply within the operating region are equally important in assessing appropriate technologies.

In this section the South African wastewater industry was reviewed which highlighted a lack of financial and labour capacity and inefficient bureaucratic processes. Further to that, the status of the energy sector was reviewed indicating substantial increases in energy costs with energy generation emanating largely from coal fired power stations. Lastly, the poor state of the South African economy, plagued by low growth and ballooning debt, was discussed. The results of this research are discussed in Section 4.6 with reference to the issues highlighted in this section to fully contextualize the appropriateness of the sidestream technologies operating within a South African environment.

2.6 Summary of Literature Review

From the literature review the following points are emphasized.

Sidestream liquor and treatment technologies:

- Sidestream liquor has a low volumetric load but contains a higher concentration of nutrients released during AD. The application of AAD such as THP results in significantly higher nutrient concentration and loads (approximately double) compared to conventional AD. If this is recycled back to the mainstream works, effluent quality can be compromised.
- Technologies to treat sidestream liquor include physical/chemical technologies and

biological processes. For large scale municipal applications, biological processes are favoured over chemical/physical due to costs.

- Biological processes include Nitrification/Denitrification (ND), nitritation/denitritation and Partial Nitritation Anammox (PN/A) with PN/A being the most novel as well as reporting the highest efficiencies in terms of aeration requirements, COD usage and sludge generation.
- Commercial ND sidestream treatment processes, such as AMTreat, operate as MLE configurations. There are a multitude if commercial PN/A processes such as ANITAmox™ (Veolia), ANAMMOX® (Paques), DEMON® (University of Innsbruck) and NAS® (Colsen) and are offered in various reactor configurations such as Sequencing Batch Reactor (SBR), Moving Bed Biofilm Reactor (MBBR), Integrated Fixed Film Activated Sludge (IFAS) and Granular Sludge
- MLE processes are more robust and generally easier to operate while PN/A processes have technical challenges such as the slow growth of autotrophic biomass, obtaining seed sludge and the control of inhibitory factors effecting the balance of the nitritation and anammox reactions.

Process modelling:

- Process modelling allows for various scenarios and designs to be assessed on a desktop basis and represent a crucial tool for decision making.
- Process modelling can broadly be split into dynamic modelling and steady-state modelling. Dynamic modelling is done on sophisticated modelling software while steady state modelling adapts dynamic mathematical equations into simplified steady state equations which can be applied to user-friendly spreadsheets.
- Standardized performance indices such as the OCI and EQI can be adapted to evaluate the results of process models.

South African contextualization:

- Sidestream treatment is not applied at municipal AD sites across South Africa creating inefficient/overloaded mainstream processes.
- The South African economy is struggling with low growth and large government debt creating increased pressure on municipal budgets impacting the ability to spend on infrastructure.
- Energy generation in South Africa is predominately from coal. Not only does this have an associated environmental impact, but the cost of energy has increased above inflation for more than a decade which is likely to remain for the foreseeable future.
- South Africa has specific challenges which need to be carefully considered when assessing sidestream treatment technologies which may not be relevant to other countries.

Considering the topics and information covered in the literature review the following research gaps have been identified:

- A comparative study on the application of sidestream treatment from a municipal AAD on a commercial scale, operating within a South African environment, has not been done. This type of study can not only be useful to select the appropriate technology for Cape Flats WWTP, but the methodology could be used to apply to similar studies at other sites around South Africa which can be used to promote the application of further treatment technologies.
- A steady-state model for a PN/A process has not been developed. The benefit of developing such a model, similar to other steady state models, is that it could allow for a user-friendly assessment of the PN/A process to obtain plant design details for costing which is valuable for decision making. This could also be used as a basis to compare steady-state and existing dynamic PN/A models to identify significant differences and justify any limitations that may exist with the use of steady-state models.

3 RESEARCH METHODOLOGY

This chapter describes the methods and procedures followed to achieve the research objectives. This is broadly outlined in Figure 3.1, which indicates a simplified research framework for this study.

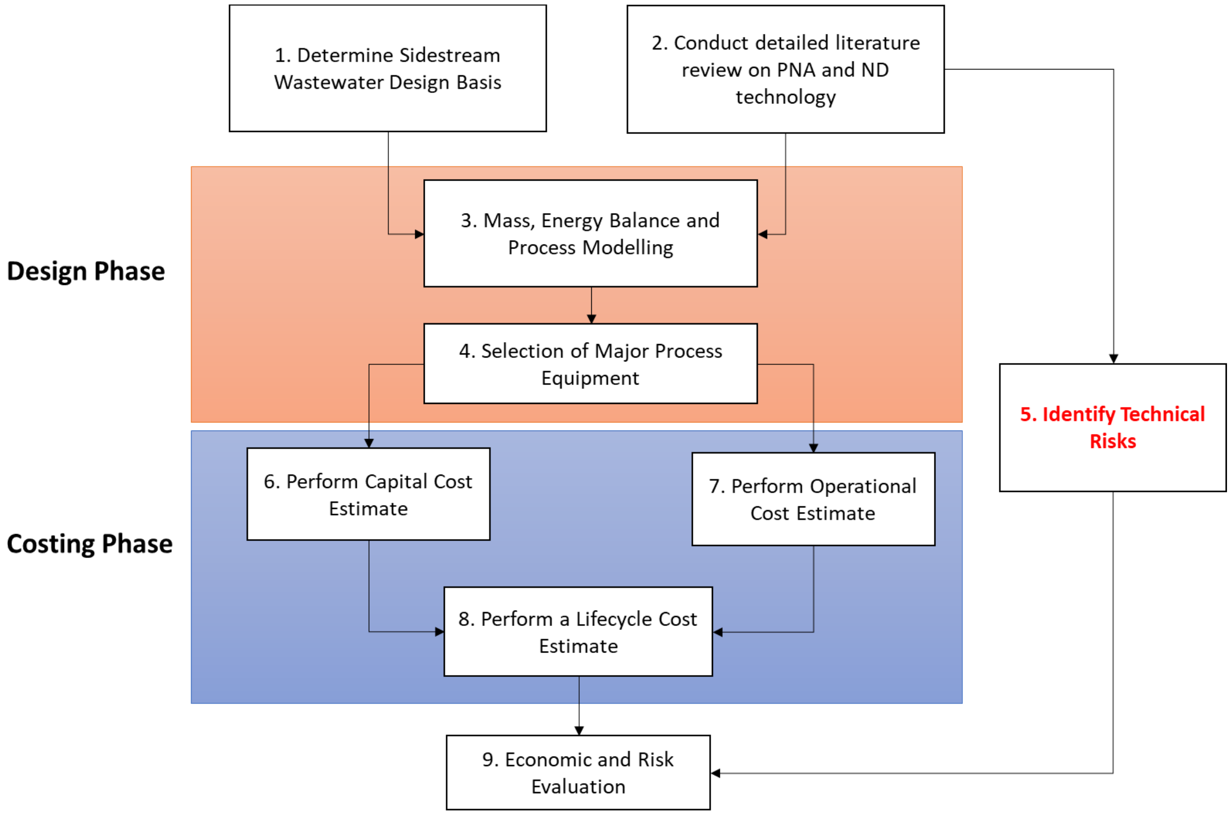

Figure 3.1: The research framework for the comparison of a ND (MLE) process and a novel PN/A process for the treatment of sidestream wastewater.

The first step was to determine the design basis for the sidestream liquor as shown in Section 3.1. Following on from this, two steady state process models describing the PN/A and MLE process were developed in an Excel spreadsheet. Section 3.2 discusses the model development in detail. The outputs of the model were used to size the major equipment required for the side-stream treatment processes. Section 3.4 discusses the costing for all the major pieces of equipment to determine the total capital cost for implementing each of the processes. All costs were taken from industry to ensure that the capital cost estimations are relevant for South Africa. Key operational costs such as energy and chemical consumption were then calculated using South African costs from major chemical suppliers, as well as the national energy provider. Once all capital and operational costs were defined, a life-cycle cost analysis was performed to determine the best economic solution as shown in Sections 3.5 and 3.6.

3.1 Sidestream wastewater design basis

The source of the sidestream liquor is centrate from an advanced anaerobic digestion (AAD) process that has not yet been built therefore sampling and analysis of the liquor directly was not possible. To gain a realistic design basis which served as inputs for the modelling the PN/A and MLE processes, a theoretical mass balance was conducted to characterize the sidestream liquor. This section discusses the methodology that was followed.

3.1.1 FSA, OP, TSS and VSS mass balance

A theoretical mass balance on free and saline ammonia (FSA), orthophosphate (OP), total suspended solids (TSS) and volatile suspended solids (VSS) was conducted over the AAD process plant using the future sludge throughput and PS/WAS split as inputs. Figure 3.2 provides a simplified process flow diagram indicating the operations involved in the regional sludge digestion facility, major assumptions and the boundary limits of the mass balance together with the stream numbers to be referenced against the results in Section 4.1, Table 4.3. The mass balance was conducted on the nutrient, VSS and TSS components within the boundary limits.

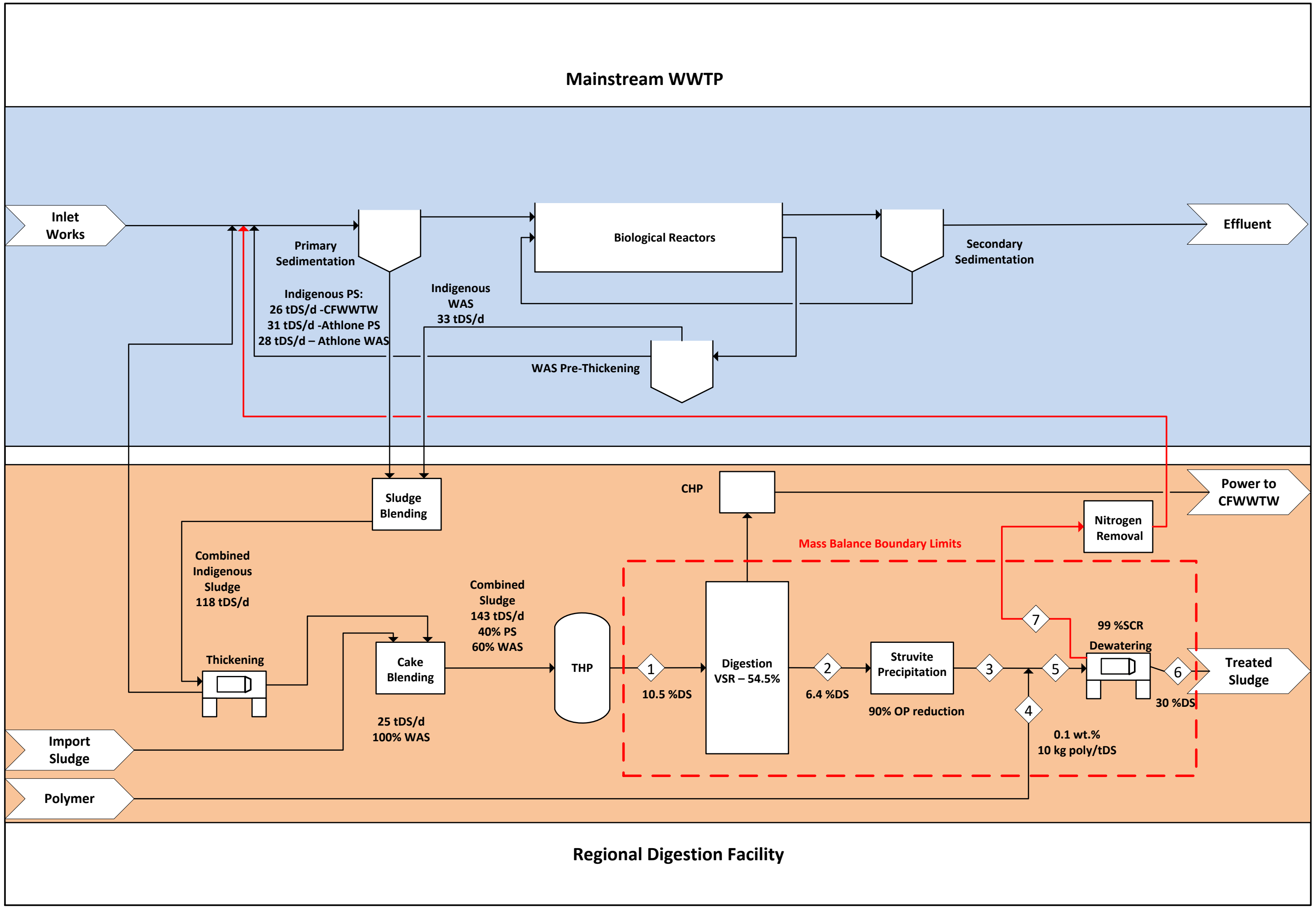

Figure 3.2: A process flow diagram indicating the major unit operations of the mainstream WWTP and the regional digestion facility

The total sludge throughput is made up of imported and indigenous sludge. Indigenous sludge is defined as the amount of sludge produced at Cape Flats WWTP from the PST's and bioreactors whereas imported sludge is transported to site from surrounding WWTPs via sludge collection trucks. The total sludge throughput was taken from the design basis for the regional digestion facility – 143 dry tons per day ($FX_{t,1}$) with a PS/WAS split of f_{PS} = 0.4 and f_{WAS} = 0.6.

The fraction of VSS in the TSS in the WAS and PS is indicated in Table 3.1 below:

Table 3.1: The fraction of VSS in TSS of WAS and PS.

	Value	Source
Fraction VSS/TSS in WAS, $f_{v,WAS}$	0.683	(Wentzel and Ekama, 1997; Ekama, 2011b)
Fraction VSS/TSS in PS, $f_{v,PS}$	0.81	(Ikumi, 2011)

The amount of ISS present in the WAS calculated assuming a fraction of 0.17 for a process running no BEPR with a sludge age of 15 days with settled wastewater (Wentzel and Ekama, 1997). The ISS fraction from the PAO's was then calculated from Equation 3.1 and added to the ISS present in the WAS without EBPR to obtain $f_{V,WAS}$ indicated in Table 3.1. The fractions of phosphorus in PAOs and WAS is indicated in Table 3.2.

$$f_{i,PAO} = 3.286(f_{p,PAO} - f_{p,OHO}) + 15 \qquad (3.1)$$

The VSS in the feed to the digester was calculated as follows:

$$FX_{V,WAS,1} = FX_{T1} \cdot f_{WAS} \cdot f_{V,WAS,1} \qquad (3.1a)$$

$$FX_{V,PS,1} = FX_{T1} \cdot f_{PS} \cdot f_{V,PS,1} \cdot f_{VSR} \qquad (3.1b)$$

$$FX_{V,1} = FX_{V,PS,1} + F \cdot X_{V,WAS,1} \qquad (3.1c)$$

A Volatile Solids Reduction (VSR, f_{VSR}) of 0.545 was used (Higgins *et al.*, 2017). The total mass flow rate (kg/d) of VSS in the digester effluent was calculated from:

$$FX_{V,WAS,2} = FX_{V,WAS,1} \cdot (1 - f_{VSR}) \qquad (3.1d)$$

$$FX_{V,PS,2} = FX_{V,PS,1} \cdot (1 - f_{VSR}) \qquad (3.1e)$$

$$FX_{V,2} = FX_{V,PS,2} + FX_{V,WAS,2} \tag{3.1f}$$

The amount of ISS in the AD influent (t/d) is carried through the AD and downstream processes and was calculated from:

$$FX_{IO,1} = FX_{T,1} - F \cdot X_{V,1} \tag{3.1g}$$

The total solids mass flux in the AD effluent (t/d) was then calculated from:

$$FX_{T,2} = FX_{V,2} + FX_{IO,1} \tag{3.1h}$$

The percentage solids ($X_{T,2}$) in the AD effluent was assumed to be 6.4% (Higgins *et al.*, 2017). The total volumetric flowrate in the digester effluent was then calculated. The contribution of MgO solution added to the struvite process was assumed negligible in terms of its effect on the overall volumetric flow rate as this represented less than 1% of the total volumetric flow.

$$Q_2 = \frac{FX_{T,2}}{X_{T,2}} = Q_3 \tag{3.1i}$$

The mass flux of inorganics released into the soluble phase as ammonia and orthophosphate were calculated by taking the PS and WAS mass flowrates into the AD and multiplying it by the percentage volatile solids reduction expected in the digester (with THP) as per the following equations:

$$FN_{a,2} = \left(FX_{V,WAS,1} - FX_{V,WAS,2}\right) \cdot f_{n,WAS} + (FX_{V,PS,1} - FX_{V,PS,2}) \cdot f_{n,PS} + \tag{3.2}$$

$$FP_{s,2} = \left(FX_{V,WAS,1} - FX_{V,WAS,2}\right)[f_{P,OHO} \cdot \{1 - f_{PAO}\} + f_{P,PAO} \cdot f_{PAO}] + f_{PS} \cdot f_{P,PS}) \tag{3.2a}$$

The mass flux of ammonia and orthophosphate in the digester influent was ignored from the calculation as this was assumed negligible due to the typical low concentration of these nutrients in the respective sludges.

The fractions indicated in Table 3.2 were used in equation 3.2 & 3.2a.

Table 3.2: Nutrient mass fractions of WAS and PS used as inputs into the nutrient mass balance over the AD process.

	WAS	PS	Source
Fraction of nitrogen in VSS (mgN/mgVSS), f_n	0.1[A]	0.04[B]	A (Ekama, 2011b) B (Sötemann, 2005; Ikumi, 2011)

Fraction of phosphorus in OHO VSS (mgP/mgVSS), $f_{p,OHO}$	0.03[A]	0.03[B]	A (Ekama, 2011b) B (Sötemann, 2005; Ikumi, 2011)
Fraction of phosphorus in PAO active VSS (mgP/mgVSS), $f_{p,PAO}$	0.12[C]	N/A	(Ekama, 2011b)(Wentzel and Ekama, 1997)
Fraction of PAO active mass is WAS, f_{PAO}	0.33	N/A	(Wentzel and Ekama, 1997)

[B] - The nitrogen and phosphorus fractions were calculated assuming a UPO fraction in the PS (f_{sup}) of 0.36 (Sötemann, 2005) and an elemental composition of $C_{1.0}H_{1.32}O_{0.44}N_{0.1}P_{0.04}$ and $C_{1.0}H_{2.19}O_{0.65}N_{0.06}P_{0.01}$ for the UPO and BPO portions of PS respectively (Ikumi, 2011).

[C] - Ekama (2011b) gives a range of 0.06-0.15 mgP/mgVSS in the activated sludge for BEPR processes. A $f_{p,PAO}$ of 0.12 mgP/mgVSS was selected to give an overall phosphorus content in the WAS of 0.06 (lower range) assuming a PAO content in the WAS of 33% (Wentzel and Ekama, 1997). The lowest range was selected as this gives the worst-case load of nitrogen to the sidestream treatment plant as less ammonia is removed in the struvite process.

The FSA and OP concentrations were calculated by dividing the mass flux of each of the nutrients in the soluble phase by the total volumetric flowrate (Q) of the stream.

$$N_{a,2} = \frac{FN_{a,2}}{Q_2} \tag{3.3}$$

$$P_{s,2} = \frac{FP_{s,2}}{Q_2} \tag{3.4}$$

In the mass balance it was assumed that a struvite precipitation process will be installed on the digester effluent stream which removes 90% of the soluble phosphorus (f_{PRE}) by dosing a magnesium source – Magnesium Oxide (Zhou *et al.*, 2015). The mass flow of struvite (FX_{STR}) was calculated stoichiometrically based on Equation 3.5. and added to the ISS carried through from the AD influent. In addition to removing phosphorus, the struvite process removes nitrogen as per the stoichiometric equation. This was accounted for using the following equations:

$$Mg^{2+} + NH_4{}^{+} + HPO_4{}^{2-} + 6H_2O \rightarrow MgNH_4PO_4 \cdot 6H_2O + H^{+} \tag{3.5}$$

$$FP_{s,3} = FP_{S,2} \cdot (1 - f_{PRE}) \tag{3.5a}$$

$$FN_{a,3} = FN_{a,2} - FP_{S,2} \cdot f_{PRE} \cdot (\frac{14}{31}) \tag{3.5b}$$

$$FX_{STR} = FP_{S,2} \cdot f_{PRE} \cdot \frac{244}{31} \tag{3.5c}$$

$$P_{s,3} = \frac{FP_{s,3}}{Q_3} \tag{3.5d}$$

$$N_{a,3} = \frac{FN_{a,3}}{Q_3} \tag{3.5e}$$

$$FX_{IO,3} = FX_{IO,1} + FX_{STR} \tag{3.5f}$$

$$FX_{T,5} = FX_{T,3} + FX_{IO,3} \tag{3.5g}$$

$$X_{T,3} = \frac{FX_{T,3}}{Q_3} \tag{3.5h}$$

After struvite precipitation, polyelectrolyte solution is added to the sludge to condition the sludge before dewatering. The volumetric flowrate of polyelectrolyte solution was calculated based on a dry polymer powder dosing rate of 10 kg poly/TDS with a polymer solution concentration of 0.1 wt.% to achieve an average cake solids ($X_{T,6}$) of 30% and a solids capture rate (f_{SCR}) of 0.99 (Higgins *et al.*, 2017).

$$Q_4 = \frac{FX_{T,3} \cdot 10\ [\frac{kgpoly}{tDS}]}{0.1\ [wt.\%] \cdot 10000} \tag{3.5i}$$

The volumetric flowrate of the dewatered cake was calculated from:

$$Q_6 = \frac{Q_5 \cdot X_{T,5} \cdot f_{SCR}}{X_{T,6}} \tag{3.5j}$$

The total volume of the sidestream liquor generated was then calculated from:

$$Q_7 = Q_3 + Q_4 - Q_6 \tag{3.5k}$$

The concentration of ammonia, phosphorus, TSS and ISS in the sidestream liquor was then determined from:

$$N_{a,7} = \frac{FN_{a,3}}{Q_3 + Q_4} \tag{3.5l}$$

$$P_{s,7} = \frac{FP_{s,3}}{Q_3 + Q_4} \tag{3.5m}$$

$$X_{T,7} = \frac{Q_5 \cdot X_{T,5} \cdot f_{SCR} - Q_6 \cdot X_{T,6}}{Q_7} \tag{3.5n}$$

$$X_{V,7} = (1 - \frac{Q_3 \cdot X_{IO,3}}{Q_3 \cdot X_{T,3}}) \cdot \tag{3.5o}$$

Finally, the mass flux of nutrients in the sidestream liquor was calculated from:

$$FN_{a,7} = Q_7 \cdot N_{a,7} \tag{3.5p}$$

$$FP_{s,7} = Q_7 \cdot P_{s,7} \tag{3.5q}$$

3.1.2 COD, TKN and TP determination

To model the sidestream liquor processes, the sidestream liquor needed to be fully fractionated. The following method was used to fractionate the sidestream liquor.

The total COD in the sidestream liquor was assumed to be 4500 mg/l (largely due to the significant presence of VFAs) which is in line with values indicated in literature as presented in Table 2.1 and Table 2.2 (Zhang *et al.*, 2016; Gu *et al.*, 2018). With this assumption and the results of the mass balance described in 3.1.1, the COD and organic nitrogen and phosphorus was fractionated. More details on how the fractionation was accomplished is described in Section 3.2.1 and 3.2.2.

3.2 Process modelling, mass and energy balances

All process modelling was conducted using the Plant-Wide Steady State Mass Balance Model (PWSSMBM) using first principle steady state equations derived for biological nutrient removal (Ekama, 2011b). The two scenarios assessed were:

1. Sidestream treatment – PN/A process
2. Sidestream treatment – MLE process

It is crucial to note that the purpose of the modelling exercise was to obtain key design parameters of the sidestream treatment solutions to allow for the comparative economical and technical assessment to be conducted. Certain assumptions were made for simplification purposes which were deemed to be within accuracy required for this type of analysis. For clarity, all critical assumptions and limitations are highlighted in Section 3.2.1 and 3.2.2. Simplifying assumptions allow for the models to be used and understood by an audience with various skillsets not only limited to a dynamic process modelling background. This allows for the models to be an important decision-making tool in the design of the proposed sidestream processes.

3.2.1 Sidestream treatment – PN/A model

In this model the sidestream liquor was treated using a single stage, granular sludge PN/A process. A continuous single stage process was selected over the two-step process as the single stage process is less capital intensive (Lackner *et al.*, 2014) and practical experience has shown this design to have lower N_2O emissions (Guo *et al.*, 2019). Out of existing commercially operated plants, granular sludge also treats the most nitrogen worldwide (Lackner *et al.*, 2014). A simplified process flow diagram is shown in Figure 3.3 below.

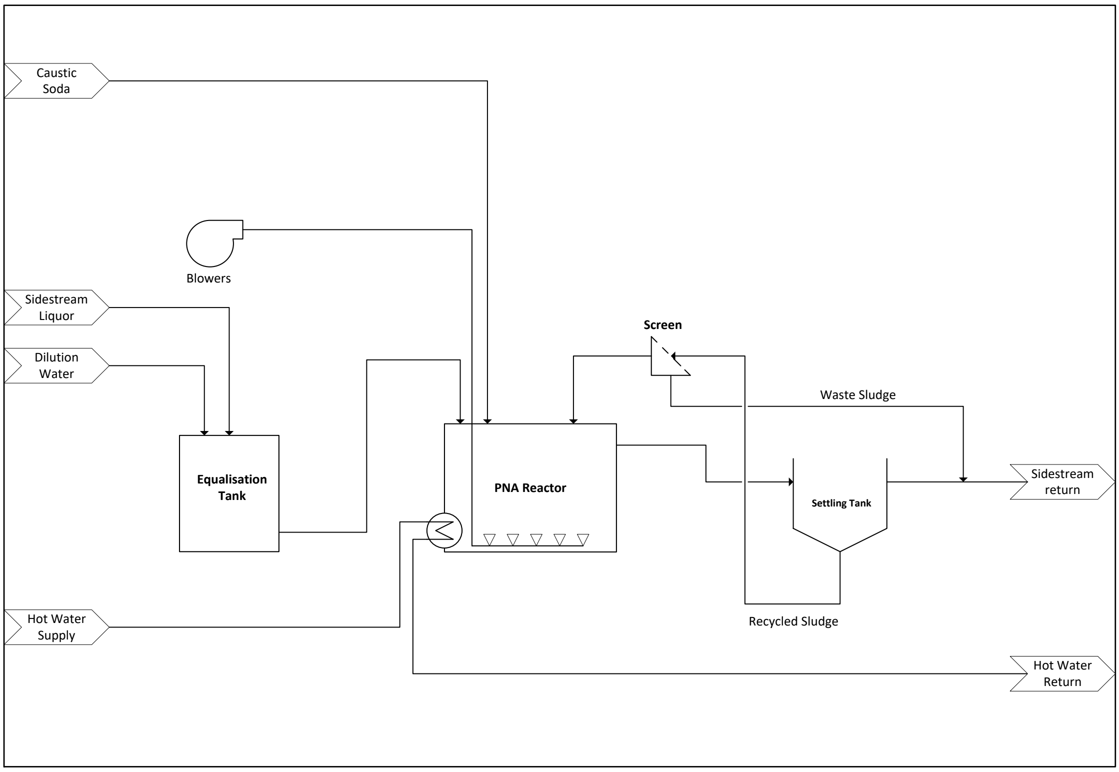

Figure 3.3: A Simplified process flow diagram indicating the PN/A process.

The steady-state model was constructed using the relevant equations and stoichiometry (Ekama, 2011b; Eskicioglu, Galvagno and Cimon, 2018), coded into a spreadsheet. The model boundary starts at the discharge of the dewatering unit operation and ends at the effluent discharge from the

PN/A settling tank. The model input variables are the characteristics of the effluent AAD process. A list of all inputs into the model and associated assumptions is listed in Section 8.1 in the appendix.

3.2.1.1 Influent fractionation

Using the assumed inputs, the various components of COD, nitrogen, phosphorus, and inorganic solids in the influent were fractionated using the following method.

From the influent TSS and VSS/TSS ratio ($\frac{Xv}{Xt}$) inputs, the total particulate COD was determined by:

$$S_p = X_t \cdot \frac{Xv}{Xt} \cdot f_{cv} \tag{3.6}$$

The unbiodegradable and biodegradable particulate COD was determined by assuming an unbiodegradable fraction of the total particulate COD ($f_{s,up,part}$) of 0.7.

$$S_{up} = S_{tp} \cdot f_{s,up,part} \tag{3.7}$$

and

$$S_{bp} = S_{tp} - S_{up} \tag{3.8}$$

The soluble COD was then determined using the total COD as an input:

$$S_{ts} = S_{ti} - S_{tp} \tag{3.9}$$

The unbiodegradable soluble COD was determined by assuming a fraction of the soluble COD that was biodegradable ($f_{s,us,sol}$):

$$S_{us} = S_{ts} \cdot f_{s,us,sol} \tag{3.10}$$

The biodegradable soluble concentration was then calculated from:

$$S_{bs} = S_{ts} - S_{us} \tag{3.11}$$

The model assumes that all S_{bs} is in the form of VFA’s considering the digestion process upstream would have hydrolyzed all the unconverted S_{bs}, therefore organic nitrogen and phosphorus associated with BSO (N_{obs} & P_{obs}) was assumed to be zero. In addition to this, THP creates a substantial amount for unbiodegradable soluble COD with a large fraction of organic nitrogen (N_{ous}).

This is dependent on the operating temperature of the THP (Higgins *et al.*, 2017). To account for this, N_{ous} was assumed to be 300 mgN/l as determined in the study conducted by Higgens *et al.* (2017) where the concentration of N_{ous} was measured in the sidestream liquor of a THP process with a blend of PS and WAS. It should be noted that the N_{ous} assumption is important for the influent and effluent characterization but will not influence either of the MLE or PN/A processes (and hence the comparative analysis) as the material is unbiodegradable.

Once all the COD fractions of the influent are determined, the respective organic groups (USO, UPO, BSO and BPO) of N_{org} and P_{org} are calculated using the f_{cv} and the respective mass ratio f_x:

$$N_{org,x} = \frac{S_x}{f_{cv}} \cdot f_{n,x} \tag{3.12}$$

$$P_{org,x} = \frac{S_x}{f_{cv}} \cdot f_{p,x} \tag{3.13}$$

Where:

S_x – respective COD fraction of the influent for USO and UPO/BPO
$f_{n,x}$ – Is the respective mass ratio of nitrogen for USO, and UPO/BPO .
$f_{p,x}$ – Is the respective mass ratio of phosphorus for USO and UPO/BPO .
The ratios indicated in Table 3.3 were used to fractionate the influent:

Table 3.3: Mass ratios used for sidestream liquor influent fractionation

Mass ratio	BPO/UPO[1]	USO	BSO
f_{cv}	1.49	1.493	n/a
f_n	0.08	0.131[2]	n/a
f_p	0.04	0.03	n/a

[1] These ratios were taken from an analysis done on PS and WAS AD effluent sludge done by Ikumi (2011) and weighted according the PS and WAS split in this study.

[2] This mass ratio was calculated as the S_{us} was calculated from Equation 3.10 and the N_{ous} was assumed to be 300 mgN/l (Higgins *et al.*, 2017).

3.2.1.2 Biological Processes

The model considers the following biological processes:

1. The growth and death of OHO sludge due to the presence of biodegradable COD as per Equations 2.7a-2.7m which is occurring as suspended flocculant sludge in the bulk liquid.
2. The conversion of ammonia to nitrite by AOB's through partial nitritation via the catabolic pathway of the AOB's as per Equation 2.5 occurring in the surface layers of the granule.

3. The conversion of ammonia and nitrite to nitrogen gas via the catabolic pathway of anammox bacteria as per Equation 2.6 occurring in the centre of the granule.

It should be noted that for simplification purposes, the following biological processes have not been considered:

1. The growth and death of NOB's (it was assumed that the conditions were ideal for the AOB's to outcompete the NOB's; see below)
2. The growth and death of OHO's under anoxic conditions i.e. denitrification.
3. The death of AOB and Anammox bacteria, and the subsequent release of organic and inorganic material.

These assumptions were made based on ideal single stage anammox reactor operation whereby the higher temperatures of the reactor favours the growth of AOB's over NOB's leading to AOB's outcompeting the NOB's (Lackner *et al.*, 2014). It is also assumed that the control of the aeration is ideal such that the DO concentrations favour AOB's, therefore, comfortably outcompeting NOB's (Lackner *et al.*, 2014). In addition to this, the reactor operates as a single continually stirred tank reactor with a uniform DO concentration of 0.3mg/l (Volcke *et al.*, 2010) which is sufficient to ensure that anoxic conditions in the bulk liquid do not occur and that all biodegradable COD is consumed aerobically resulting in negligible denitrification.

3.2.1.3 Kinetics

The kinetic parameters and growth yields used for biological process 1 is listed in Table 3.1. Nitrification and denitrification parameters are not applicable due to these processes not being considered as discussed in Section 3.2.1.2. The kinetics of biological process 2 and 3 were also not considered in the model due to the limitations imposed by the steady state model. Volcke *et al.* (2010) and Mozumder *et al.* (2014) present dynamic models involving Monod kinetics whereby biomass growth and death within multiple layers of a granule are calculated. Due to this complexity, deriving steady state equations from the dynamic model, as done for conventional AS systems (Ekama, 2011b), is prohibitive. This model does consider the amount of autotrophic biomass required at steady state conditions which is discussed in Section 3.2.1.4. The outcome of the PN/A process stoichiometry was calculated based on assumed conversions in literature using mass balanced stoichiometry and the performance of existing full-scale single stage PN/A reactors.

Table 3.4: Kinetic parameters used for the PN/A process model.

	PN/A Process	**Reference**
OHO growth/death		
Yield Coefficient, Y_h (mg VSS/mg COD)	0.45	(Ekama, 2011b)
Endogenous respiration rate, b_h (/d)	0.43[1]	
Endogenous residue fraction, f_{EH} (mgVSS/mgVSS)	0.2	
Fraction of Inorganic content of OHO (mgISS/mgOHOVSS)	0.15	

1 – $b_{h,20}$ = 0.24. Adjusted for temperature using:

$b_{h,T} = b_{h,20} \cdot 1.04^{(T-20)}$ (Ekama, 2011b)

3.2.1.4 Reactor Sizing

The reactor sizing was based on literature volumetric Nitrogen Loading Rates (NLR) (kg N/m^3/d) that ensure the concentration of ammonia in the reactor does not reach levels that will inhibit the biological processes at design loads (Jin *et al.*, 2012). In a survey conducted by Lackner (2014), commercially operated granular sludge reactor loading rates vary between 0.04 - 0.65 kg N/m^3/d. Subsequently, a loading rate of 0.5 kg N/m^3/d was selected for this design based on the findings by Figdore *et.al.* (2011) that studied reactor loading rates treating sidestream liquor from a THP digestion process similar to the process in this case study. The total reactor volume was calculated from:

$$Vp = \frac{Q_i \cdot N_{ti}}{NRL} \tag{3.13a}$$

For the purpose of this study, it was assumed that the PN/A process was split into two duty process trains i.e. 2 reactors and 2 settling tanks. One of the reasons was to allow for greater turndown capacity when starting up the PN/A process at low nitrogen loads i.e. start up a single reactor at a time. Secondly, running two reactors minimizes the risk of losing all the anammox sludge through washout or through irreversible inhibition due to poor process control. If such instances occurred on one of the reactors, the “healthy” reactor could be used to seed the “unhealthy” one.

The MLSS consists of a combination of autotrophic granular sludge (MX_{tA}) and OHO suspended sludge (MX_{tH}). MX_{tH} was calculated from Equations 2.7e-2.7j. MX_{tA} was calculated assuming a VSS/TSS ratio of 0.8 in the granules:

$$MX_{tA} = \frac{MX_{VA}}{0.8} \tag{3.13b}$$

Where MX_{VA} was calculated from equation 3.31 in Section 3.2.1.13. The total reactor MLSS was calculated using the following equation:

$$MLSS = \frac{MX_{tH} + MX_{tA}}{Vp} \tag{3.13c}$$

3.2.1.5 Oxygen Demand

The model calculates the biological Actual Oxygen Demand (AOR) by taking into account the OHO carbonaceous oxygen demand and the AOB oxygen demand (biological process 1 and 2 in Section 3.2.1.2). The OHO carbonaceous oxygen demand is a function of the growth of OHO active biomass and endogenous respiration which are dependent on sludge age and the amount of utilizable biodegradable COD in the sidestream influent as per equation 2.7k.

The PN/A model assumes ideal separation of granular anammox sludge from flocculant OHO sludge in the sludge recycle line consequently decoupling the sludge age of the OHO sludge from the granular anammox sludge. In this case the sludge age of the OHO sludge is wasted continuously and is assumed to be equal to the hydraulic retention time of the reactor (V_p/Q_i). The oxygen demand for the AOB's was determined from the anammox reaction stoichiometry as per Equation 2.5 in Section 2.2.2.3

The AOR was converted to a Standard Oxygen Requirement (SOR) to calculate the size of the aeration system using the following equations(Ekama, 2011a):

$$\frac{AOR}{SOR} = \frac{[C_{s,corrected} \cdot \beta - C_L] \cdot \left(1.024^{(T-20)}\right) \cdot \alpha \cdot F}{9.07} \tag{3.14}$$

$$C_{s,corrected} = C_{s,T} \cdot \frac{(P_{site} + h' - p_{site})}{(760 - 17.51)} \qquad (3.15$$

$$C_{s,T} = C_{s,20} \cdot \frac{51.6}{(31.6 + T)} \tag{3.15a}$$

$$h' = \frac{1000 \cdot h}{13.6} \tag{3.15b}$$

$$P_{site} = 10^{(2.88117 - 0.000053407 \cdot H)} \tag{3.15c}$$

$$p_{site} = 10^{\left(8.10765 - \frac{1750.286}{T+235}\right)} \text{ (Felder and Rousseau, 2000)} \tag{3.15d}$$

Where:

SOR – Standard Oxygen Requirement, aeration equipment capacity measured by the amount of oxygen the equipment can deliver in potable water at 20°C (kgO_2/h)

β – salinity-surface tension correction factor, 0.95

α – oxygen transfer correction factor, 0.6

F – fouling factor, 0.9

C_L – Oxygen concentration in the aeration tank (mg/l), 2 mg/l for MLE (Ekama, 2011b) and 0.3 mg/l for PN/A (Volcke *et al.*, 2010)

$C_{s,T}$ – Oxygen saturation concentration corrected for temperature (mg/l)

$C_{s,corrected}$ – average dissolved oxygen saturation concentration in clean water in aeration tank at temperature T and altitude H (mg/l)

$C_{s,20}$ - dissolved oxygen saturation concentration in clean water in aeration tank at 20°C, 9.07 (mg/l)

T – operating temperature (°C), 24°C for MLE and 35°C for PN/A

P_{site} – atmospheric pressure at altitude H (mmHg)

p_{site}– vapour pressure of water at water temperature T, (mmHg)

H – altitude at site (m)

h – Depth of FBDA network

It was assumed that Fine Bubble Diffused Aeration (FBDA) equipment would be used to deliver oxygen to the system. Knowing the SOR, the total power consumption of the system was calculated using an assumed standard efficiency for FBDA equipment of 5 kgO_2/kWh (Redmon, 2018) and a shaft to wire efficiency of 70% (Ekama, 2011a).

3.2.1.6 Reactor Alkalinity Consideration

The optimum pH range for an PN/A process is 6.7-8.3 (Van Hulle *et al.*, 2010). Inhibition of the process can occur outside of this range and should be controlled to a neutral pH (Jin *et al.*, 2012). If there is insufficient alkalinity in the sidestream liquor, additional chemical dosing (in the form of sodium hydroxide, sodium bicarbonate etc.) is required to prevent the pH from dropping below the optimal range. The model calculates an alkalinity change due to the various biological processes occurring. As shown in Equations 2.5 and 2.6 in Section 2.2.2.3, the PN/A process consumes alkalinity. From the concepts described by Loewenthal *et al.* (1989) and Ekama (2011b) and considering the stoichiometric equations in the biological processes occurring within the PN/A reactor, the net change in alkalinity was calculated as follows:

$$\Delta Alk_{system} = Alk_{feed} + Alk_{ammonification} - Alk_{OHO\ growth} - Alk_{nitritation} - Alk_{anammox} \tag{3.16}$$

Where:

Alk,$_{feed}$ = alkalinity present in the side stream liquor
Alk,$_{ammonification}$ = alkalinity gain from the ammonia releases in the breakdown of biodegradable COD.
Alk,$_{OHO\ growth}$ = alkalinity lost from the ammonia uptake during OHO sludge growth
Alk,$_{nitritation}$ = alkalinity lost from the conversion of ammonia to nitrite
Alk,$_{anammox}$ = alkalinity gain from the conversion of ammonia to nitrogen gas

The model calculates the amount of sodium hydroxide to be dosed to ensure that Alk$_{system}$ > 50 (Ekama, 2011b). For the purpose of this design, the alkalinity of the sidestream liquor was assumed to be 5500 mgCaCO$_3$/l which is a reasonable average of the alkalinity ranges stated in various literature sources (Zhang *et al.*, 2016; Gu *et al.*, 2018).

3.2.1.7 Heat Balance

The PN/A process has an optimal temperature set point of approximately 35°C (Van Hulle *et al.*, 2010). Since this occurs above general ambient wastewater temperatures, the heating requirements were assessed in the model by performing an energy balance over the major equipment units. The amount of heating required for the process is equal to the sensible heat (Q$_{sensible}$) requirement to heat up the incoming sidestream feed to 35°C plus any additional heat lost to the environment (Q$_{lost}$). The heat balance was conducted based on the following equations (Felder and Rousseau, 2000):

$$Q_{heating} = Q_{sensible} + Q_{lost} \tag{3.17}$$

$$Q_{sensible} = \dot{M}_{SRL} \cdot Cp_{SRL} \cdot (T_{process} - T_{SRL}) \tag{3.18}$$

The heat lost to the environment from the major equipment units was calculated from the following equation(The Engineering Toolbox, 2021):

$$Q_{lost} = U_{overall} \cdot A \cdot (T_{process} - T_{ambient}) \tag{3.19}$$

For the heat loss calculation, only the surface areas of the major units – equalization/buffer tank and reactor(s) were considered. The overall heat transfer co-efficient was calculated from (The Engineering Toolbox, 2021):

$$\frac{1}{U_{overall} \cdot A} = \frac{1}{h_{ci} \cdot A_i} + \Sigma \left(\frac{s_n}{k_n \cdot A_n}\right) + \frac{1}{h_{co} \cdot A_o} \tag{3.20}$$

The AD process is expected to occur at 40°C, however, to account for energy losses through the downstream processes i.e. struvite precipitation, dewatering etc. it was assumed that the side stream liquor enters either the PN/A or MLE process at 25°C. The ambient temperature ($T_{ambient}$) was taken as the daily average (winter average) or daily low (winter low) of the winter months for Cape Town i.e. June, July and August (Climate-Data.org, 2020). Winter average and winter low temperatures were assumed to be 13°C and 8°C respectively. Heat generated through biological activities and via the FBDA system (warm air) were ignored as a worst-case scenario.

It is important to compare the heating demand of the sidestream processes with the heating energy available. CHP engines are commonly installed on AD plants where the combustion of biogas within the CHP engine co-generates electricity and heat (Clarke Energy, 2020). Heat energy is in the form of high-grade energy (steam) and low-grade energy (hot water approximately 85°C). Hot water from the CHP engines is used to heat the PN/A process using internal heating coils (Colsen, Site visit to Hengelo WWTP, November 2018). The amount of low grade energy available is typically 25% of the total energy recovered from the methane produced during digestion (Clarke Energy, 2020). The total amount of low-grade energy available for heating the side stream treatment processes was calculated by calculating the approximate flowrate of methane generated in the digestion process using the following equation:

$$Q_{CH4} = FX_{TSS} \cdot \frac{VSS}{TSS} \cdot \omega \qquad (3.21a)$$

Where

Q_{CH4} = Volumetric flowrate of methane in the biogas (Nm^3/d)

FX_{TSS} = Mass flux of sludge to the digestion facility (ton TSS/d)

$VSS/_{TSS}$ = Fraction of VSS in the combined sludge flow to the digester, 0.734 as per nutrient mass balance

ω = specific methane production, 300 Nm^3/ton VSS_{added}.

A conservative specific methane production was selected for this study. Ranges of 295 Nm^3/ton VSS_{added} (Jeong *et al.*, 2019) up to 341 Nm^3/ton VSS_{added} (Higgins *et al.*, 2017) have been reported for AAD processes with THP temperatures of 160°C.

The total energy available for low grade heating purposes was then calculated from:

$$Q_{LHG} = \frac{LHV_{CH4} \cdot Q_{CH4} \cdot \varepsilon_{LGH}}{(3600 \cdot 24)} \qquad (3.21b)$$

Where:

Q_{LGH} = Low grade heat available (MW)

ϵ_{LGH} = Heat recovery efficiency of low grade heat, 0.25 (Clarke Energy, 2020)

LHV_{CH4} = Lower heating value of methane (MJ/Nm^3) taken as 8.026 x 10^8 J/kmol (Perry, Green and

Maloney, 1997) and converted as follows using the ideal gas law PV=nRT:

$$LHV_{CH4} = 8.026\ x\ 10^8 \left[\frac{J}{kmol}\right] \cdot \frac{101325[Pa]}{8.314[J \cdot mol^{-1} \cdot K^{-1}] \cdot 293[K]} \cdot \frac{1[kmol]}{1000[mol]} \cdot \frac{1[MJ]}{10^6[J]} \quad (3.21c)$$

3.2.1.8 Mixing and Pumping Energy

The model calculates the mixing energy requirment using a factor of 5 W/m^3 (Tchobanoglous *et al.*, 2003; De Ketele, Davister and Ikumi, 2018). For the PN/A process, mixers will only be running during periods when the aeration rate is not high enough to sustain adequate mixing. As this depends on transient conditions in the reactor, which the steady state model cannot simulate, a factor of 0.25 was applied to the mixing energy to account for times when the mixers are off.

Pumping energy is calculated by (Perry, Green and Maloney, 1997):

$$E_{pumping} = \frac{\frac{Q}{3600} \cdot \rho \cdot g \cdot \Delta h}{\eta_{eff}} \quad (3.22)$$

Where
E = Pumping energy (kW)
Q = Volumetric flowrate being pumped (m^3/h)
ρ = liquid density (1000 kg/m^3)
g = 9.81 (m/s^2)
Δh = differential head over the pump
η_{eff} = hydraulic pumping efficiency (shaft to wire efficient) assumed to be 50%. This can change depending on the pumps selected however the same efficiency has been used in both PN/A and MLE models for comparative purposes.

3.2.1.9 Sludge Production

Sludge is produced due to the growth of OHO, AOB, NOB and Anammox bacteria as well as the accumulation of unbiodegradable and inert particulates. Due to the low specific growth yields of AOB, NOB and Anammox sludge, together with their significantly longer sludge ages, their contribution to the sludge mass is ignored in the model.

The bulk of the sludge produced by the PN/A process is from OHO growth, as well as the inert organic and inorganic solids in the feed and was calculated from Equation 2.7d-2.7j. Accepting that the sludge age of the OHO bacteria in the PN/A reactor is very short (approximately equal to the hydraulic retention due to continuous wasting), the daily sludge production was calculated from

Equation 2.7l.

3.2.1.10 Effluent Quality and Process Efficiency

The model calculates the following effluent parameters and fractionates them into Organic/Inorganic and soluble/particulate phases where applicable:

COD
TKN
TP
NO3/NO2
FSA
OP
TSS
ISS

Since the treated sidestream liquor is returned back to the mainstream works, the model combines the waste sludge and the effluent overflow into a single waste stream.

For COD, S_{te} is calculated by assuming all the S_{bi} (S_{bsi}+S_{bpi}) is consumed by OHO bacteria. The effluent COD is subsequently made up of S_{use} (which is equal to the S_{usi} in the influent, as unbiodegradable solubles are not produced during the process) and the COD from the continuously wasted OHO sludge.

$$S_{te} = S_{usi} + \frac{MX_v}{Vp} \cdot f_{cv} \qquad (3.22a)$$

The Ammonia (FSA) (N_{ae}) was calculated in the effluent stoichiometrically by setting the ammonia conversion of Equation 2.5 such that an overall ammonia conversion of 90% is achived as supported by literature (Van Hulle *et al.*, 2010). Nitrate was calculated in the effluent stoichiometrically from Equation 2.6 assuming a 100% conversion of nitrite. This provides an overall TIN conversion of 80% which is supported by literature (Van Hulle *et al.*, 2010).

The TKN, OP, TSS and ISS concentrations in the effluent were calculated with the following equations:

$$N_{te} = N_{ousi} + \frac{MX_v \cdot f_n}{R_{s,OHO} \cdot Q_i} + N_{ae} \qquad (3.22b)$$

$$P_{se} = P_{ti} - \frac{MX_v \cdot f_p}{R_{s,OHO} \cdot Q_i} \qquad (3.22c)$$

$$X_{te} = \frac{MX_t}{R_{s,OHO} \cdot Q_i} \quad (3.22d)$$

$$X_{IOe} = \frac{(MX_t - MX_v)}{R_{s,OHO} \cdot Q_i} \quad (3.22e)$$

Due to the continuous wasting of OHO sludge, $R_{s,OHO}$ was made equal to the hydraulic retention time $R_{hn.}$

Due to the sidestream liquor being returned to the mainstream treatment process, the effluent pollution load in the sidestream liquor results in an incremental increase in sludge generation and aeration required at the mainstream works. These incremental increases were calculated by taking the fractionated effluent quality concentrations and total flow of the sidestream process and calculating the wasted sludge mass (FX_t) and the additional aeration requirement (FO_T) from the steady state model equations listed in Table 2.3. It was assumed that the mainstream works operates at a sludge age of 15 days and at an MLSS of 4 g/l. The additional aeration power required at the mainstream works was calculated from Equations 3.14 and 3.15a-d (Section 3.2.1.5) and making the same assumptions with regards to alpha, beta and FBDA efficiency discussed in Section 3.2.1.5.

The PN/A model determines the process efficiency by calculating the percentage ammonia removal and the percentage Total Inorganic Nitrogen (TIN) removal using the following equations:

$$\%N_{a,rem} = (\frac{N_{ae}}{N_{ai}} - 1) \cdot 100 \quad (3.22f)$$

$$\%TIN_{rem} = (\frac{N_{ai} - N_{ae} - N_{ne}}{N_{ai}}) \cdot 100 \quad (3.22g)$$

Equation 3.22g assumes all inorganic nitrogen in the influent is in the form of ammonia. This assumption is reasonable considering the nitrates/nitrites typically account for less than 1% of the total inorganic nitrogen in the sidestream influent as indicated in Table 2.1 and Table 2.2. The removal of TIN is an important metric to measure the performance of the PN/A system. High levels of ammonia removal through nitrification/nitritation are still achievable even if the anammox performance is poor. In this instance, high nitrates will occur in the effluent. The %TIN removal specification quantifies the efficiency of the process in converting ammonia to nitrogen gas which is the end goal.

3.2.1.11 Settling Tank

The capacity and size of the settling tank is calculated in the model using one dimensional idealized flux theory (Ekama, 2012). The maximum overflow rate in the settling tank is calculated knowing the mixed liquor concentration from the reactor (X_f) using the following equation:

$$q_{Amax} = V_0 \cdot e^{(-n \cdot Xf)} \tag{3.23}$$

V_0 and n are calculated with an assumed DSVI of 125 ml/g via the following relationship:

$$SSVI = DSVI \cdot 0.67 \tag{3.24}$$

$$\frac{V_0}{n} = 67.9 \cdot e^{(-0.016 \cdot SSVI)} \tag{3.25}$$

$$n = 0.88 - 0.393 \cdot \log\left(\frac{V_0}{n}\right) \tag{3.26}$$

A DSVI of 125 ml/g is a reasonable worst-case scenario for flocculant sludge (Ekama, 2012). For the PN/A process, the settling velocity of granules is much higher than the flocculant sludge. Lu *et al.* (2013) reported a settling velocity of anammox granules > 54 m/h while the settling velocity of flocculant sludge is typically <1.5 m/h (Ekama, 2012). Since there is a combination of flocculant and granular sludge in the PN/A system, selecting an overflow rate as if it was pure flocculant sludge is a worst-case assumption as in reality it is likely that much higher overflow rates would be acceptable.

The area of the settling tank (A_{ST}) is then calculated from (Ekama, 2012):

$$A_{ST} = \frac{Q}{q_{Amax} \cdot F} \tag{3.27}$$

Where:

Q – volumetric flowrate to the settling tank (m^3/h)

F – the flux rating factor. Assumed to be 80%.

3.2.1.12 Mass Balance Verification

The COD mass balance is used as a tool for verifying the accuracy of results calculated from steady-state models (Ekama, 2011b). In addition to COD, mass balances can be conducted over other elements of the process (phosphorus and nitrogen) to ensure all components of the system are correctly accounted for. When the mass balances are complete, the model is verified to be internally consistent and allows for confidence in the calculated results. In this regard, the steady state PN/A model is verified by conducting a mass balance on COD, nitrogen and phosphorus.

COD is balanced according to the following equation:

$$F_i \cdot S_{t,i} = F \cdot O_c + F_e(S_{us,e} + f_{cv,OHO} \cdot X_{v,OHO}) \tag{3.28}$$

Wasted sludge is combined with the overflow into a combined effluent stream recycled back to the mainstream works therefore $Q_i = Q_e$. The effect of autotrophic biomass growth on the COD balance is ignored as it is assumed that the sludge age is sufficiently long to result in negligible sludge production when compared to the amount of OHO sludge generated.

Nitrogen is balanced according to the following equation:

$$F_i N_{t,i} = F_i N_{N2} + F_e N_{a,e} + F_e N_{NO,e} + F_e\left(N_{ous,e} + f_{n,OHO} \cdot X_{v,OHO}\right) \tag{3.29}$$

Phosphorus is balanced according to the following equations:

$$F_i P_{t,i} = F_e P_{OP,e} + F_e\left(P_{ous,e} + f_{p,OHO} \cdot X_{v,OHO}\right) \tag{3.30}$$

Refer to Figure 8.1 which presents the results of the mass balances.

3.2.1.13 Seed Sludge Requirements

Obtaining seed sludge was listed as a significant risk to the PN/A process as discussed in Section 2.3.2. and is therefore a crucial parameter to calculate as a strategy and cost for obtaining the seed sludge needs to be determined. This is particularly relevant to South African conditions as seed sludge would need to be transported from the United States of America (USA) or Europe. The quantity of seed sludge required was calculated assuming the following:

- Seed sludge is granular with radius > 500 µm and the amount of anammox bacteria in each granule is 27% (Li *et al.*, 2020)
- The Total Autotrophic Nitrogen Removal Rate (TANRR) possible from the sludge is 0.14 kg N/kg VSS/d (Li *et al.*, 2020)
- Doubling time of anammox sludge of 11 days (van der Star *et al.*, 2007; Eskicioglu, Galvagno and Cimon, 2018)
- The VSS/TSS fraction of the granules is 0.8 and the solids concentration of the seed material being transported is 30 kg TSS/ m^3 (Colsen, personal E-mail communication, 20 May 2020)

The amount of autotrophic granular sludge required (VSS_{auto}) was related to total nitrogen load removed by the following equation (Li *et al.*, 2020):

$$MX_{VA} = \frac{TN\ removal[kg\ N \cdot d^{-1}]}{TANRR[kg\ N \cdot d^{-1} \cdot kgVSS^{-1}]} \tag{3.31}$$

Knowing the amount of granular sludge required to achieve the required nitrogen removal at full load ($VSS_{auto,final}$), the initial seed sludge ($VSS_{auto,initial}$), was related to the start up time (t) by the following equation:

$$t\ [days] = \left[\frac{\log\left(\frac{MX_{VA,final}}{MX_{VA,initial}}\right)}{log2} \right] \cdot 11 \tag{3.32}$$

The volume of seed sludge was them calculated as follows:

$$Volume\ [m^3] = \frac{MX_{VA}[kgVSS]}{0.8[kgVSS/kgTSS] \times 30[kgTSS/m3]} \tag{3.32a}$$

3.2.2 Sidestream treatment – MLE process

In this model the sidestream liquor was treated using a conventional activated sludge process with an MLE reactor configuration. The process was modelled as a high rate MLE reactor whereby the process would be maintained at approximately 24°C to benefit from the improved denitrification kinetics. The model allows for the addition of a COD source into an equalization tank upstream of the MLE reactor. A simplified process flow diagram of the MLE process is indicated in Figure 3.4 below.

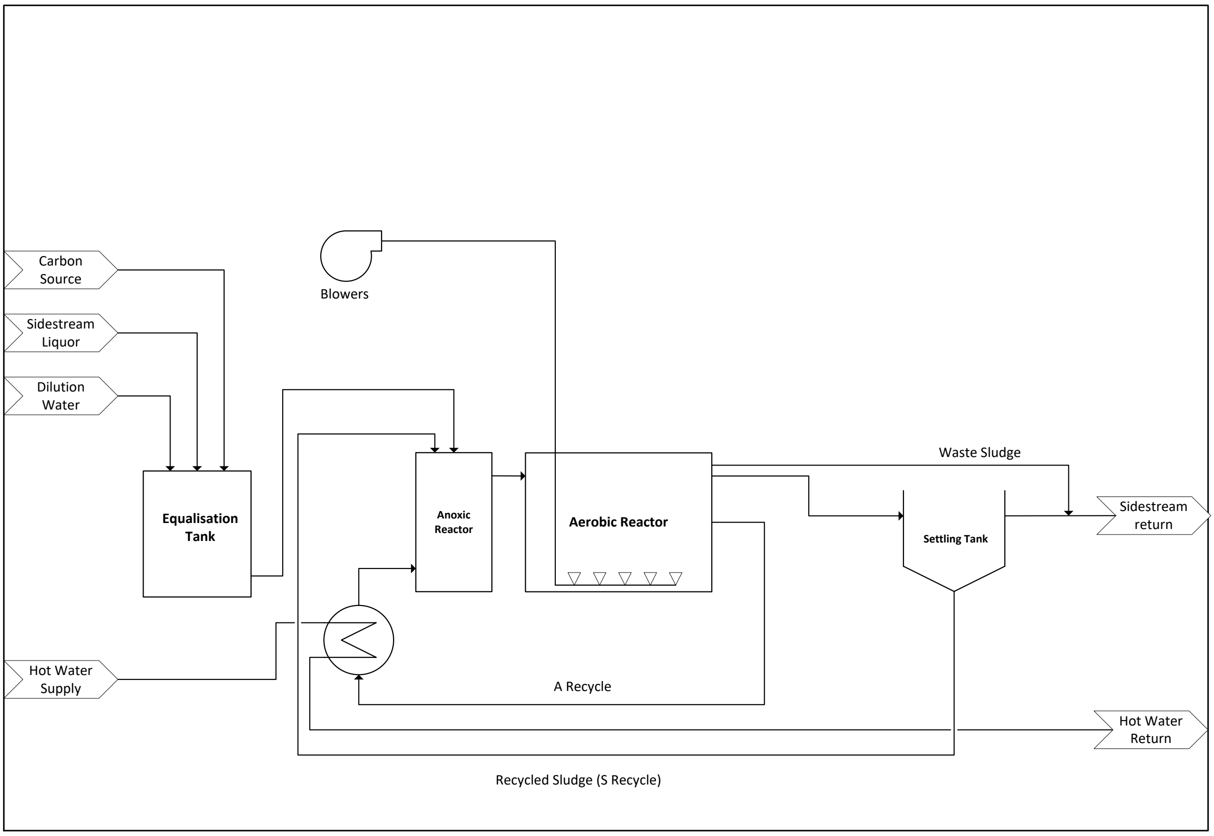

Figure 3.4: A simplified process flow diagram of the MLE process.

The model is based on steady state equations and all calculations were performed using a spreadsheet. The model boundary starts at the discharge of the dewatering unit operation and ends at the effluent discharge from the MLE settling tank. The model uses the sidestream liquor chemical quality parameters as inputs. A list of all inputs into the model and associated assumptions is listed in Section 8.2 in the appendix.

3.2.2.1 Influent Fractionation

The influent fractionation is done using the same method described for the PN/A model (Section 3.2.1.1) except in this instance, there is an additional step to include the addition of dosing an organic carbon source. The model calculates the amount of organic dosing required to ensure that the blended influent feed to the reactor has a TKN/COD ratio sufficient to achieve the maximum level of denitrification possible. To provide a fair comparison of the MLE and PN/A processes, the TKN/COD set points were decided such that the MLE process produces the same nitrate load in the effluent as the PN/A process. The TKN/COD ratio was solved for iteratively until the nitrate load in the effluent equaled the PN/A process. The value for the TKN/COD ratio came to 0.215.

The organic source used for dosing in this study was methanol. Methanol was selected due to it

being used extensively as an organic source in wastewater treatment around the world, it's relatively low cost and high COD value (1,185 g/l), and due to it being readily available in South Africa.
The volumetric flow rate of methanol (Q_{meth}) was calculated from the following equations:

$$S_{ti} = \frac{S_{t,SL} \cdot Q_{SL} + S_{t,meth} \cdot Q_{meth}}{Q_{SL} + Q_{SL}} \tag{3.32b}$$

$$N_{ti} = \frac{N_{t,SL} \cdot Q_{SL}}{Q_{SL} + Q_{SL}} \tag{3.32c}$$

$$\frac{N_{ti}}{S_{ti}} = 0.215 \tag{3.32d}$$

Where $S_{ti}/S_{t,SL}/S_{t,meth}$ refers to the total COD concentration in the blended influent to the MLE, the total COD concentration in the sidestream liquor and the total COD concentration of the methanol respectively. Similarly, $N_{ti}/N_{t,SL}$ refer to the total TKN concentration of the blended influent to the MLE and the sidestream liquor respectively.

3.2.2.2 Biological processes

The following biological processes were considered in the MLE model:

1. The aerobic growth and death of OHO sludge due to the presence of biodegradable COD as per Equations 2.7a-2.7m
2. The growth and death of nitrifiers (AOB and NOB considered as one, i.e. ANO's) as per Equations 2.8a-2.8f.
3. The anoxic growth and death of OHO sludge within the anoxic zone as per Equations 2.8i-2.8l.

It is important to note that the growth of Polyphosphate Accumulating Organisms (PAOs) has not been considered in the model as the anoxic zone is assumed to be completely anoxic meaning that there are no areas within the zone that have turned anaerobic such that PAO growth occurs. The basis of this assumption is that the size of the anoxic zone is selected to completely denitrify the nitrate load in the A recycle based on the kinetics (refer to Section 3.2.2.3 and 3.2.2.6).

3.2.2.3 Kinetics

The kinetic parameters and growth yields used for biological process 1 is listed in Table 3.5. Due to the substantially different yield coefficients of methanol and acetate (Dold *et al.*, 2008; Ekama, 2011b), the overall yield co-efficient was weighted based on the substrate split between acetate and methanol. Nitrification kinetic parameters used for biological process 2 and denitrification kinetic parameters used for biological process 3 are also indicated in Table 3.5.

Table 3.5: Kinetic parameters used for the MLE process model

	MLE	**Reference**
OHO growth/death		
Yield co-efficient Acetate, $Y_{h,acetate}$ (mg VSS/mg COD)	0.45	(Ekama, 2011b)
Yield co-efficient Methanol, $Y_{h,methanol}$ (mg VSS/mg COD)	0.27	(Dold *et al.*, 2008)
Weight Yield Coefficient, Y_h (mg VSS/mg COD)	0.33[1]	1 -
Endogenous respiration rate, b_h (/d)	0.28[2]	(Ekama, 2011b)
Endogenous residue fraction, f_{EH} (mgVSS/mgVSS)	0.2	
Fraction of Inorganic content of OHO (mgISS/mgOHOVSS)	0.15	
Nitrifier growth/death		
Maximum specific growth rate of nitrifiers, μ_{Am} (/d)	0.72[3]	(Ekama, 2011b)
Half saturation constant, K_n (mg NH_3-N)/l)	1.59[4]	
Specific endogenous mass loss rate of nitrifying bacteria, b_{AT} (mg/mg/d)	0.045[5]	
Yield Coefficient for Nitrogen, Y_n (mg VSS/mgN)	0.1	
Kinetic constant of degradation, K_r	0.017[6]	
Denitrification		
Denitrification rate constant with RBCOD in anoxic zone K1 (mgNO_3-N/mg VSS/d)	1.49[7]	(Ekama, 2011b)
Denitrification rate constant with SBCOD in anoxic zone K2 (mgNO_3-N/mg VSS/d)	0.137[8]	

[1] – Yield co-efficient weighted based on the split of methanol and acetate substrate

[2] – $b_{h,20}$ = 0.24. Adjusted for temperature using:

$b_{h,T} = b_{h,20} \cdot 1.04^{(T-20)}$ (Ekama, 2011b)

[3] - $\mu_{A,20}$ = 0.24. Adjusted for temperature using:

$\mu_{Amax} = \mu_{A,20} \cdot 1.123^{(T-20)}$ (Ekama, 2011b)

[4] – K_N = 1. Adjusted for temperature using:

$K_{N,T} = K_{N,20} \cdot 1.123^{(T-20)}$ (Ekama, 2011b)

[5] – $b_{A,20}$ = 0.040. Adjusted for temperature using:

$b_{A,T} = b_{A,20} \cdot 1.029^{(T-20)}$ (Ekama, 2011b)

[6] – $K_{r,20}$ = 0.015. Adjusted for temperature using:

$K_{r,T} = K_{r,20} \cdot 1.029^{(T-20)}$ (Ekama, 2011b)

[7] – $K_{1,20}$ = 0.72. Adjusted for temperature using:

$K_{1,T} = K_{1,20} \cdot 1.2^{(T-20)}$ (Ekama, 2011b)

[8] – $K_{2,20}$ = 0.101. Adjusted for temperature using:

$K_{2,T} = K_{2,20} \cdot 1.08^{(T-20)}$ (Ekama, 2011b)

3.2.2.4 Reactor Sizing

The model calculated the reactor size by selecting a sludge age that would satisfy complete nitrification. For this study, a sludge age of 4 days was selected. In order to ensure that the comparison between the MLE and PN/A processes were comparable, the size of the anoxic zone was selected to provide partial denitrification to match the nitrate load being returned to Cape Flats WWTP with the PN/A process. The sludge mass was from equations 7d – 7j in Table 2.3.

A mixed liquor concentration (MLSS) of 4 g/l was selected and used for calculating the size of the settling tank. This is a typical MLSS concentration for AS systems that would provide a reasonable reactor and settling tank size considering that the reactor size decreases with an increasing MLSS concentration but the converse is true for the settling tank (Ekama, 2016). Knowing the sludge age and mixed liquor concentration (and subsequently the mass of total solids in the reactor), the total reactor volume was calculated from:

$$Vp = \frac{M(X_t)}{MlSS} \tag{3.33}$$

3.2.2.5 Oxygen Demand

The model calculates the biological Actual Oxygen Demand (AOR) by taking into account the OHO carbonaceous oxygen demand and the AOB oxygen demand. The OHO oxygen demand was calculated using the same equation used for the PN/A model (Equation 2.7k, Section 2). The AOB oxygen demand was calculated using Equation 2.8b, 2.8f-h and 2.8n. The AOR was converted to an SOR and power consumption using the same method described for the PN/A process in Section 3.2.1.5.

3.2.2.6 Internal A recycle

The selection of the internal A recycle is an important input consideration for the MLE model. The function of the A recycle is to recirculate nitrates, generated in the aerobic zone, back to the anoxic zone (as indicated in Figure 2.1) for denitrification to occur (Ekama, 2011b). The objective of the A recycle is to balance the nitrate load recycled from the aerobic zone with the denitrification potential of the anoxic zone. At this condition, an optimum A recycle (A_{opt}) is achieved allowing for maximum denitrification to occur, which results in the minimum concentration of nitrates in the effluent. The denitrification potential of the anoxic zone is determined by (Ekama, 2011b):

$$D_p = S_{bs,i} \cdot \frac{(1 - f_{cv} \cdot Y_h)}{2.86} + S_{b,i} \cdot \frac{K_2 \cdot fx1 \cdot Y_h \cdot R_s}{\left(1 + b_{H,T} \cdot R_s\right)} \tag{3.34}$$

Where fx1 refers to the anoxic zone mass fraction.

The maximum anoxic mass fraction needs to consider the minimum aerobic mass fraction required for nitrification. This is achieved using Equation 3.34a which takes into account the maximum growth rate and death rate of nitrifying bacteria (Ekama, 2011b):

$$fx1 = 1 - Sf \cdot \frac{(b_{AT} + \frac{1}{R_S})}{\mu_{AmT}} \tag{3.34a}$$

SF refers to the safety factor applied to the minimum sludge age required for nitrification and is generally accepted as 1.25 (Ekama, 2011b)

The nitrate load (mg NO_3-N/l influent) to the anoxic zone is determined by (Ekama, 2011b):

$$N_L = \left(\left(\frac{N_c}{A + S + 1}\right) + \frac{O_a}{2.86}\right) \cdot A \tag{3.35}$$

Where

N_c = Nitrate concentration in to the aerobic (mg NO_3-N/l)

A = A recycle ratio

S = S recycle ratio

O_a = Dissolved oxygen concentration in the A recycle.

When N_L = D_p, the nitrates in the effluent are calculated from (Ekama, 2011b):

$$N_{n,e} = \left(\frac{N_c}{A + S + 1}\right) \tag{3.36}$$

3.2.2.7 Settling Tank

The capacity and size of the settling tank is calculated in the model using one dimensional idealized flux theory (Ekama, 2012) as per the methodology outlined in Section 3.2.1.11.

3.2.2.8 Chemical Consumption

The sidestream liquor requires an additional COD source for denitrification. Refer to Section 3.2.2.1 which discusses organic dosing and influent fractionation.

3.2.2.9 Heat Balance

An energy balance was conducted over the primary equipment using the same method described for the PN/A process in Section 3.2.1.7.

3.2.2.10 Effluent Quality and Process Performance

The MLE model calculates the same effluent parameters listed in Section 3.2.1.10. Since the treated sidestream liquor is returned back to the mainstream works, the model combines the waste sludge and the effluent overflow into a single waste stream.

The COD, TKN, OP, TSS and ISS concentrations were calculated as per Equations 3.22a–3.22e in Section 3.2.1.10. The Ammonia (FSA) (N_{ai}) concentration in the effluent was calculated from Equation 2.8b which considers the selected sludge age and anoxic zone size. The nitrate concentration in the effluent was calculated from Equation 3.36 in Section 3.2.2.6.

The incremental increase in aeration and sludge generated at the mainstream works was calculated using the same methods discussed in Section 3.2.1.10.

The process efficiency, in terms of percentage ammonia and TIN removal, was calculated as described in Section 3.2.1.10

3.2.2.11 Mass Balance Verification

Similar to the PN/A model, the MLE model is verified by conducting a mass balance on COD, nitrogen and phosphorus.

COD is balanced as per Equation 3.28 in the PN/A model in Section 3.2.1.12.

Nitrogen is balanced according to the following equation:

$$F_i \cdot N_{t,i} = F_e \cdot N_{a,e} + F_e \cdot N_{NO,e} + F_e\left(N_{ous,e} + f_{n,OHO} \cdot X_{v,OHO}\right) \qquad (3.37)$$

Phosphorus is balanced as per Equation 3.30 in the PN/A model in Section 3.2.1.12. Refer to Figure 8.3 which presents the results of the mass balances.

3.3 Effluent Quality Index

The primary purpose of a wastewater treatment facility is to achieve a desirable effluent quality that can be safely discharged and should therefore be prioritized over economic factors to a practical degree. The effluent quality of the sidestream liquor process should therefore not compromise the mainstream treatment works. To compare the effluent qualities from each of the sidestream liquor processes, an Effluent Quality Index (EQI) was used. The EQI was developed by the IWA benchmark simulation modelling task group (Jeppsson *et al.*, 2007) and modified to a South African context by de Ketele *et al*. (2018). The EQI is defined as:

$$EQI = \frac{1}{T \cdot 1000}\int_{t_0}^{t_{end}} [\beta_{COD} \cdot \left(COD(t)\right) + \beta_{FSA} \cdot \left(FSA(t)\right) + \beta_{OP} \cdot \left(OP(t)\right) + \beta_{NO} \cdot \left(NO(t)\right) + \beta_{TSS} \cdot \left(TSS(t)\right)] \cdot Q_e(t)d(t) \qquad (3.38)$$

Where:

EQI = Effluent Quality Index (kg pollution/d)

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

Q_e = total effluent flowrate (m^3/d)

$Limit_X$ = The regulatory discharge limit for the respective wastewater parameter

X(t) = Concentration of the respective wastewater parameter in the effluent

β_x = a weighting factor for each wastewater parameter relative to COD where $Limit_x$ was obtained from the Cape Flats WWTP's water use license (Department of Water and Sanitiation, 2016)

$$\beta_X = \frac{COD_{limit}}{Limit_X} \qquad (3.39)$$

The impact of each pollutant on the EQI is weighted based on the effluent limits of the mainstream works.

3.4 Capital cost estimate

Once all the major equipment for each design was specified, capital costs were estimated from the results. The capital costing was broken down into a civil portion and a Mechanical and Electrical (M&E) portion for each proposed solution. The civil portion covered:

- Earthworks
- Access Roads
- Major Concrete installations – Equalization tank, reactors and blower house

The M&E portion covered:

- Blowers and FBDA equipment
- Major pumps
- Mixers
- Piping
- Chemical Handling and dosing equipment
- Control and automation equipment – instrumentation, MCC's, cabling etc.

All costing for the PN/A process was obtained from prominent South African companies in South African Rand to make the analysis specific for the region. The following South African consulting companies contributed to the cost estimate for their respective disciplines using proprietary costing methods:

- Project Assignments – Mechanical
- JGP – Electrical
- JG Afrika – Civil

Where equipment capacity differed and actual costs were not obtained from technology suppliers, the costs were scaled with the following formula used for engineering cost estimation (Tribe and Alpine, 1986):

$$\frac{C_1}{C_2} = \left(\frac{V_1}{V_2}\right)^{\alpha} \qquad (3.40)$$

Where:

C_1 = Cost of equipment item 1

C_2 = Cost of equipment item 2

V_1 = Capacity of equipment item 1

V_2 = Capacity of equipment item 2

α = Scale co-efficient, 0.6 for most M&E equipment and 1 for civil costs

The following factors were used to account for the total Cost of Works based on the mechanical and electrical equipment cost estimates:

	Adjustment factor
Civil Contractor P&G's	0.15

Mechanical Contractor P&G's	0.2*
Electrical Contractor P&G's	0.2
Mechanical equipment Installation fee	0.2
Contingency	0.15

* Note – these factors are estimated by the various consultants based on experience and costing on previous projects of this nature.

The electrical cost for the PN/A process was costed based on the design however, for the MLE process, the electrical cost was linked to the cost of the mechanical equipment by assuming that the cost of the electrical equipment for the MLE process was the same percentage of the mechanical equipment as costed for the PN/A process. This amounted to 37.5%.

The capital cost estimate included professional services fees charged by engineering consultants to assist with tender preparation and contract management for municipal projects such as this. The calculation of the fee was based on the Engineering Council of South Africa's (ECSA) fee guidelines. The guidelines stipulate that the fee must be based on Cost of Works (COW) according to the Equation 3.41a for projects greater than R11.5m (Engineering Council of South Africa, 2015):

$$\%\,Fee = 29.9\% - 0.023 \cdot Log(COW) \tag{3.41}$$

Equation 3.41 is based on project fee category F applicable for the construction of wastewater treatment works (Engineering Council of South Africa, 2015).

3.5 Operational cost

The modelling and mass balance exercise provided a basis for calculating the energy and chemical requirements and total sludge production. The operational cost for each design was then evaluated using the Operational Cost Index (OCI) approach as per the IWA benchmark simulation modelling (BSM) task group (Jeppsson *et al.* (2007); de Ketele *et al*.,(2018); Ikumi *et. al. (*2019)) but adapted to include terms relative to these processes. The OCI is given in the equation below:

$$OCI = (AE + PE + ME + HE) \cdot Energy\ Cost + SP \cdot Sludge\ Disposal\ Cost + CA \cdot Chemical\ Cost \tag{3.42}$$

Where:

AE = Aeration energy (kWh/d)

PE = Pumping energy (kWh/d)

ME = Mixing energy (kWh/d)

HE = Heating energy ((kWh/d)

SP = Sludge produced (kgTSS/d)
CA = Chemical addition, carbon source or caustic soda (m3/d)

The aeration energy included the aeration energy of the sidestream liquor treatment process in addition to the aeration required to process the effluent generated by the respective processes at the mainstream WWTP. This was calculated knowing the fractionated effluent quality and using Equation 2.7k in Section 2.4.1 and assuming that the mainstream works operates with a sludge age of 15 days. The sludge produced (SP) refers to the incremental increase in sludge that would occur at the mainstream works because of the treated sidestream liquor being recycled. The incremental increase in sludge was calculated from Equation 2.7e-2.7j and 2.7l in Section 2.4.1.The energy cost was taken from a municipal electricity account in January 2019 from Cape Flats WWTP and escalated by the approved tariff increases published by the national power supplier, ESKOM (ESKOM, 2020b, 2020a). The chemical cost for methanol taken from a prominent chemical supplier in South Africa (Brenntag, personal email quotation, 27 Jan 2021). For this study, a price of R10.60/kg was used for methanol.

The sludge produced by both processes would be sent back to the mainstream wastewater works which will add to the indigenous sludge throughput to the regional sludge digestion facility. This was calculated knowing the fractionated effluent quality and using Equations 7e – 7l in Section 2.4.1 to calculate the incremental increase in sludge mass generated and aeration required at the mainstream WWTP assuming a sludge age of 15 days. This additional sludge would divert import sludge away from the regional digestion facility which would need to be disposed of to landfill at a higher cost. The cost for disposal of the sludge was taken as R812/ per ton of sludge as per the City of Cape Town’s rate to dispose of primary sludge to the closest landfill at Visserhok (Ladouce, 2020).

3.6 Life-cyle cost analysis

A Life-Cycle Cost Analysis (LCCA) is used to evaluate capital investment decisions which have different capital and operational costs (Fuller and Petersen, 1996). The purpose of the LCCA is to provide an accurate indication of the full financial burden for both processes over the study period.

The study period of the LCCA refers to the time frame that the analysis is performed over and is usually the investment horizon of the investor (Fuller and Petersen, 1996). A study period of 23 years was chosen. Being a municipal treatment plant, the study period was selected to be the expected full operational lifetime of each process solution (which was taken to be 20 years) plus the construction period (which was taken to be 3 years)

The base date is the point in time to which all project related costs are discounted (Fuller and Petersen, 1996). The base date was selected as the year this study was conducted, 2020. For simplification purposes, all costs were inflated and discounted in yearly increments.

The LCCA calculated the Present Value (PV) of all the costs relating to each process after escalating the present day costs to the date they will occur and then discounting them to the base date again. The following formula was used to calculate the PV for each annual amount (A) occurring at time t over n years:

$$PV = A_0 \cdot \sum_{t=1}^{n} \left(\frac{1+e)}{(1+d)}\right)^t \tag{3.43}$$

Where:

A_0 – cost at base date

n – number of years

e – escalation rate

d – discount rate

Table 3.6 indicates the inputs used in the LCCA.

Table 3.6: A summary of the key inputs into the LCCA.

Parameter	**Value**	**Source**
number of years, n	23	3 years construction, 20 year operation
methanol escalation, $e_{methanol}$	6.19 %	Table 3.7
energy escalation, e_{energy}	10.2 %[1]	(Moolman, 2019)
sludge disposal escalation, e_{sludge}	5.20 %	SA average CPI (Inflation.eu, 2021)
Discount rate	7.00 %	SA prime lending rate

[1] The average annualized energy price escalation from 2007 to 2019 was 15.2% (Moolman, 2019). To account for realistic above inflation energy price increases over the period of the LCCA, an average of the increase from 2007-2019 and CPI of 5.20% (Inflation.eu, 2021) was used.

For the purpose of the LCCA it was assumed that the capital costs would be incurred over a 3-year period during the construction phase of the project. The capital expenditure outside of the base date was escalated at an average CPI rate of 5.20% (Inflation.eu, 2021). From year 4, the operational expenses were factored into the assessment as indicated by the OCI for that year.

The chemical (methanol), energy and sludge disposal costs were escalated at their own applicable rates. For the escalation of methanol, a historical price list was obtained from a methanol supplier (Methanex, 2021). The average year on year escalation over a 20-year period was calculated from this list. Due to the price being reported in US dollars, the average year on year depreciation in the rand was factored into the escalation rate over the same period to give an escalation rate in Rand terms as indicated in Table 3.7.

Table 3.7: Methanol price escalation rate

	Rate	Source
Methanol price average escalation in US$ terms	3.21%	(Methanex, 2021)
Devaluation in Rand/Dollar terms	2.98%	(Excelrates FX, 2021)
Methanol price average escalation in Rand terms	6.19%	

The discount rate was based on the prime lending rate for South Africa. A discount rate of 7.0% was used.

4 RESULTS AND DISCUSSION

4.1 Side stream wastewater quality

The results of each stream of the mass balance conducted over the AD, struvite and dewatering process (refer to Figure 3.2) are indicated in Table 4.3. This represents the solutions of Equations 3.1 to 3.5q in Section 3.1.1. Table 4.1 below summarizes the influent characteristics of the side stream liquor, such as the flowrate and overall concentrations as calculated using the results of the mass balance from stream 7 (refer to Table 4.3) together with the methodology discussed in Section 3.1.2. It is important to note that the calculated nutrient loads are dependent on the nutrient characteristics of the incoming sludge.

Table 4.1: Calculated side stream liquor quality parameters

Parameter	Value
Volumetric flowrate (Ml/d)	2.04
COD (mgCOD/l)	4500[1]
TKN (mg N/l)	1674
FSA (mg NH3-N/l)	1355
TP (mg P/l)	186
TSS (mg/l)	511
ISS (mg/l)	142

1- Assumed value based on literature values (Zhang *et al.*, 2016; Gu *et al.*, 2018).

The calculated FSA concentration in Table 4.1 is lower than the literature values listed in Section 2.1 while the calculated TP concentration is substantially higher. The nutrient concentrations in the side stream liquor are dependent on the nutrient loads in the various sludges imported to the facility in addition to the WAS/PS split. The nutrient concentrations in the side stream stated in the literature sources could have been produced from the digestion of a higher percentage of WAS which would result in higher FSA or alternatively the nitrogen component in the WAS could have been higher. One also needs to consider the removal of nitrogen in the struvite precipitation process which may not have occurred in the literature sources. It is possible that the literature sources did not include digestion of WAS from EBPR processes in which case the contribution of phosphorus from polyphosphates in the PAO organisms would be zero whereas in this study the contribution is significant. Varying polymer dosing rates and concentrations can also result in significantly different side stream concentrations due to the dilution of digestate before dewatering. One of the main objectives of this study is to conduct the analysis relative to the conditions at Cape Flats WWTP to make the results case specific.

Table 4.2 lists the calculated nutrient loads leaving the digestion process and leaving the dewatering process in the side stream liquor.

Table 4.2: Calculated side stream liquor nutrient loads

Parameter	Value	Equation/Source
Digestate		Table 4.3
Ammonia load leaving the digester (kg N/d)	4,281	
OP load leaving the digester (kg P/d)	1,932	
Side stream		
Ammonia load in side stream (kg N/d)	2,762	
TKN load in side stream liquor (kg N/d)	3,415	
P load in side stream liquor (kg P/d	379	
% ammonia removal with struvite precipitation and dewatering	55 %	$\frac{FN_{a,7}}{FN_{a,2}} - 1$

Although the side stream liquor represents approximately 1-2% of the hydraulic load to the mainstream works, it accounts for 29% of the design TKN treatment capacity[1]. This represents a significant percentage of the mainstream treatment design capacity and is aligned with what is presented in literature sources investigating similar facilities in other parts of the world as discussed in Section 2.1. A struvite precipitation process can remove nominally 90% of the OP from the digestate (Zhou *et al.*, 2015) as well as a portion of the ammonia that is bound into the precipitated struvite ($MgNH_4PO_4\ 6H_2O$). Consequently, the struvite process in conjunction with the nitrogen leaving in the dewatered sludge cake, accounts for approximately a 55 % reduction in the nitrogen load.

As discussed in Section 2.2.2, when considering various PN/A reactor types, granular PN/A reactors treat the highest nitrogen loads with an average load of approximately 3 250 kg/d. The calculated nitrogen load in the side stream liquor at Cape Flats WWTP is 3 415 kgN/d. This indicates that the nitrogen load at Cape Flats WWTP is above the average nitrogen loads treated by PN/A granular reactors. However, as Lackner *et al.* (2014) points out, at the time of their publication, the average nitrogen load being treated per plant from municipal sludge was approximately 800 kg N/d where higher average nitrogen loads are treated in industrial wastewater (approximately 2 700 kg N/d). This puts the significance of the scale of the nitrogen load into perspective at Cape Flats WWTP.

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[1] The TKN design treatment capacity of Cape Flats is assumed to be 11 712 kg N/d (NINAM SHAND (South) (Pty) Ltd and AFRICON Engineering International (Pty) Ltd, 1999).

Table 4.3: Mass balance stream table indicating the results of the FSA, OP, VSS and TSS mass balance conducted over the AD facility.

Stream	Symbol	Unit	1	2	3	4	5	6	7
Description	**-**	**-**	**AD Feed**	**AD Effluent**	**Struvite Reactor Effluent**	**Polymer solution**	**Dewatering feed**	**Dewatered Cake**	**Side stream Liquor**
Total solids flux	FX_T	tDS/d	143	85.8	104.2	0	104.2	103.1	1.042
WAS solids flux	$FX_{T,WAS}$	tDS/d	85.8	51.5	51.5	0.0	51.5	51.0	0.515
PS solids flux	$FX_{T,PS}$	tDS/d	57.2	34.3	34.3	0.0	34.3	34.0	0.343
Volatile Solids Flux	FXv	tDS/d	104.9	47.7	47.7	0.0	47.7	47.3	0.477
WAS volatile solids flux	$FX_{V,WAS}$	tDS/d	58.6	26.7	26.7	0.0	26.7	26.4	0.267
PS volatile solids flux	$FX_{V,PS}$	tDS/d	46.3	21.1	21.1	0.0	21.1	20.9	0.211
Inorganic solids flux	FX_{IO}	tDS/d	38.1	38.1	56.4	0.0	56.4	55.9	0.564
Total solids Concentration	X_T	g/l	105	64.0	77.7	0	43.7	300	0.511
Total VSS concentration	X_V	g/l	77.0	35.6	35.6	0.0	20.0	137.5	0.234
Total Volumetric Flow	Q	Ml/d	1.36	1.34	1.34	1.04	2.38	0.34	2.04
FSA flux	FN_a	t/d	-	4.28	3.23	0	3.23	0.47	2.762
FSA concentration	N_a	g/l	-	3.19	2.41	0	1.35	1.35	1.355
OP flux	FP_s	t/d	-	2.59	0.26	0	0.26	0.037	0.222
OP concentration	P_s	g/l	-	1.93	0.193	0	0.109	0.109	0.109
Struvite Flux	FX_{STR}	t/d	-	-	18.4	0	18.4	18.2	0.184

4.2 PN/A and ND process design comparison

This section compares the critical design aspects of the conceptual plant designs for the PN/A and MLE process treating the SRL at Cape Flats WWTP.

4.2.1 Process Operating Conditions

Table 4.4 below indicates the operating conditions for the PN/A and MLE processes.

Table 4.4: A summary of the operating conditions of the PN/A and MLE processes.

	PN/A Process	**MLE**	**Equation/Source**
Sidestream volumetric flow (Ml/d)	2.04	2.04	Table 4.2
COD load (kg/d)	9 180	15 886	2.7a
TKN load (kg N/d)	3 415	3 415	$Q_i \cdot N_{Ti}$
FSA load (kg N/d)	2 764	2 764	Table 4.2
% FSA removal	89.9	99.9	3.22g
% TIN removal	80.2	90.4	3.22f
MLSS (kg/m^3)	3.7	4.0^2	1 – 3.13c 2 - specified
Nitrogen loading rate (kgN/m^3/d)	0.5^3	1.1	3 - specified 4 - $\frac{Qi \cdot Nti}{Vp}$
Sludge age (days)	3.4^6	4^7	7- specified

[6] This sludge age refers to the flocculant OHO sludge age which is assumed to be approximately equal to the hydraulic retention time ($\frac{Vp}{Qi}$) due to continuous purging and wasting of all OHO sludge.

[7] The model assumes that the autotrophic granular sludge age is infinitely long as the biomass is retained in the reactor indefinitely. In reality, small amounts of autotrophic biomass will slough off the granules and be wasted with the OHO sludge albeit at a negligible rate compared to the OHO sludge.

Both processes treat the full side stream nitrogen load therefore the TKN and FSA loads are equal. The COD load being treated by the MLE process is higher due to the dosing of methanol as an organic source to ensure that the TKN/COD ratio is sufficient such that denitrification can occur. The FSA and TIN conversion are lower for the PN/A process than the MLE process. The percentage FSA and TIN removal for the PN/A process have been set based on what is typically achievable for full scale installations (Van Hulle *et al.*, 2010). It is possible to achieve an even higher TIN conversion in the MLE process but this would require a larger amount of organic dosing which would increase the operational cost. Therefore, for comparative reasons, the MLE model was set up such that the nitrate concentration in the effluent was similar to the PN/A process. The MLE process achieves a high FSA removal as the sludge age in the model was set such that near complete nitrification is achieved. Operating the MLE process as a high rate system at 24°C greatly improves the nitrification kinetics. Consequently, high FSA removal is achievable at short sludge ages relative to ambient MLE processes.

4.2.2 Major equipment sizes

Table 4.5 provides a comparison of the major equipment sizes for the PN/A and MLE process options.

Table 4.5: A comparison of the major equipment sizes between a PN/A process and a conventional MLE process treating side stream liquor.

Major Civil Structures	**PN/A**	**MLE**	**Equation/Source**
Number of Reactors	2	1	input
Anoxic Zone (m^3)	n/a	1583	$Vp \cdot fx$
Aerobic Zone (m^3)	n/a	1648	$Vp \cdot (1 - fx)$
Reactor Size (m^3)	3415[a]	3231[b]	a – 3.13a b – 3.33
Number of Settling Tanks	2	1	input
Settling Tank Area (m^2)	38.5	79	3.23-3.27
Settling Tank Diameter (m)	7.0	10	$\sqrt{\frac{4 \cdot A_{ST}}{\pi}}$

The MLE process has a smaller reactor volume, when compared to the PN/A process. The MLE process uses methanol as an organic source which results in a significantly lower growth yield of biomass (0.33 mgVSS/mgCOD) when compared to activated sludge growth using organic sources found in municipal wastewater i.e. VFA's (0.45 mgVSS/mgCOD). Methanol being 100% readily biodegradable COD (RBCOD), significantly increases the denitrification rate in the anoxic zone which results in a relatively small anoxic volume. In addition to this, dosing methanol does not add any inert material or ISS which would accumulate in the reactor resulting in a larger sludge mass and subsequently, a larger reactor volume. The MLE process is not subjected to the same inhibitory factors as the PN/A process and is therefore not limited by nitrogen loading rates which results in a smaller overall reactor volume. The assumption that the anoxic zone remains completely anoxic (limitation of the steady state model as discussed in Section 3.2.2.2) simulates a case whereby the growth of PAO's is negligible. Due to the high VFA concentration in the side stream liquor, should any areas in the anoxic zone become anaerobic, the growth of PAOs could become significant which would increase the sludge mass, and therefore the reactor volume, in order to achieve the design MLSS of 4 kg/m^3. This is due to PAOs having a lower endogenous respiration rate when compared to OHOs (Wentzel and Ekama, 1997; Ekama, 2011b).

The size of the PN/A reactor volume is based on a nitrogen loading rate of 0.5 kg N/m^3/d. A conservative loading rate (producing a worst case scenario for reactor sizing) was selected for this study based on the existing full scale plants (Lackner *et al.*, 2014) and recommendations in literature to prevent the inhibitory effects associated with treating side stream liquor from THP (Figdore *et al.*, 2011).

The size of settling tanks are dependent on hydraulic loading rates, sludge settleability and concentration. The overall required settling tank surface area for the PN/A system and MLE system is approximately equal due to the hydraulic loads and reactor mixed liquor concentrations being similar as indicated in Table 4.4. For comparative purposes, the design for both processes were based on conventional raked settling tanks as this provides a worst case scenario in terms of space and capital cost. A further reduction in space and capital cost could be achieved by using lamella type settling tanks. These are already widely used in existing commercial installations whereby the lamella settling tank is sometimes installed within the reactor.

The MLE process has a significantly larger aeration requirement which is discussed in more detail in Section 4.2.3. Larger aeration requirement results in larger blower equipment which requires a larger blower house to enclose the equipment.

4.2.3 Electrical Energy requirements

This section focuses on the energy requirements of each design. Energy consumption has been split into:

- Pumping energy
- Mixing energy
- Aeration energy
- Heating energy

The energy requirements for pumping and mixing are listed in Table 4.6.

Table 4.6: A comparison of the major pumping and mixing energy requirements of an MLE and PN/A process.

Pumps		**PN/A**	**MLE**	**Equation/source**
Reactor Feed Pump (kW)	Capacity (m^3/h)	85	85	$\frac{Qi \cdot 1000}{24}$
	Power (kW)	4.6	4.7	3.22
Sludge recycle pump (kW)	Capacity (m^3/h)	85	85	$\frac{Qi \cdot S_{rec} \cdot 1000}{24}$
	Power (kW)	4.6	4.6	3.22
Anoxic Zone A Recycle Pump (kW)	Capacity (m^3/h)	N/A	656	$\frac{Qi \cdot A_{rec} \cdot 1000}{24}$
	Power (kW)	N/A	3.6	3.22
Mixing	Power (kW)	8.5[a]	7.9[b]	a - $Vp \cdot 5 \cdot 0.25/1000$ b - $Vp \cdot fx1 \cdot 5/1000$

It is imperative to note that the PN/A process only requires one internal sludge recycle per reactor whereas the MLE process requires a sludge recycle and a larger anoxic zone “A” recycle. The A recycle is an important control parameter in the MLE process which returns nitrates back to the anoxic zone to ensure denitrification takes place. The MLE process therefore results in a higher pumping energy than the PN/A process. An A recycle ratio of 7.7 was calculated for the MLE reactor operating conditions. It should be noted that a maximum A recycle ratio of 6 is generally accepted as the highest practical limit for mainstream wastewater treatment processes whereby above this limit typically makes the energy consumption to pump the wastewater prohibitively high (Ekama, 2011b). However, due to the lower influent hydraulic load of the side stream liquor compared to mainstream processes; this limit is not necessarily applicable as a ratio of 7.7 yields a 3.6 kW pump which is deemed acceptable.

Mixing energy is proportional to the volume being mixed which was calculated on the basis of 5 W/m^3 (Gernaey *et al.*, 2004). An MLE configuration has mechanical mixers in the anoxic zone to keep the sludge in suspension whereas the aerobic zone relies on aeration for mixing. For the PN/A configuration there are no distinct anoxic zones and mixing is performed by a combination of aeration and mechanical mixing. Therefore, in times when the oxygen demand is low, aeration alone is not sufficient for mixing, and subsequently, the mechanical mixers run. The mixing energy for the PN/A process is expected to be similar to MLE with the assumption that mixers are only running for 25% of the time when the aeration rate is too low to maintain sufficient agitation.

Table 4.7: A comparison of the aeration requirements of a PN/A process versus an MLE process for treating side stream liquor.

Parameter	**PN/A**	**MLE**	**Equation/source**
Actual oxygen requirement, AOR (kg/h)	304	534	2.7k, 2.8n, 2.8o & 2.8p
Water temperature (°C)	35	24	input
AOR/SOR correction factor [1]	0.669	0.524	3.14 & 3.15
Standard oxygen requirement, SOR (kg/h)	454	1019	2.7k, 2.8n, 2.8o, 2.8p, 3.14 & 3.15
Aeration energy requirement, E_{shaft} (kW)[2]	91	204	$SOR \cdot 5$
Total aeration energy, E_{total}(kW)[3]	130	291	$\frac{E_{shaft}}{0.7}$

[1] An alpha, beta and fouling factor 0.6, 0.95, and 0.9 respectively was used to calculate the AOR/SOR conversion factor (Tchobanoglous *et al.*, 2003)

[2] The FBDA equipment efficiency was assumed to be 5 kg O_2/kWh (Redmon, 2018)

[3] A shaft to wire efficiency of 70% was used (Ekama, 2011a).

It is shown in Table 4.7 that the AOR for the PN/A process is 43% less than the AOR for the MLE process due to short circuiting of the partial nitritation process. It is important to note that carbonaceous oxygen demand of the OHOs accounts for 36% of the AOR of the PN/A process due to the presence of biodegradable COD contained in the side stream liquor feed, with the balance being used by the AOBs for partial nitritation. The amount of biodegradable COD in the side stream liquor significantly affects the efficiency of the PN/A process. However, this amount of biodegradable COD in the side stream liquor is dependent on AD operation and can represent a significant portion of the total AOR (Zhang *et al.*, 2019). This needs to be carefully considered at design stage to ensure that the aeration equipment is appropriately sized.

It is important to note that for both the MLE and PN/A processes, the mixing and pumping energy represents a very small fraction of the total energy requirements with aeration energy making up the bulk of the energy split.

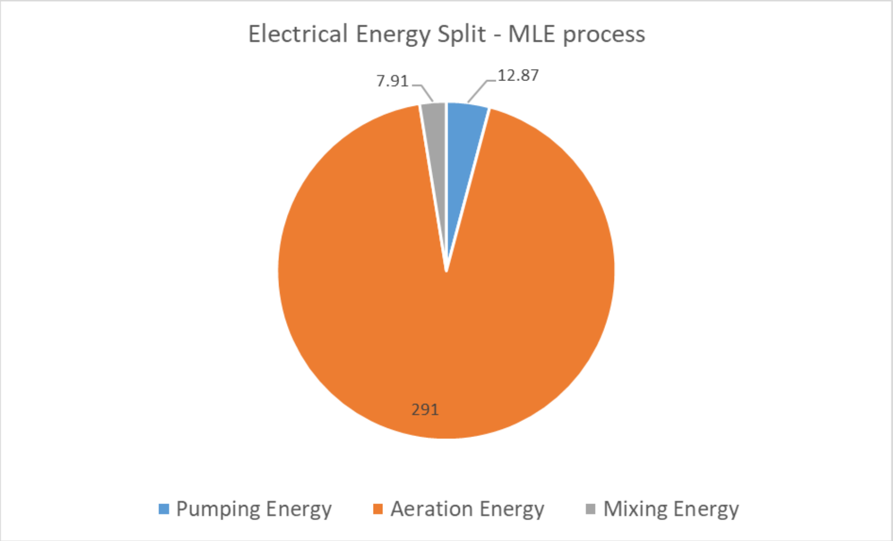

Figure 4.1: Electrical power consumption split -MLE

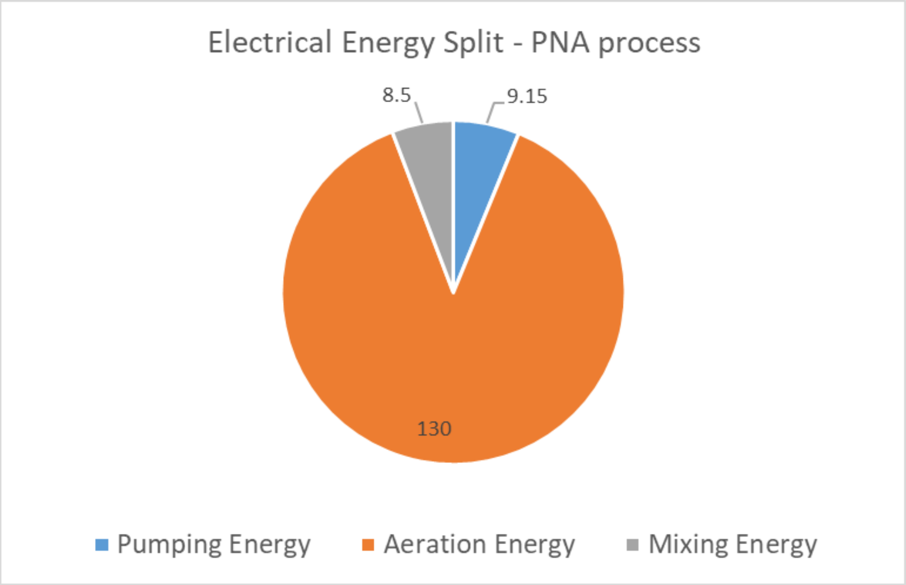

Figure 4.2: Electrical power consumption split - PN/A

Figure 4.1 and Figure 4.2 show that for both design cases, aeration consumes the greatest percentage of the total energy consumption. The selection and efficiency of the aeration equipment is therefore a critical aspect for any design. The specific energy consumption was calculated to be 1.48 kWh/kg N for the PN/A process compared to 2.85 kWh/kg N for the MLE process. The specific energy consumption for the PN/A process is in line with values achieved in existing commercial applications whereas the MLE process is lower than 4 kWh/kg N presented by Lackner *et al.* (2014). A possible reason for this is that the MLE system is modelled using high efficiency FBDA equipment whereas the process in literature could be using less efficient aeration equipment. The modelling exercise indicates that the PN/A process uses 36% less energy to achieve the same nitrogen removal. In Section 2.2.2, an aeration savings of 60% is mentioned. This saving is based on having a purely autotrophic process whereby there is no additional biodegradable COD present in the side stream liquor to promote OHO growth in the PN/A reactor. For a commercial AAD process, this is unrealistic to expect. Therefore, the assumptions and results of this thesis indicate a more realistic aeration savings. However, should there be less biodegradable COD in the side stream liquor than what is assumed in Section 3.2.1.1, the energy savings will greater.

4.2.4 Heating Energy requirements

Both the MLE and PN/A processes occur above general ambient temperatures, and therefore, heat needs to be added to the process under certain conditions. This section discusses the results of the energy balances conducted over each of the proposed processes to determine the heating requirements. Figure 4.3 below illustrates the heating requirements of the PN/A and MLE process. The heating requirements are split into the energy required to heat the side stream liquor to the reactor operating temperature i.e. sensible heat and the energy required to account for heat losses to the environment.

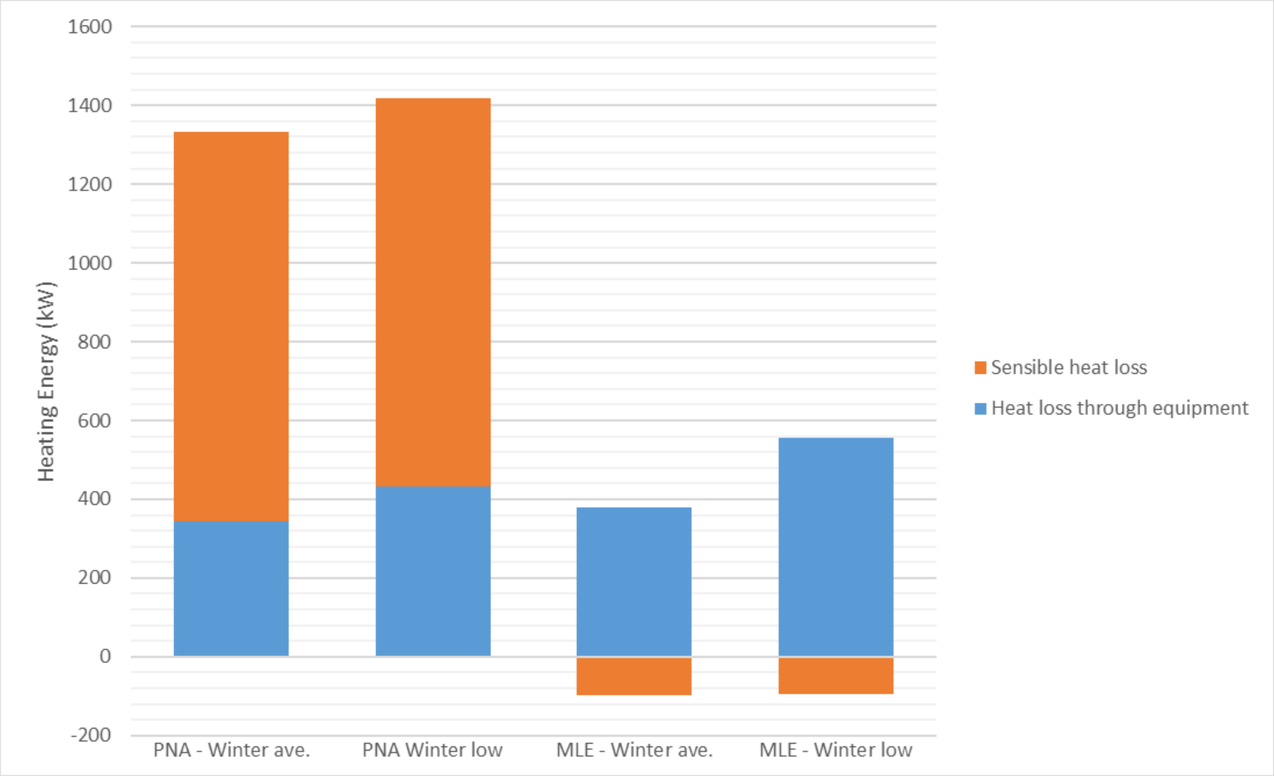

Figure 4.3: A comparison of the heating duties required by the PN/A and MLE process

Figure 4.3 shows that the PN/A process overall requires more heating energy than the MLE process. Due to the PN/A process being operated at 35°C, a substantial portion of the heating requirement goes towards heating the incoming sidestream liquor. Due to the MLE process operating at 24°C, there is a negative value which indicates that some of the sensible heat (refer to Section 3.2.1.7) in the side stream liquor can be used to overcome heat losses in the MLE system. The PN/A system has a lower heat loss to the environment compared to the MLE system despite operating at a higher temperature and having larger equipment and therefore more surface area for heat loss. This is because the PN/A system was modelled as an enclosed reactor while the MLE system is open to the surrounding environments. This makes a significant difference to the heat balance as energy is lost through evaporation especially at Cape Flats WWTP where wind speeds are regularly above 50 km/h. Enclosing the reactor creates an additional capital cost and due to the low heating requirements of the MLE system, this was not deemed necessary. The PN/A system on the other hand, requires significant heating, and therefore, it was opted to incur the additional capital cost to enclose the reactor.

The objective of conducting the energy balance is to verify and optimize the designs to ensure that the heating energy required does not exceed the heating energy available. As highlighted in Section 3.2.1.7, low grade energy will be supplied from the CHP engines at the digestion facility in the form of warm water at approximately 85°C. The calculated maximum low grade heat available is shown in Table 4.8 below:

Table 4.8: Low grade heat available in the form of hot water produced by the CHP engines at the digestion facility.

	Amount	**Equation/Source**
Low grade heat available, Q_{LGH} (kW)	3 040	3.21a, 3.21b & 3.21c

The surplus low-grade heat available from the CHP engines at digestion facilities such as the one proposed for Cape Flats WWTP is suited for side stream technologies requiring heating as there are typically no other major low grade heat users in the digestion process and mainstream works. One can see from the energy balance that for this particular application, there is no additional cost associated with heating for either process as the total heat demand is substantially less. It should be noted that this is case specific and should be assessed based on each specific design and climatic region with the intent of balancing the heating requirements with the available heat from the CHP engines.

4.2.5 Chemical Dosing and Alkalinity

This section focuses on chemical consumption rates for each process. For the PN/A process, it was calculated that if the side stream liquor has an incoming alkalinity of >5100 mg/l as $CaCO_3$, the predicted effluent alkalinity will be >50 mg/l as $CaCO_3$ and therefore no additional chemical alkalinity dosing is required. With an assumed side stream liquor alkalinity (entering each respective process) of 5500 mg/l as $CaCO_3$ (Zhang *et al.*, 2016; Gu *et al.*, 2018), it is unlikely that any additional alkalinity dosing is required under normal operation which presents a significant operational saving for the PN/A process when compared to MLE.

MLE requires a significant amount of methanol to achieve the required level of denitrification in the design. Dosing methanol has the added benefits of improving the denitrification rate resulting in a smaller reactor footprint. However this comes with a large chemical consumption cost. This is discussed in more detail in Section 4.4. The minimum alkalinity in the feed to the MLE process must be > 5325 mg/l as $CaCO_3$ to ensure that the effluent alkalinity is > 50 mg/l as $CaCO_3$. Should the influent alkalinity drop below the assumed range, additional denitrification could be promoted to recover alkalinity. This however would require additional methanol dosing and an increase in the A recycle ratio which would come at an additional chemical and power cost.

4.2.6 Effluent quality

In this section, the effluent quality leaving the PN/A and MLE process is presented and discussed. Table 4.9 and Table 4.10 indicate the predicted effluent qualities of the MLE and PN/A processes.

Table 4.9: Predicted Effluent concentrations leaving the side stream treatment processes.

Effluent Concentrations	**PN/A**	**MLE**	**Equation/ Source**
COD (mg/l)	3220	4023	3.22a
TKN (mg N/l)	499	418	3.22b
Ammonia (mg NH4-N)/l	136	1.1	A – 3.22c B – 2.8b
Nitrite (mg NO2-N/l)	0	0	2.5 (nitrite limiting)
Nitrate (mg NO3-N/l)	132	129	C – 2.6 D – 3.36
Phosphorus (mgP/l)	186	185	P_{ti}
Orthophosphate (mg OP-P/l)	98	82	3.22c
TSS (mg/l)	964	1579	3.22d
ISS (mg/l)	331	400	3.22e

Table 4.10: Predicted effluent loads returned to the mainstream wastewater works

Effluent Loads (kg/d)	PN/A	MLE	Equation/ Source
Flow rate, Q_i (m^3/h)	83	83	Input
COD	6569	8229	$Qi \cdot S_{te}$
TKN	1019	855	$Qi \cdot N_{te}$
Ammonia	278	2.2	$Qi \cdot N_{ae}$
Nitrate	269	264	$Qi \cdot N_{ne}$
Phosphorus	379	379	$Qi \cdot P_{te}$
Orthophosphate	425	167	$Qi \cdot P_{se}$
TSS	2054	3231	$Qi \cdot X_{te}$
ISS	391	819	$Qi \cdot X_{IOe}$

The treated effluent from the side stream treatment process is made up of the overflow from the settling tank combined with the wasted sludge. The sludge ages for both processes are sufficiently long enough (>2.5 days) to assume that all the biodegradable COD in the influent is consumed. The COD in the effluent is therefore made up of the unbiodegradable soluble and particulate components together with the biodegradable and unbiodegradable particulate portion of the biomass in the wasted sludge. The COD of the MLE process is higher due to a higher sludge mass being wasted which contributes to the effluent COD. It should be noted that the soluble unbiodegradable COD portion in the effluent is 2283 mgCOD/l and 2277 mgCOD/l for PN/A and MLE processes respectively. A high soluble unbiodegradable COD fraction is produced by THP processes (Figdore *et al.*, 2011) which will end up being recycled back to the mainstream works irrespective of the sidestream treatment process. It is estimated that this could increase the COD in the effluent at Cape Flats WWTP by 23-46 mg/l[2].

The conversion of ammonia in the MLE processes is 99.9% which is higher than the ammonia conversion of 89.9% in the PN/A process. For the PN/A process, as per the stoichiometry, approximately half the incoming ammonia is oxidized to nitrite however the anammox reaction requires 1.32 moles of nitrite for every 1 mole of ammonia. Subsequently, the nitrite is the limiting reagent in the anammox reaction resulting in residual ammonia in the effluent. Although it is theoretically possible to achieve higher than 90% conversion of ammonia, practical experience from single stage PN/A processes presented in literature, limit the conversion at 90% (Van Hulle *et al.*, 2010).

Both processes will have nitrates present in the effluent. The production of nitrates in the anammox reaction is unavoidable. As mentioned in Section 4.2.1, the MLE process can achieve a lower

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[2] The increase in soluble unbiodegradable COD concentration is estimated based on diluting 2.04 Ml/d of treated sidestream liquor recycled into 100 to 200 Ml/d of raw wastewater.

effluent nitrate concentration at the cost of higher organic dosing but for comparative reasons, the process has been modelled such that it achieves approximately the same nitrate removal as the PN/A process. For both processes the nitrate concentration to the mainstream wastewater works is expected to increase by 1-2.6 mg NO_3-N/l[3]. This would introduce nitrates into the anaerobic zone, creating a slightly anoxic environment which would negatively affect EBPR (Ekama, 2011b) therefore returning the side stream liquor straight into the primary anoxic zones should be considered.

The total phosphorus loads returned to the mainstream works are identical for both the MLE and PN/A processes. The OP loads are different with the MLE process returning a lower OP load due to a larger P uptake as a result of higher OHO growth in the MLE reactor compared to the PN/A process. Both processes will not reduce the phosphorus being returned to Cape Flats WWTP as these processes are focused on nitrogen removal only. The amount of phosphorus being returned to the mainstream plant is based on the performance of the struvite process upstream of the MLE/PN/A process.

PN/A produces the lowest TSS due to the low sludge generation rates. The TSS is almost exclusively due to OHO generation from the biodegradable COD fraction in the influent. The MLE process produces 64% more sludge than the PN/A process which is returned back to the mainstream works.

The EQI was calculated for each process according to Equation 3.38 and 3..39 to compare the respective pollutant loads against one another. Although the EQI is commonly used to compare the effluent qualities leaving the mainstream works and entering a water body, any pollution load leaving the side stream treatment will add an additional burden onto the existing mainstream works which could risk the final effluent quality leaving the plant. In this regard, it is important to not only compare the economics of each process but to also assess if there is a significant difference in process performance which could relate to potential environmental risks. The values obtained are displayed in Table 4.11.

Table 4.11: EQI comparison of the PN/A and MLE processes

EQI	**(kg pollutants/d)**	**Equation/Source**
PN/A	22 260	3.38 and 3.39
MLE	22 036	3.38 and 3.39

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[3] The increase in Nitrate concentration to the biological reactors is estimated based on diluting 2.04 Ml/d of treated sidestream liquor recycled into 100 to 200 Ml/d of raw wastewater.

Table 4.11 shows that the EQI of each process is very similar with the MLE process being 1% less. However, the contribution of the individual pollutants to the EQI is different due to the difference in effluent quality. Figure 4.4 and Figure 4.5 indicate the relative contributions of each pollutant to the overall EQI of the respective process. One can see that COD is by far the largest contributor with the MLE process contributing the largest due to COD present as sludge generated in the process. The COD has been split into soluble and particulate to show the significant contribution of the unbiodegradable soluble COD to the pollutant load which is present irrespective of the side stream treatment process and is a result of THP pretreatment.

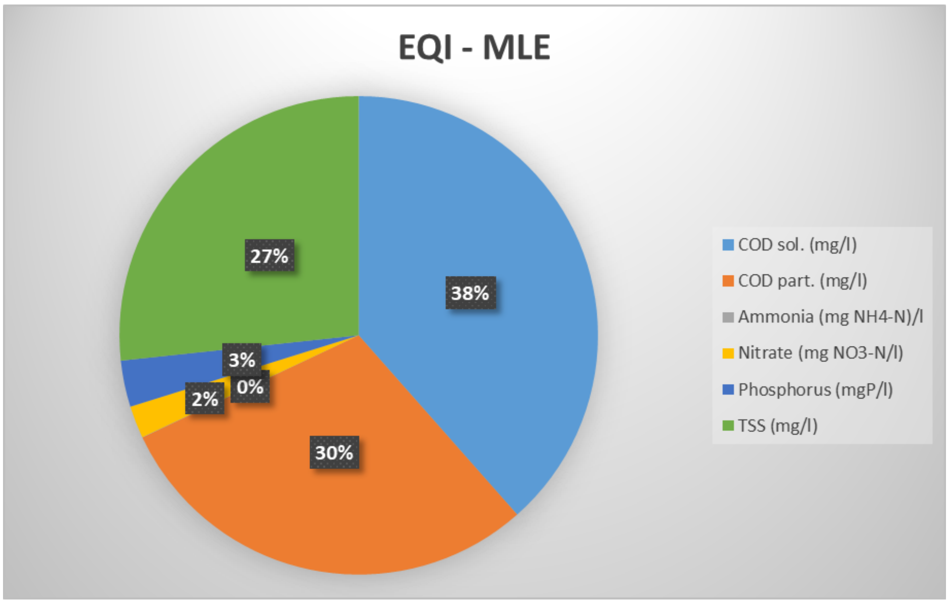

Figure 4.4: The relative pollutant contribution to the EQI for the MLE process.

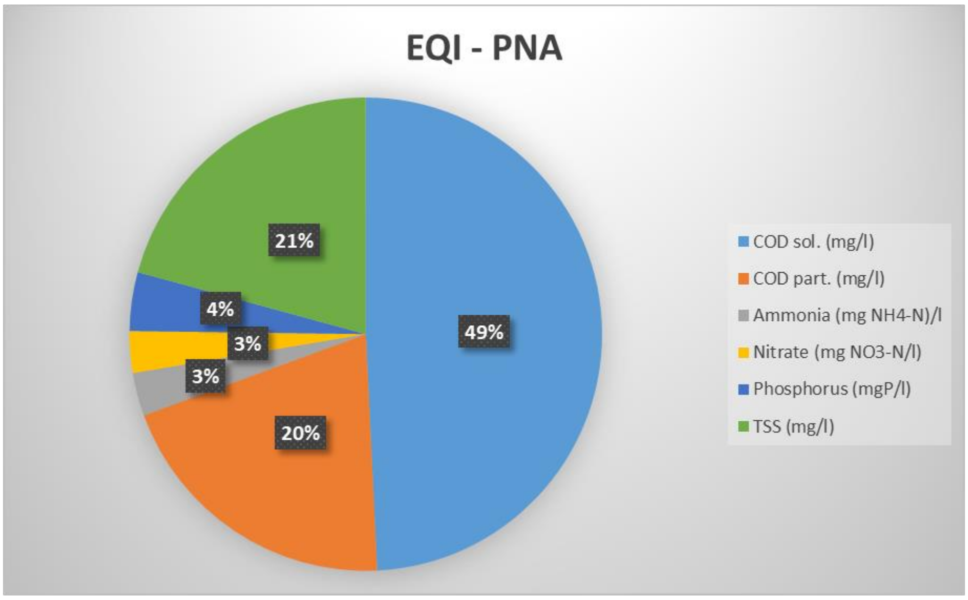

Figure 4.5: The relative pollutant contribution to the EQI for the PN/A process

Due to the higher ammonia removal in the MLE process, the relative contribution of ammonia to the EQI is negligible however the contribution is significant for the PN/A process. Overall, the higher ammonia levels in the PN/A process is largely offset by the higher COD present in the MLE process resulting in the MLE process only having a 1% lower EQI than the PN/A process. The contribution of nitrates is nominally equal. As discussed previously, this could be further reduced on the MLE process but would come at a higher economic cost through additional organic dosing. The contribution of TSS from the MLE process is higher due to higher sludge mass being wasted when compared to the PN/A process.

It should be noted that the EQI quantifies the pollutant loads from each of the processes by weighting the individual pollutant load with the pollutant limit in the mainstream works. This assumes that the effluent is discharged directly to a water body. In this study, the effluent is recycled to the mainstream works which presents some limitations to the use of the EQI. This approach does not assess the ability of the mainstream works to deal with each pollutant load and assumes that the difficulty of processing the pollutants at the mainstream works is proportional to the effluent limits. The ability of the mainstream works to handle the individual pollutants may not be proportional in this regard as it would depend on the capacity and condition of the existing infrastructure. The effect of the effluent quality in the side stream liquor on the mainstream works would need to be modelled to determine the exact effect on final effluent quality which could then be further assessed with another EQI. This presents an opportunity for future research work whereby the treated side stream liquor returned to the biological reactors is modelled to determine the exact effect on the effluent given the characteristics of the raw effluent being sent to the mainstream works.

In this study however, the incremental increase in aeration and sludge generated at the mainstream works as a result of processing the side stream liquor effluent was calculated (as outlined in Section 3.2.1.10 and is given in Table 4.12

Table 4.12: The incremental increase in aeration and sludge generated at the mainstream treatment process as a result of recycling the effluent of the side stream processes.

	PN/A	**MLE**	**Equation/Source**
Aeration (kW)	40.7	49.8	2.7k, 2.8n, 2.8o & 2.8p 3.14 & 3.15
Sludge generated (tDS/d)	1.43	2.00	2.7a-j & 2.7l

Although the MLE process has a slightly lower EQI, the burden on the mainstream wastewater works is higher with the MLE effluent requiring 23% more aeration and creating 40% more sludge. The incremental increase in aeration is due to the ammonia and biodegradable COD (in the form of wasted active sludge mass) in the side stream effluent. Even though the PN/A process has a larger ammonia concentration in the effluent than the MLE process, the MLE's biodegradable COD portion is higher than the PN/A process therefore requiring more aeration energy. The higher TSS, ISS and biodegradable COD of the MLE side stream effluent results in greater sludge production at the mainstream works. Any additional sludge generated at the mainstream works takes away capacity from the digestion facility which diverts import sludge away and results in an increased sludge disposal cost for the City of Cape Town.

4.3 Capital Cost Analysis

This section presents and discusses the results of the capital costing exercise (as described in Section 3.4) for each of the treatment options. Figure 4.6 illustrates the total capital cost estimate for each process together with the major breakdown of cost into its major components.

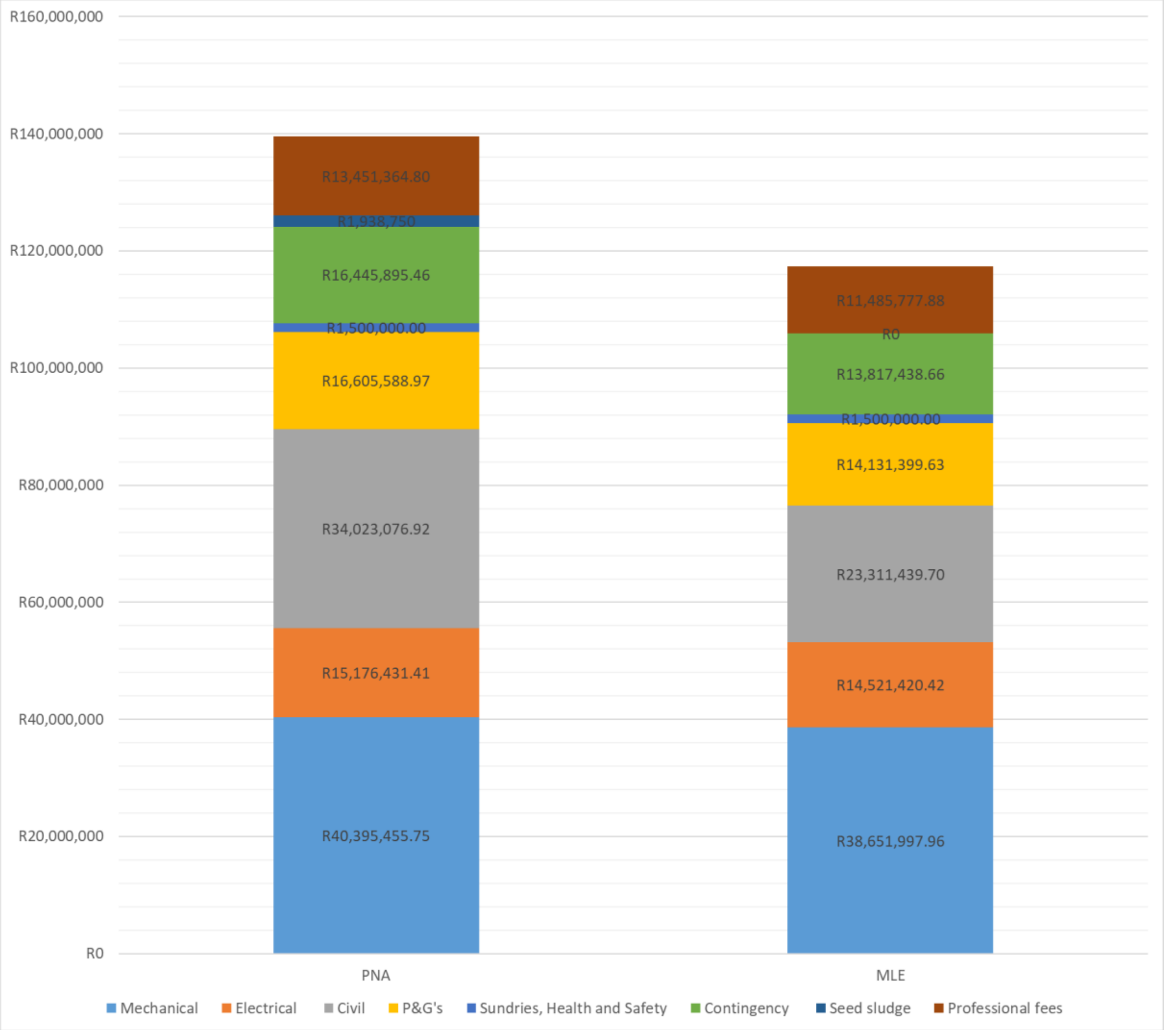

Figure 4.6: Capital cost comparison of the PN/A and MLE process

Figure 4.6 shows that the MLE process costs 16% less to implement compared to the PN/A process. When one looks at the relative component contributions to the total cost one can see that the mechanical equipment is 4% less expensive for the MLE process than the PN/A process. This is mostly due to the additional cost of heating utilities for the PN/A process which outweighs the larger aeration equipment for the MLE process. Similarly, the electrical cost of the MLE process is 4% less expensive for the MLE process compared to the PN/A .

The civil costs for the MLE process are 31% less than the PN/A process with most of this additional cost due to larger reactor volumes. Because of the inhibitory effects to the PN/A process, the process must run at much lower nitrogen loading rates when compared to the MLE process. This results in larger reactors. In addition to that, as discussed in Section 4.2.4, in order to conserve heat energy, the PN/A process requires fully enclosed reactors which significantly increases the cost.

Seed sludge is a cost that is unique to the PN/A process and represents 1.4% of the total project cost. In order to obtain reasonable start-up times i.e. within 6 months, large volumes or seed sludge are required. Seed sludge is discussed further in Section 4.6.1.

4.4 Operational Cost Analysis

In this section each operational cost element for the PN/A and MLE process is calculated and presented as an operational cost index. Table 4.13 compares the various components and the total OCI for each process.

Table 4.13: An Operational Cost Index comparison between a PN/A and MLE process treating sidestream liquor.

		PN/A	**MLE**	**Equation/Source**
AE (kW)[1]	Sidestream	130	291	Table 4.7
	Mainstream	40.7	49.8	
PE (kW)		9.3	12.9	Table 4.6
ME (kW)		8.5	7.9	Table 4.6
HE (kW)[2]		1332	281	3.17-3.20
SP (kg TSS/d)[3]		1431	2005	2.7e-2.7j & 2.7l
Chemical consumption (m^3/d)		-	5.7	3.32b, 3.32c & 3.32d
Energy Cost[4]		R1,765,155	R3,834,320	3.42
Sludge disposal Cost[5]		R424,146	R594,382	
Chemical Consumption Cost[6]		N/A	R17,339,917	
OCI		**R2,189,301**	**R21,768,619**	

[1] This includes the power consumption of the sidestream treatment process and the additional power consumption required to treat the effluent from the respective sidestream process at the mainstream WWTP.

[2] At these heating duties there is a surplus of heating energy from the CHP engines as discussed in section 4.2.4. Heating energy does not contribute the OCI for both processes.

[3] This is the incremental increase in waste activated generated at Cape Flats WWTP mainstream works due to the recycling of the sidestream liquor effluents from the PN/A and MLE process.

[4] Assumes an energy cost of R1.21 R/kWh for 2020.

[5] Assumes a sludge disposal cost of R689/kg TSS

[6] Assumes a chemical cost of R10.60/kg for methanol (Brenntag, Email quotation, 27 Jan 2021).
sidestream

Table 4.13 shows that the OCI is substantially lower for a PN/A process compared to an MLE process. Figure 4.7 indicates the breakdown of the OCI for each process. The chemical consumption costs, in the form of methanol as an organic source, makes up most (80%) of the OCI cost for the MLE process. The high concentration of biodegradable COD in the sidestream liquor helps to reduce the organic dosing requirement whereby the methanol accounts for 42.2% of the total COD to the MLE process. The relative chemical dosing cost would be significantly higher for conventional AD processes (without THP) where typically the amount of biodegradable COD in the sidestream liquor is substantially lower (Zhang *et al.*, 2016). For the PN/A process, with the high alkalinity level of THP digestate, no additional chemical dosing is required under normal operating conditions which makes electrical energy costs the biggest contribution to the OCI. The sludge disposal cost for the MLE process is 40% higher than the PN/A process due to the additional sludge generated as discussed in section 4.2.6. which takes away sludge importing capacity whereby the City of Cape Town would incur the cost of sending that sludge to landfill.

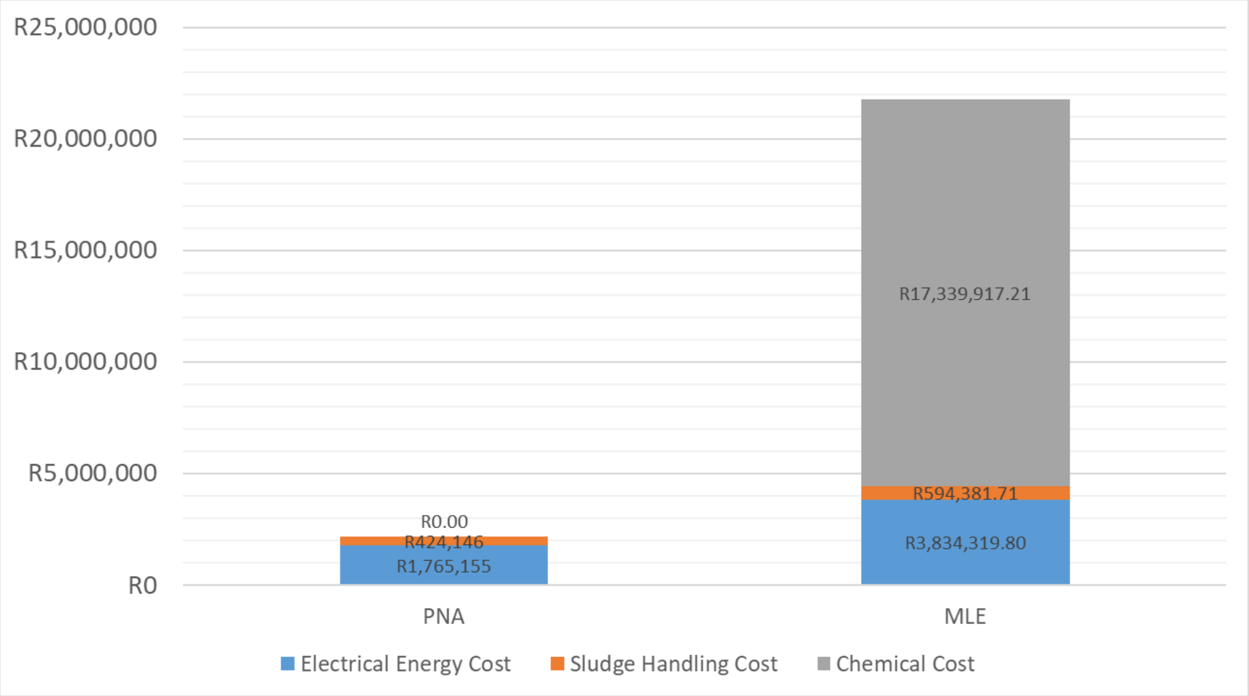

Figure 4.7: A comparison of the OCI for the PN/A and MLE process.

4.5 Life-Cycle Cost Analysis

This section combines the results of the OCI and capital costing sections and uses these results to formulate an overall LCCA for each of the processes. Refer to Section 3.6 which describes the methodology used to complete the LCCA.

Table 4.14: A comparison of the total lifecycle cost of the PN/A and MLE process in present day value over a 20 year operational period

	PN/A	MLE	Equation/Source
LCCA over 20-year operational period[1], PV	R195,862,286	R553,239,250	3.43

1- Assumes a 3 year construction period

Table 4.14 indicates that implementing an MLE process will cost 2.8 times the cost of the PN/A process over a 20 year operational period. Figure 4.8 below shows the capital and operational split of the total cost represented as Present Value (PV) for each process. One can see that for an MLE process, operational costs represent 79% of the total life-cycle cost as opposed to the PN/A process which represents 30%. This indicates that a much larger proportion of the total costs are operational for the MLE which are incurred continuously once the process goes online. The PN/A process, incurring higher capital costs upfront, only incurs a fraction of the operational costs of the MLE process making it far less costly over the long term.

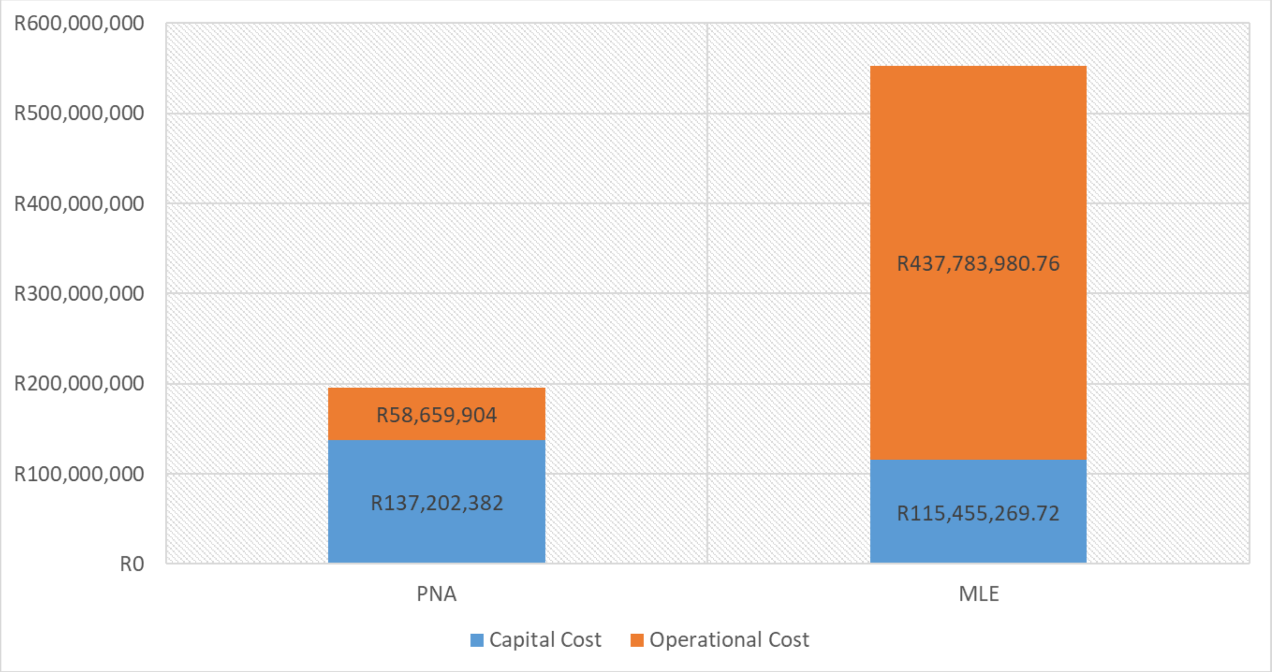

Figure 4.8: A comparison of the capital and operational breakdown of the PV cost for the PN/A and MLE process over an operation life of 20 years.

Figure 4.9 indicates the cumulative PV life-cycle cost as a function of years for each of the processes. One can see that the initial capital cost incurred with the installation of the PN/A process is quickly overcome by the operational costs of the MLE process such that by the end of the 1st year of operation, the total life-cycle costs are approximately equal. From that point onwards, the MLE is more expensive.

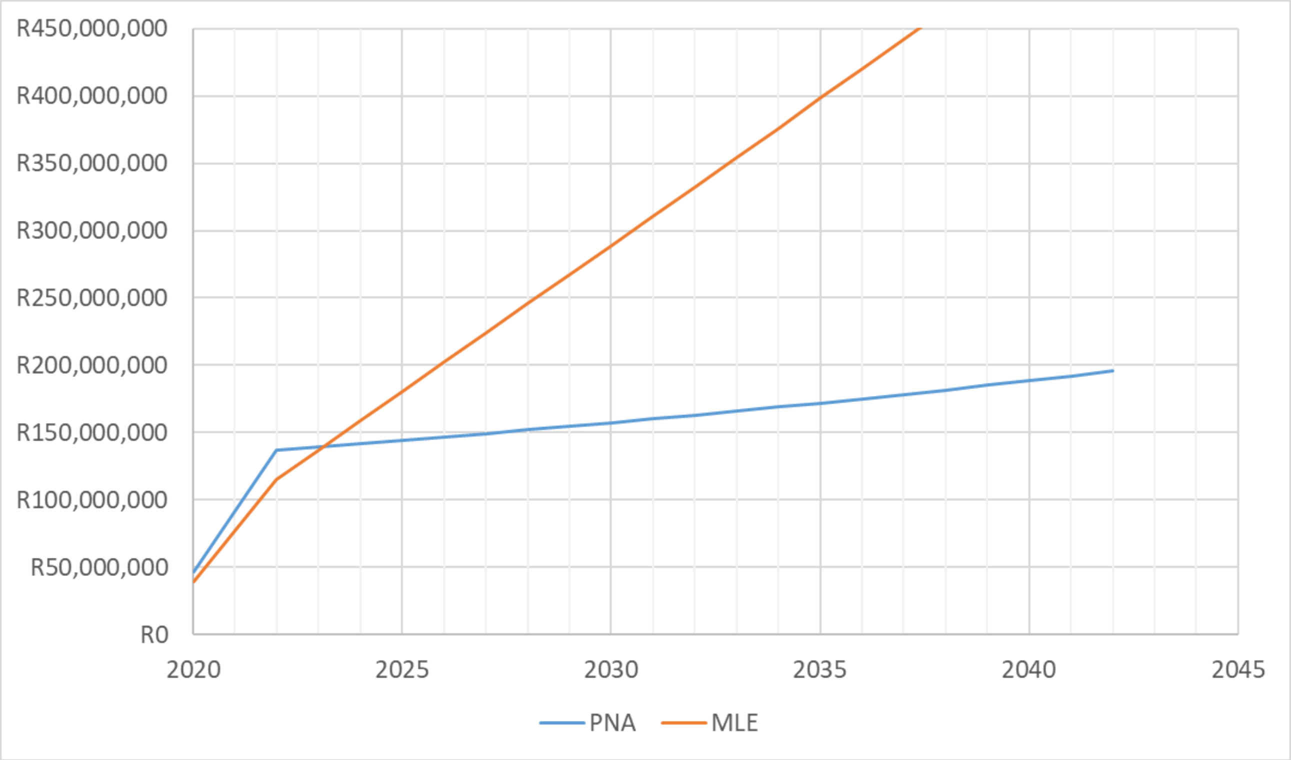

Figure 4.9: The cumulative cost of the PN/A process versus the MLE process represented as Present Value.

4.6 PN/A within a South African Context

In this section, the results of the research are discussed considering specific risks and challenges to South Africa in particular.

4.6.1 Seeding Requirements for the PN/A process

The amount of seed sludge is determined by the timeframe to start up the reactors. As discussed in Section 2.3.2, increasing the amount of seed sludge can dramatically reduce the start-up time which must be off set with the cost of harvesting and transporting the sludge to South Africa. Figure 4.10 indicates the time to achieve full production given that the reactors are inoculated with a given mass of seed sludge.

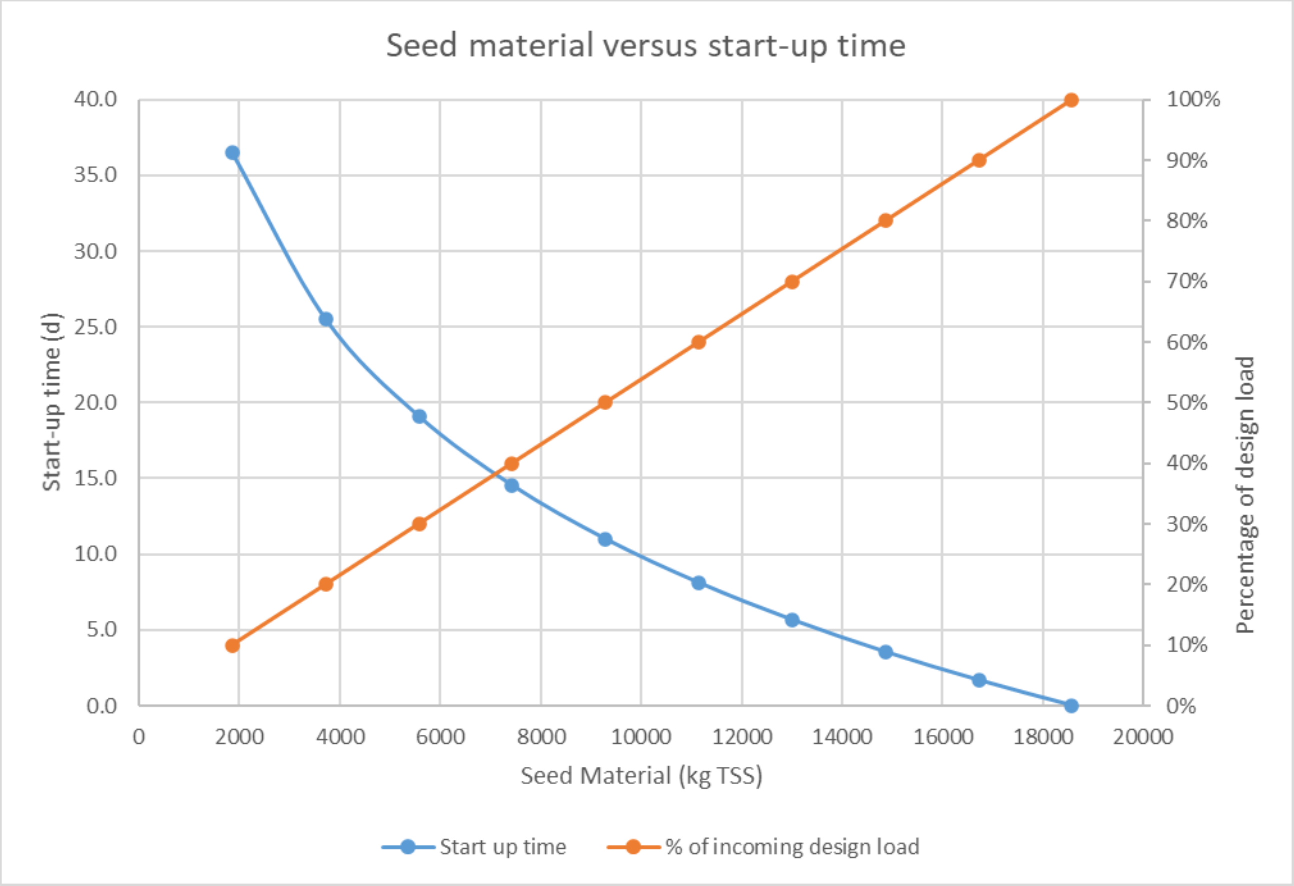

Figure 4.10: The amount of granular seed sludge versus theoretical start-up time to reach full production for a PN/A process

It can be seen from Figure 4.10 that the start-up time is not linear when it comes to seed material and therefore there is a diminishing return on the effect the amount of seed material has on reducing the start-up time. It is reasonable to assume that the cost of transporting the seed material will be approximately linear. When deciding on the amount of seeding material, one also needs to consider the turndown of the aeration system to ensure that the DO set point can be maintained such that the biological oxygen demand and the system is not over aerated. If too little seed material is used, a portion of the air from the blowers would need to be vented or perhaps smaller blowers would need to be rented for start-up. Considering a maximum turndown for the centrifugal blowers of 50%, and that each reactor would have two duty blowers and that each reactor treats half the sidestream nitrogen load (100% x 0.5 x 0.5 x 0.5= 12.5%), the PN/A system could operate at 12.5% of design load. From Figure 4.10, this would relate to a total seed mass of 2322 kg TSS (Approximately 77 m^3 at 30 kg TSS/m^3) and a start up time of 33 days. Allowing a reasonable safety factor for the seed sludge, 100 m^3 of sludge would be sufficient.

It is important to note that the literature values used to calculate the seed material requirements were based on laboratory tests on a single stage PN/A system (Li *et al.*, 2020) which represent ideal conditions. Although the results obtained in this study are significantly longer than the start-up times and seed sludge concentrations presented by Wang *et al.*(2018)*, f*or full scale applications it is likely that start-up times will be substantially longer than the ones indicated in Figure 4.10. This is due to factors such as practical commissioning challenges with full scale equipment, process upsets upstream of the PN/A plant at the new sludge AD plant, variability in sidestream liquor quality at start-up of the anaerobic digesters etc. In fact, from discussions with a prominent supplier of PN/A systems, a practical timeframe to start up a PN/A process can be up to 6 months (Colsen, personal E-mail communication, 20 May 2020).

A major risk in applying PN/A over an MLE process is the lack of seed sludge to start the process. Although this is not unique to South Africa, the risk is exacerbated by the distance South Africa is away from the seed sludge sources. This does not only pose a risk for start-up but also during operation in the case that the reactor sludge be killed or washed out due to poor operating conditions. Even in Europe, where there are many commercially operated plants, the availability and price of anammox seed sludge is variable creating a risk that when the seed sludge is needed, it is either not available or is at a premium price due to high demand.

This risk can be managed by explicitly stating the volumes of seed sludge and outlining the responsibility upfront in contractual agreements with technology suppliers. Technology suppliers should also only be selected with experience in transporting seed sludge across continents. This would not be the first time that an anammox project involved transporting seeding sludge across continents. As an example, projects in China have been seeded with sludge from Europe (Aquest Colsen, personal email communication, 02 Oct 2018).

The process risk of killing/washing out of sludge in a reactor is mitigated by having two reactors operating in parallel. Besides providing turndown which is beneficial during start-up, one reactor could be used to seed another if sludge loss occurs.

In contrast, for the ND process, as it utilises activated sludge, the seed could be sourced from sludge onsite at CF. This is clearly a lower risk and less costly application.

4.6.2 Economic challenges in the wastewater industry

From an economic perspective, Section 2.5.1 highlighted the budgetary issues faced by many municipalities in terms of prioritizing wastewater treatment projects. Section 2.5.3 further outlines the poor shape of the South African economy with structural reforms aimed at cutting public spending budgets. If one considers that the LCCA of the PN/A process costs only 35% of the cost of the MLE process, from an economic perspective the PN/A process is preferred and meets the budgetary challenges of the wastewater sector in South Africa.

4.6.3 Energy cost and climate change

Section 2.5.2 describes the status of the energy cost and supply in South Africa. The LCCA assumed a 10.2% increase in energy cost as an estimate. The actual increases are difficult to predict but one can say with a high degree of certainty that they will be above inflation for the next few years at least. The PN/A process, consuming only 46% of the power of the MLE process, is favoured in this regard and the overall cost to run the process will be less sensitive to electrical cost increases in the future. When one considers that 83% of the electrical power is from coal in South Africa (as highlighted in Section 2.5.2) the energy savings from the PN/A process will contribute to a reduction in greenhouse gases and limit climate change[4].

4.6.4 Availability of Technical Skills for PN/A Application

As presented in section 2.2.2.3, PN/A technology has been widely applied in Europe, the US and China (Bungay, 2012; Lackner *et al.*, 2014; Ali and Okabe, 2015) with no full scale applications to municipal wastewater in Africa to date. As with any new technology applied to a region, there is the lack of experience and skilled operational staff to install and run the facility. Most of the technical skills will need to be “imported” to construct and commission the PN/A plant but the real risk lies in what happens to the infrastructure once the plant is handed over to the municipality. In section 2.5.1, obtaining the correct expertise to service the requirements for current wastewater plants was highlighted as a challenge. If one considers that the PN/A process is completely novel within South Africa, this adds additional risk. From a skills perspective, running a conventional MLE process would carry lower risk as the process makes use of conventional activated sludge which municipal operators are experienced in at the mainstream works.

One mitigating factor to this is that Cape Flats WWTP falls within the Cape Town Metropolitan area. The human resource capacity issues highlighted, Section 2.5.1, although pertinent, are felt more acutely in the rural areas of the country in contrast to major metropoles. In the construction and commissioning phase, the skills risk can be further mitigated through careful tender procedural management to ensure that the winning South African contractor is paired up with an appropriate technology supplier who can demonstrate adequate experience (reference sites) and qualifications to provide the best level of technical support. The technology supplier should also have successfully seeded reactors across continents. Once the plant is handed over, one possible way of transferring

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[4] The electricity generated by the digestion facility will not be sufficient to offset the electrical demand of the mainstream works at Cape Flats WWTP. Any additional energy savings in the sidestream treatment process means more of the mainstream works electrical energy costs can be offset.

skills to the local municipality is by entering into an Operation and Maintenance (O&M) contract with the private sector whereby the correctly skilled personnel run the facility for a fixed term conducting training and knowledge transfer with municipality staff with the intention of the municipality taking over at the end of the contract. This will ensure a smooth transition of from the private sector operation back into the control of the municipalities. In addition to this, running an O&M contract could get past additional bureaucratic hurdles to plant maintenance as outlined in Section 2.5.1 brought on by the Municipal Finance Management Act (Act No. 5 of 2003). In fact, the introduction of O&M style contracts align with the National Water and Sanitation Master Plan, created by the South African Department of Water and Sanitation, which outlines a key action to "investigate and promote service delivery models such as BOTT (Build, Operate, Train and Transfer)" (South African Department of Water and Sanitation, 2018). Another mitigating factor can be the use of modern technology to allow remote connectivity. With the onset of the 4th Industrial Revolution, technology allows for advanced control systems to be installed which can allow the remote monitoring and control of processes. This can ensure that technical assistance is available in real-time during the commissioning, operation and training periods shortly after the plant is running and well into the O&M contract period. This type of technology is already widely available from technology suppliers. An example of this is "AQUAVISTA™ " supplied by Veolia which is a prominent technology supplier of PN/A processes (Veolia Water Technologies, 2020).

5 CONCLUSION

In this study the technical and economic feasibility of a PN/A process was compared to that of a MLE process for treating sidestream liquor from a regional digestion facility. The evaluation was performed making use of steady state models to calculate critical plant design factors to determine the required equipment sizes, operational parameters and capital and operational expenses. The models were evaluated using recognised plant performance indices such as the OCI and EQI and furthermore, the overall economic cost of each process was compared by performing a LCCA. The results of the evaluation were then contextualized within a South African environment by comparing the risks and benefits of each process considering the current state of the wastewater industry and other macro-economic factors.

The nitrogen load in the sidestream liquor was calculated by conducting a nutrient mass balance over the digestion facility. Furthermore, by considering the amount of nitrogen being removed in a struvite precipitation process upstream of the dewatering unit operation, it was determined that the sidestream liquor, although only representing approximately 1-2% of the hydraulic load to the mainstream works, accounted for 29% of the design TKN treatment capacity. The expected nitrogen load to the sidestream treatment process was expected to be 3415 kgN/d with approximately 80% of it being FSA. These values, calculated via a desktop study, aligned with expected values from existing commercial plants (Jardin, Thöle and Wett, 2014; Zhang *et al.*, 2016; Eskicioglu, Galvagno and Cimon, 2018; Gu *et al.*, 2018) .

When it came to sizing of the major process equipment, the total reactor volume of the MLE process was 53% smaller than the PN/A process. The primary reason for the larger PN/A reactor volume is that the PN/A process requires lower nitrogen loading rates to ensure that inhibitory effects, such as high concentrations of ammonia, are avoided. Due to both processes operating at similar hydraulic loads and MLSS (85 m^3/h and 3.7 g/l, and 85 m^3/h and 4.0 g/l for the PN/A and MLE process respectively), the overall settling tank surface area was approximately equal at 77 m^2 and 79 m^2 for the PN/A and MLE process respectively. The size of the aeration equipment for the MLE process was 76% larger than the PN/A system.

In terms of electrical energy usage, the treatment efficiencies of the PN/A process and MLE process came to 1.48 kWh/kg N and 2.85 kWh/kg N respectively. For both processes, the amount of energy required for aeration alone amounted to >88% of the required total electrical consumption of the process. The significantly lower aeration demand of the PN/A process was the greatest contributor to its superior electrical efficiency. It was observed that the PN/A process required more heating than the MLE process in winter conditions. However, the energy balance concluded that there is sufficient heat generated (in the form of hot water, refer to Section 3.2.1.7) at the digestion facility to meet the heating demand of either process. This confirms the suitability of processes operating above ambient temperatures to treat sidestream liquor in this application.

The OCI of the MLE process is more than 10.2 times that of the PN/A process. The largest contribution to the OCI (80%) of the MLE process is the cost for dosing methanol as an external carbon source. The capital cost of the MLE process is 16% less than the PN/A process due to smaller civil structures, less utilities and no requirement for the importation of seed sludge. However, it is imperative to note that even although the capital cost is initially less for the MLE process, the overall LCCA indicated that the MLE process is 2.8 times more expensive than the PN/A process over a 23 year period with the cost of the MLE process becoming more expensive after approximately 1 year of operation. From an economic perspective, the PN/A process is the outright favourite.

An important consideration for the PN/A process is that it has not been implemented on a commercial scale in South Africa and is completely novel for the region. This poses a unique set of challenges based on the environment South Africa currently finds itself in. The key issues identified were:

- The South African economy has shown low levels of growth for a number of years even before the COVID-19 pandemic which has resulted in large fiscal deficits. This filters down to the wastewater industry putting increased pressure on municipal budgets to build and operate infrastructure.
- The cost of electricity has far outstripped inflation since 2008 with the likelihood of above inflation increases well into the future. Coupled with this, South Africa has a constrained electrical grid supplied by greenhouse gas emitting, coal fire power stations.
- There is a lack of skills capacity in the wastewater industry.
- The municipalities are challenged with inefficient bureaucratic procedures that make operating and maintain existing plants problematic.

It was found that the lower overall life-cycle cost and higher energy efficiency of the PN/A process meets the economic and energy challenges currently faced in South Africa. However, the PN/A process poses more technical risks. It requires a higher level of skill and experience to operate and maintain as it is a novel biological process that is not as robust as conventional activated sludge systems. This is on the backdrop of an industry that is already struggling to maintain and operate the status quo. Further to this there is currently no significant source of seed sludge available in South Africa resulting in the need to import an estimated 100 m^3 of seed sludge from Europe or the USA in order to start the process within a reasonable timeframe i.e. 6 months. Consequently, from an operational and reactor start-up perspective, the MLE process is preferred.

The technical challenges highlighted can be mitigated with the following considerations:

- Careful control in the tendering procedure to ensure that the winning contractor is paired with a suitable technology supplier that can demonstrate successful reference sites and has successfully seeded reactors across continents.
- Introduction of Operation and Maintenance (O&M) contracts with the private sector who can run the new facility with appropriately sourced skills which must then be transferred to the municipality over the contract period to bridge the skills gap and generate technical capacity within the municipality.
- Making use of state-of-the-art technology (4th Industrial Revolution) to allow remote connection to skilled resources to provide real time technical assistance when required.

In closing, this study aimed to address the problem of selecting a suitable treatment process to treat sidestream liquor from a combined THP and AD process at Cape Flats WWTP. At the beginning of the research it was hypothesized that, based on the successful implementation of PN/A processes in other parts of the world for similar applications, a PN/A process would be the preferred process to implement at Cape Flats WWTP. From the conclusions above, this thesis determined that although a PN/A process is completely novel to South Africa for this application, the economic and process benefits are superior to a conventional MLE process thereby proving the hypothesis true. It is therefore recommended that the PN/A process be selected over conventional MLE for the treatment of sidestream liquor at Cape Flats WWTP.

6 RECOMENDATIONS

Based on the conclusions of this thesis, the following recommendations are made:

1. If the cost of an external carbon source could be reduced, this would make an MLE process more favourable. It may be possible to completely replace methanol or partly augment the carbon addition with another cheaper source of carbon. Some examples of potential cheaper sources:
    a. Glycerol by-product from local biodiesel suppliers (Consultation with a local supplier seems to indicate that crude glycerol at this scale would need to be imported which would negatively affect costs)
    b. Fermentation of primary sludge
    c. The use of water treatment residue from bulk water treatment plants.

    This presents an opportunity for further work where additional scenarios can be modelled and evaluated using the same methods presented in this thesis. The cost savings would need to be weighed up against the availability of the carbon source and the potential contaminants that could be introduced which could negatively affect the operation and effluent quality of the mainstream works.
2. The assumption regarding the negligible growth of PAOs occurring in the MLE configuration should be verified against a dynamic model to determine the accuracy of the assumption. In this work, the growth and death kinetics of PAOs with a methanol substrate should be determined. Currently the growth of PAOs with VFA substrate is well understood but this is not the case for methanol. This presents another opportunity for further research.
3. The sidestream liquor effluent quality of each process being returned to the mainstream treatment process should be modelled to determine the effect on the final effluent quality and hence a more accurate representation of the EQI. This presents another opportunity for further research.
4. A more detailed investigation into the operation of the anammox process by using more complex mathematical models could highlight possible bottlenecks. Furthermore, dynamic models could be used to investigate non-ideal scenarios such as pipe blockages resulting in poor feeding conditions, the effect of power cuts to the process, as well as the effect of metals and other dissolved compounds within the municipal sludge to name a few. The PN/A process is a novel technology with the potential to greatly benefit the South African wastewater treatment industry and laboratory and pilot scale plants should be set up and tested within a South African context such that the feasibility of implementing the technology at other sites can be investigated, as well as allowing for a local seed sludge to be available, resulting in the increased popularity of incorporating anammox technologies.
5. With reference to dynamic modelling mentioned in point 4, the results of the dynamic models

should be compared with the steady state models developed in this thesis to determine the key differences and more importantly, the significance of these differences to decision making and the commercial application of these technologies.

6. Once PN/A or MLE treatment is applied to sidestream liquor at Cape Flats WWTP, a study should be done taking actual plant data and using it to calibrate the steady state and dynamic models to further improve the tools for designing and troubleshooting future sidestream treatment commercial applications.
7. A formal Technical Risk Analysis should be done on the MLE and PN/A processes whereby the risks are numerically weighted for comparative purposes to verify if there are additional risks over and above the ones mentioned in this study, and to quantitatively compare to see how it would affect the conclusions of this work.

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8 APPENDIX

8.1 PN/A Process

8.1.1 PN/A model inputs

Table 8.1: PN/A model sidestream liquor input parameters

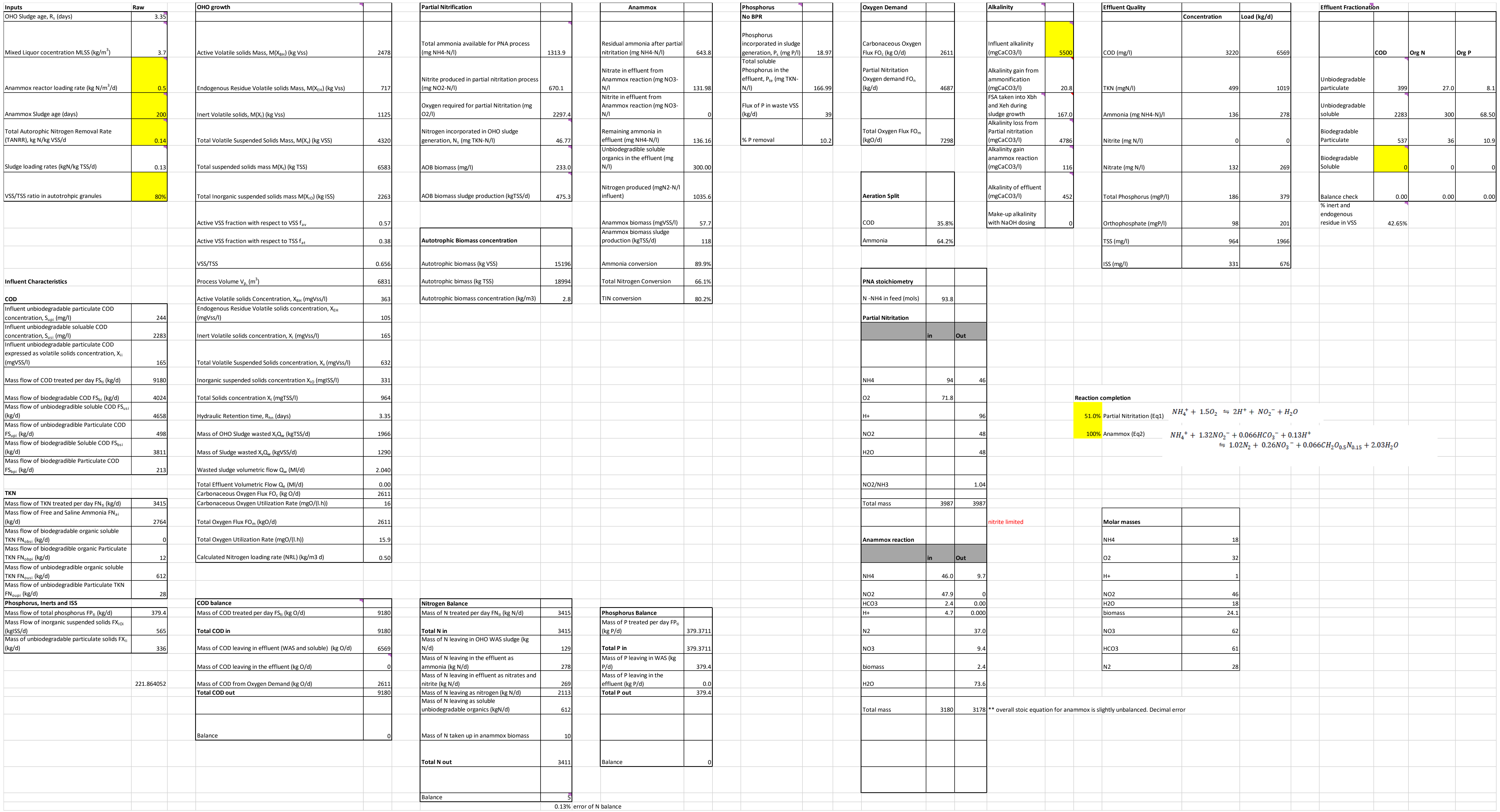
Sidestream liquor								
	COD	**TKN**	**TP**	**FSA**	**OP**	**TSS**	**ISS**	**VFA**
Sidestream liquor (unfiltered)	4500	1674	186			511	277	
Sidestream liquor (filtered)	4151	1655	177	1355	109			1868
Sidestream flow (Ml/d)	2.04							
Reactor Temperature (°C)	35							
VSS/TSS of solids in sidestream liquor	0.4583							

Table 8.2: Sidestream liquor fractionation results

Sidestream Liquor Fractionation	
COD (mg/l)	
Total	4500
Total soluble	4151
Unbiodegradable soluble	2283
Biodegradable soluble	1868
Total particulate	349
Unbiodegradable particulate	244
Biodegradable particulate	105
TKN (mg N/l)	
Total	1674.2
Ammonia	1355
Org N	319.42
Total soluble	300.00
Unbiodegradable soluble	300.00
Biodegradable soluble	0
Total particulate	19.4
Unbiodegradable particulate	13.59
Biodegradable particulate	5.83
P (mg P/l)	
Total	186.0
OP	109
Org P	77.21
Total soluble	68.5
Unbiodegradable soluble	68.5

Biodegradable soluble	0
Total particulate	8.7
Unbiodegradable particulate	6.10
Biodegradable particulate	2.61

8.1.2 PN/A Model Raw Outputs



Inputs	Raw
OHO Sludge age, R_s (days)	3.35
Mixed Liquor cocentration MLSS (kg/m^3)	3.7
Anammox reactor loading rate (kg N/m^3/d)	0.5
Anammox Sludge age (days)	200
Total Autorophic Nitrogen Removal Rate (TANRR), kg N/kg VSS/d	0.14
Sludge loading rates (kgN/kg TSS/d)	0.13
VSS/TSS ratio in autotrohpic granules	80%
Influent Characteristics	
COD	
Influent unbiodegradable particulate COD concentration, S_{upi} (mg/l)	244
Influent unbiodegradable soluable COD concentration, S_{usi} (mg/l)	2283
Influent unbiodegradable particulate COD expressed as volatile solids concentration, X_{ii} (mgVSS/l)	165
Mass flow of COD treated per day FS_{ti} (kg/d)	9180
Mass flow of biodegradable COD FS_{bi} (kg/d)	4024
Mass flow of unbiodegradible soluble COD FS_{usi} (kg/d)	4658
Mass flow of unbiodegradible Particulate COD FS_{upi} (kg/d)	498
Mass flow of biodegradible Soluble COD FS_{bsi} (kg/d)	3811
Mass flow of biodegradible Particulate COD FS_{bpi} (kg/d)	213
TKN	
Mass flow of TKN treated per day FN_{ti} (kg/d)	3415
Mass flow of Free and Saline Ammonia FN_{ai} (kg/d)	2764
Mass flow of biodegradable organic soluble TKN FN_{obsi} (kg/d)	0
Mass flow of biodegradible organic Particulate TKN FN_{obpi} (kg/d)	12
Mass flow of unbiodegradible organic soluble TKN FN_{ousi} (kg/d)	612
Mass flow of unbiodegradible Particulate TKN FN_{oupi} (kg/d)	28
Phosphorus, Inerts and ISS	
Mass flow of total phosphorus FP_{ti} (kg/d)	379.4
Mass Flow of inorganic suspended solids FX_{IOi} (kgISS/d)	565
Mass of unbiodegradable particulate solids FX_{Ii} (kg/d)	336
	221.864052

OHO growth	
Active Volatile solids Mass, $M(X_{BH})$ (kg Vss)	2478
Endogenous Residue Volatile solids Mass, $M(X_{EH})$ (kg Vss)	717
Inert Volatile solids, $M(X_i)$ (kg Vss)	1125
Total Volatile Suspended Solids Mass, $M(X_v)$ (kg VSS)	4320
Total suspended solids mass $M(X_t)$ (kg TSS)	6583
Total Inorganic suspended solids mass $M(X_{IO})$ (kg ISS)	2263
Active VSS fraction with respect to VSS f_{av}	0.57
Active VSS fraction with respect to TSS f_{at}	0.38
VSS/TSS	0.656
Process Volume $V_{p,}$ (m^3)	6831
Active Volatile solids Concentration, X_{BH} (mgVss/l)	363
Endogenous Residue Volatile solids concentration, X_{EH} (mgVss/l)	105
Inert Volatile solids concentration, X_i (mgVss/l)	165
Total Volatile Suspended Solids concentration, X_v (mgVss/l)	632
Inorganic suspended solids concentration X_{IO} (mgISS/l)	331
Total Solids concentration X_t (mgTSS/l)	964
Hydraulic Retention time, R_{hn} (days)	3.35
Mass of OHO Sludge wasted X_tQ_w (kgTSS/d)	1966
Mass of Sludge wasted X_vQ_w (kgVSS/d)	1290
Wasted sludge volumetric flow Q_w (Ml/d)	2.040
Total Effluent Volumetric Flow Q_e (Ml/d)	0.00
Carbonaceous Oxygen Flux FO_c (kg O/d)	2611
Carbonaceous Oxygen Utilization Rate (mgO/(l.h))	16
Total Oxygen Flux FO_m (kgO/d)	2611
Total Oxygen Utilization Rate (mgO/(l.h))	15.9
Calculated Nitrogen loading rate (NRL) (kg/m3 d)	0.50

COD balance	
Mass of COD treated per day FS_{ti} (kg O/d)	9180
Total COD in	9180
Mass of COD leaving in effluent (WAS and soluble) (kg O/d)	6569
Mass of COD leaving in the effluent (kg O/d)	0
Mass of COD from Oxygen Demand (kg O/d)	2611
Total COD out	9180
Balance	0

Partial Nitrification	
Total ammonia available for PNA process (mg NH4-N/l)	1313.9
Nitrite produced in partial nitritation process (mg NO2-N/l)	670.1
Oxygen required for partial Nitritation (mg O2/l)	2297.4
Nitrogen incorporated in OHO sludge generation, N_s (mg TKN-N/l)	46.77
AOB biomass (mg/l)	233.0
AOB biomass sludge production (kgTSS/d)	475.3

Autotrophic Biomass concentration	
Autotrophic biomass (kg VSS)	15196
Autotrophic bimass (kg TSS)	18994
Autotrophic biomass concentration (kg/m3)	2.8

Nitrogen Balance	
Mass of N treated per day FN_{ti} (kg N/d)	3415
Total N in	3415
Mass of N leaving in OHO WAS sludge (kg N/d)	129
Mass of N leaving in the effluent as ammonia (kg N/d)	278
Mass of N leaving in effluent as nitrates and nitrite (kg N/d)	269
Mass of N leaving as nitrogen (kg N/d)	2113
Mass of N leaving as soluble unbiodegradable organics (kgN/d)	612
Mass of N taken up in anammox biomass	10
Total N out	3411
Balance	5
0.13%	error of N balance

Anammox	
Residual ammonia after partial nititation (mg NH4-N/l)	643.8
Nitrate in effluent from Anammox reaction (mg NO3-N/l	131.98
Nitrite in effluent from Anammox reaction (mg NO3-N/l	0
Remaining ammonia in effluent (mg NH4-N/l)	136.16
Unbiodegradible soluble organics in the effluent (mg N/l)	300.00
Nitrogen produced (mgN2-N/l influent)	1035.6
Anammox biomass (mgVSS/l)	57.7
Anammox biomass sludge production (kgTSS/d)	118
Ammonia conversion	89.9%
Total Nitrogen Conversion	66.1%
TIN conversion	80.2%

Phosphorus Balance	
Mass of P treated per day FP_{ti} (kg P/d)	379.3711
Total P in	379.3711
Mass of P leaving in WAS (kg P/d)	379.4
Mass of P leaving in the effluent (kg P/d)	0.0
Total P out	379.4
Balance	0

Phosphorus	
No BPR	
Phosphorus incorporated in sludge generation, P_s (mg P/l)	18.97
Total soluble Phosphorus in the effluent, P_{te} (mg TKN-N/l)	166.99
Flux of P in waste VSS (kg/d)	39
% P removal	10.2

Oxygen Demand	
Carbonaceous Oxygen Flux FO_c (kg O/d)	2611
Partial Nitritation Oxygen demand FO_n (kg/d)	4687
Total Oxygen Flux FO_m (kgO/d)	7298

Aeration Split	
COD	35.8%
Ammonia	64.2%

PNA stoichiometry		
N -NH4 in feed (mols)	93.8	
Partial Nitritation		
	in	Out
NH4	94	46
O2	71.8	
H+		96
NO2		48
H2O		48
NO2/NH3		1.04
Total mass	3987	3987
Anammox reaction		
	in	Out
NH4	46.0	9.7
NO2	47.9	0
HCO3	2.4	0.00
H+	4.7	0.000
N2		37.0
NO3		9.4
biomass		2.4
H2O		73.6
Total mass	3180	3178

nitrite limited

** overall stoic equation for anammox is slightly unbalanced. Decimal error

Alkalinity	
Influent alkalinity (mgCaCO3/l)	5500
Alkalinity gain from ammonification (mgCaCO3/l)	20.8
FSA taken into Xbh and Xeh during sludge growth	167.0
Alkalinity loss from Partial nitritation (mgCaCO3/l)	4786
Alkalinity gain anammox reaction (mgCaCO3/l)	116
Alkalinity of effluent (mgCaCO3/l)	452
Make-up alkalinity with NaOH dosing	0

Reaction completion

51.0%	Partial Nitritation (Eq1)
100%	Anammox (Eq2)

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

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

Molar masses	
NH4	18
O2	32
H+	1
NO2	46
H2O	18
biomass	24.1
NO3	62
HCO3	61
N2	28

Effluent Quality	Concentration	Load (kg/d)
COD (mg/l)	3220	6569
TKN (mgN/l)	499	1019
Ammonia (mg NH4-N)/l	136	278
Nitrite (mg N/l)	0	0
Nitrate (mg N/l)	132	269
Total Phosphorus (mgP/l)	186	379
Orthophosphate (mgP/l)	98	201
TSS (mg/l)	964	1966
ISS (mg/l)	331	676

Effluent Fractionation

	COD	Org N	Org P
Unbiodegradable particulate	399	27.0	8.1
Unbiodegradable soluble	2283	300	68.50
Biodegradable Particulate	537	36	10.9
Biodegradable Soluble	0	0	0
Balance check	0.00	0.00	0.00
% inert and endogenous residue in VSS	42.65%		

Figure 8.1: Spreadsheet results from the PN/A model biological processes

Table 8.3: SST design inputs and results for PN/A process.

SST design Inputs		
PWWF factor	1	
PDWF	1	
Sludge DSVI	125	ml/g
Sludge SSVI	83.75	
V0/n	17.78	
n	0.389	
V0	6.91	
Flux rating	80%	
MLSS, Xf	3.7	g/l
Design Equation outputs		
$X_{R,crit}$	10.3	kg/m3
$R_{,crit}$	0.516	
$q_{R,crit}$	0.94	m/h
X_{Lmin}	5.14	kg/m3
J_{Lmax}	9.6	kg/m2 h
$q_{AmaxSHCII}$	1.61	m/h
$R_{min,PWWF}$	0.580	
q_{ASHCI}	1.61	m/h
From Diurnal data		
ADWF	85	m3/h
PDWF	85	m3/h
PWWF	85	m3/h
Required settling tank area, A_{ST}	66	m2
Conventional settling tank sizing		
SST diameter	7.0	m
number of SSTs	2.0	
Area per SST	38.5	m2
Total area	77	m2
$Q_{R, crit}$	72	m3/h
qa at PWWF	1.10	m/h
Rmin at PWWF	0.58	
Qr at PWWF	49	m3/h

8.1.3 PN/A EQI outputs

Table 8.4: EQI calculation results for the PN/A process

Effluent Quality Index				
	Limit	**Predicted**	**βx**	**PUx**
COD sol. (mg/l)	75	2283	1	2283.3
COD part. (mg/l)	75	937	1	936.7
Ammonia (mg NH4-N)/l	10	136	7.5	136.2
Nitrate (mg NO3-N/l)	10	132	7.5	132.0
Phosphorus (mgP/l)	5	185.97	15	186.0
TSS (mg/l)	25	964	3	963.7
EQI	**22260**	**kg pollutants/d**		

8.1.4 PN/A – Aeration calculation

Table 8.5: Inputs and calculation results of the aeration requirements for the PN/A process.

Fine Bubble Aeration Calculation	
Actual oxygen requirement (kg/h)	304
Diffuser submergence (est.)	5.2
OTE per meter	7
Oxygen Transfer Efficiency, OTE	36.40
Water temperature	35
Elevation (m)	0
Atmospheric pressure at elevation, Psite (kPa)	101
Atmospheric pressure at elevation, Psite (mmHg)	761
Water saturated vapour pressure, psite (mmHg)	42.2
DO saturation concentration at atmospheric pressure corrected for temperature, C_s (mg/l)	7.03
DO saturation corrected for water depth and temperature, $C_{s,corrected}$ (mg/l)	8.61
Pressure at aerator depth (mmHG gauge)	382
Assumed operating oxygen concentration (mg/l)	0.3
Beta factor to account for wastewater characteristics	0.95
Alpha factor to correct for aerator type, basin geometry, degree of mixing and wastewater	0.6
Fouling factor for diffusers	0.9
Combined AOR/SOR correction factor	0.669
Standard oxygen requirement, SOR (kg/h)	454
Adjust SOR for partial dissolution, i.e. incorporate OTE	1248
Mass flow of air required, (kg/h)	5379
Volumetric air flow required at 20°C and 0% RH, (m3/h)	4482

Assumed FBDA equipment efficiency (kgO2/kWh)	5
Aeration energy requirement (shaft power)	90.8
Shaft to wire efficiency	70%
Total aeration energy (kW)	129.8

8.1.5 PN/A – Energy Balance

Table 8.6: PN/A energy balance inputs

Operating Conditions	**Winter ave**	**Winter low**	
Sidestream Feed Flow rate	85	85	m3/h
Sidestream Feed Temperature	25	25	°C
Reactor Operating temperature	35	35	°C
Ambient Temperature	13	8	°C
Ground temperature	13	8	°C

Table 8.7: PN/A major equipment dimensions and heat transfer areas

Balancing tank dimensions		
Design retention time of tank	12	hours
Volume of balancing tank working	1020	m3
Tank feeboard	30%	
Total tank volume	1457	m3
Tank Height	6	m
Tank Diameter	18	m
Tank Wall thickness	0.25	m
Surface area of walls	339	m2
Surface area through floor	254	m2
Surface area of roof	254	m2
Reactor Dimensions		
Number of reactors	2	
Reactor working volume	3415	m3
Reactor Freeboard	10%	
Reactor Total volume	3795	m3
Reactor Height	6	m
Reactor L/W ratio	2	
Reactor Length	36	m
Reactor Width	18	
Reactor Wall thickness	0.25	m
Total Surface area of walls	1280	m2
Total Surface area through floor	1265	m2
Total Surface area of roof	1265	m2
SST dimensions		

number of SST's	2	
Sidewall height	3	
SST diameter	7	
SST Area	38	m2
SST Volume	115	m3
SST wall thickness	0.25	m
Surface area of walls	66	m2
Surface area through floor	38	m2
Surface area of roof	38	m2

Table 8.8: Calculation of overall heat transfer co-efficients for PN/A process

Overall heat transfter Coefficient		
Convective Heat Transfer Inside Wall, h_i	10000	W/m2 K
Concrete Wall thickness	0.25	m
Concrete Plinth thickness	0.5	m
Concrete conductivity	1.21	W/m K
Convective Heat Transfer Outside Wall, h_o	40.8	W/m2 K
Convective Heat Transfer tank head space, h_o	26.2	
$U_{\text{concrete walls, no insulation}}$	4.3	W/m2 K
$U_{\text{concrete walls, with insulation}}$	1.2	W/m2 K
$U_{\text{roof, closed concrete}}$	3.7	W/m2 K
$U_{\text{roof, open}}$	40.6	W/m2 K
$U_{\text{floor concrete}}$	1.6	W/m2 K

Table 8.9: Calculation of heating requirements of PN/A process.

Heat losses to environment	**Winter**	**Winter low**	
Balancing tank			
Walls (concrete)	17.4	24.9	kW
Floor (concrete)	4.9	7.0	kW
Roof (closed)	11.2	16.1	kW
Total	33.5	48.0	kW
Reactors			
Walls (concrete)	121.1	149.5	kW
Floor (concrete)	44.6	55.1	kW
Roof (closed)	102.7	126.8	kW
Total	268.4	331.4	kW
SST			
Walls (concrete)	6.2	7.7	kW
Floor (concrete)	1.4	1.7	kW
Roof (open)	34.2	42.2	kW
Total	41.8	51.6	kW
Total Heat loss through equipment	**344**	**431**	**kW**
Total Sensible heat loss	**988**	**988**	**kW**
Total heating requirement	**1332**	**1419**	**kW**

8.1.6 Operational Cost Index Calculations for PN/A process

Table 8.10: OCI inputs and calculation results for the PN/A process

Operational Cost Index calculations		
PN/A Aeration Energy	130	kW
Additional aeration energy to process effluent in Cape Flats WWTP	40.7	kW
Pumping Energy		
S recycle		
Volumetric flow rate	85	m3/h
Hydraulic Lift Height	10	m
Hydraulic power	2.3	kW
Hydraulic to wire efficiency	50%	

Total Power	4.63	kW
Feed Pump Power		
Volumetric flow rate	85	m3/h
Pump differential head	10	m
Hydraulic power	2.3	kW
Hydraulic to wire efficiency	50%	
Total Power	4.63	kW
Total pumping energy (kW)	**9.27**	**kW**
Mixing Energy		
Reactor Volume	6831	m3
Mixing Energy requirement	5	W/m3
Mixing power reduction factor	0.25	
Total Mixing Energy	**8.5**	**kW**
Electrical energy cost		
Total power consumption	188.3	kW
Electricty cost at base date	1.07	R/kWh
Electricity price increase 2019/20	5.80%	
Electricty price increase from 2020/21	6.90%	
Calculated electricty cost for 2020	1.21	R/kWh
Total electricity cost at base date	**R1,765,155**	
Heating energy		
Total heating demand	**1332**	**kW**
Sludge Handling Cost		
Total Waste sludge produced	1431	kg TSS/d
Sludge disposal cost	R812.00	R per t
Sludge disposal cost per day	R1,162.04	
Sludge disposal cost per year	R424,146	
OCI	**R2,189,301**	

8.1.7 Capital cost breakdown for PN/A process

Table 8.11:A breakdown of the capital cost estimate for the PN/A process

Item No.	Description	Qty	Budget Unit Cost	Section Budget Estimate
100.00	**PN/A Process Capital Cost Estimate**			
101.00	**Reactors**			
101.01	FBDA network in reactor 1 and 2	1	R 1,675,714	R 1,675,713.52
101.02	Drum seive	2	R 585,640	R 1,171,280.00
101.03	Sludge pumps	4	R 100,000.00	R 400,000.00
101.04	Sludge piping and fittings	100	R 2,420.00	R 726,000.00
101.05	Valves on sludge piping	8	R 20,000.00	R 160,000.00
101.06	Treated water piping	100	R 2,420.00	R 726,000.00
101.08	Settling tank	2	R 282,303.78	R 564,607.56
101.09	Mixers	6	R 300,000.00	R 1,800,000.00
101.11	Galvanized steel platforms and walkways	1	R 450,000.00	R 450,000.00
101.12	Installation, commissioning and upholding			R 1,534,720.22
				R 9,208,321.30
102.00	**Aeration system**			
102.01	Blowers	5	R 1,046,973	R 5,234,866.98
102.02	Aeration piping main header	50	R 9,146.10	R 1,143,263.00
102.03	Aeration piping sub headers	30	R 9,146.10	R 823,149.36
102.04	Control valves	4	R 100,000.00	R 400,000.00
102.05	Subheader isolation valves	2	R 50,000.00	R 100,000.00
102.06	Blower Isolation valves	5	R 40,000.00	R 200,000.00
102.07	Blower NRV's	5	R 40,000.00	R 200,000.00
102.08	Galvanized steel pipe bridge	1	R 175,000.00	R 175,000.00
102.09	Installation, commissioning and upholding			R 1,655,255.87
				R 9,931,535.21
103.00	**Utilities**			
103.01	Hot Water Piping	300	R 3,000.00	R 2,250,000.00
103.02	Hot Water booster pumps	2	R 100,000.00	R 500,000.00
103.03	Cooling water piping	300	R 5,000.00	R 3,000,000.00
103.04	Cooling water pumps	3	R 150,000.00	R 450,000.00
103.05	Heating Control valves	2	R 75,000.00	R 150,000.00
103.06	Treated Effluent piping	150	R 2,000.00	R 750,000.00
103.07	Treated Effluent control valves	2	R 75,000.00	R 150,000.00
103.08	NRVs on treated effluent lines	2	R 20,000.00	R 40,000.00
103.09	Washwater booster pump	2	R 50,000.00	R 100,000.00

103.10	NRV's on heating and cooling lines	5	R 20,000.00	R 100,000.00
103.11	Isolation valves on heating and cooling pipes	20	R 40,000.00	R 800,000.00
103.12	Installation, commissioning and upholding			R 1,658,000.00
				R 9,948,000.00
104.00	**HVAC**			
104.01	HVAC fans	3	R 55,000.00	R 165,000.00
104.02	HVAC filters	sum	R 40,000.00	R 40,000.00
104.03	Blower house louvres	4	R 25,000.00	R 100,000.00
104.04	Blower house ducting	sum	R 25,000.00	R 25,000.00
104.05	Air conditioning for MCC room	1	R 15,000.00	R 15,000.00
104.06	Blower house chemical pump room HVAC	1	R 30,000.00	R 30,000.00
104.07	Blower house office HVAC	1	R 35,000.00	R 35,000.00
104.08	Blower house bathroom HVAC	1	R 8,000.00	R 8,000.00
104.09	Installation, commissioning and upholding			R 4,062,800.00
				R 4,480,800.00
105.00	**Chemical Dosing**			
105.01	Stainless Steel Caustic Soda Storage tank	1	R 600,000.00	R 600,000.00
105.02	Caustic dosing pumps	3	R 60,000.00	R 180,000.00
105.03	Heat tracing	120	R 383.33	R 45,999.36
105.04	Anti foam dosing pumps	3	R 40,000.00	R 120,000.00
105.05	Piping and fittings SS316	120	R 450.00	R 162,000.00
105.06	Caustic offloading piping	20	R 650.00	R 39,000.00
105.07	Valves	sum	R 5,000.00	R 5,000.00
105.08	Installation, commissioning and upholding			R 230,399.87
				R 1,382,399.23
106.00	**Buffer Tank**			
106.01	Vertical shaft mixers	4	R 150,000.00	R 600,000.00
106.02	Lifting davits/ guide rails	3	R 45,000.00	R 135,000.00
106.03	Galvanized steel platforms and walkways	1	R 300,000.00	R 300,000.00
106.04	Feed pumps	4	R 100,000.00	R 400,000.00
106.05	Feed piping to tank	130	R 3,000.00	R 1,170,000.00
106.06	Discharge piping to reactor	120	R 2,000.00	R 720,000.00
106.07	Overflow/drain piping	30	R 5,200.00	R 312,000.00
106.08	Isolation valves	20	R 45,000.00	R 900,000.00
	Installation, commissioning and upholding			R 907,400.00
				R 5,444,400.00
107.00	**Seed Sludge**			
107.01	Seed sludge	100	R 15,030	R 1,503,000
	Shipping containers	5	R 8,000	R 40,000
	Duties and fees	5	R 4,000	R 20,000

107.02	Commissioning and upholding			R 375,750
				R 1,938,750.00
108.00	**Electrical and instrumentation**			
108.01	MV ring distribution network			R 919,321
108.02	Transformers			R 899,609
108.03	LV MCC			R 4,721,992
108.04	BBF SCADA Integration			R 98,450
108.05	Ancilliary			R 431,753
108.06	Instrumentation			R 3,187,100
108.07	Indicators & Valves			R 1,810,652
108.08	Lighting			R 305,923
108.09	CCTV			R 738,772
108.10	Access Control & Fire Detection			R 175,040
108.11	Cabling and termination			R 1,065,115
108.12	Earthing and bonding			R 170,490
108.13	Cable supports			R 474,700
108.14	Civils Electrical work			R 177,515
	Total			**R 15,176,431.41**
109.00	**Civil Costs**			
109.01	Site reshaping	sum	R5,493,846	R 5,493,846.15
109.02	Anammox reactors	sum	R13,356,923	R 13,356,923.08
109.03	SST's		R1,141,538	R 1,141,538.46
109.04	Anammox reactor buffer tank	sum	R3,907,692	R 3,907,692.31
109.05	Sodium hydroxide tank ground slab	sum	R112,308	R 112,307.69
109.06	Anammox MCC and blower house	sum	R5,098,462	R 5,098,461.54
109.07	Road and offloading bay	sum	R3,626,154	R 3,626,153.85
109.08	Underground drainage	sum	R824,615	R 824,615.38
109.09	Application, approvals and permits	sum	R76,923	R 76,923.08
109.10	Additional civil items (stairs, ladders, piping support, etc)	sum	R384,615	R 384,615.38
				R 34,023,076.92
110.00	**P&G's**			
110.01	Mechanical P&Gs			R 8,466,841
110.02	Electrical P&G's			R 3,035,286
110.03	Civil			R 5,103,462
				R 16,605,588.97
110.00	**Extras**			
111.01	Health and Safety			R 500,000.00
111.02	Environment			R 250,000.00
111.03	Sundries			R 750,000.00
				R 1,500,000.00
	SUB- TOTAL (equipment & contracts)			R 109,639,303.05

PROJECT CONTINGENCIES 15 % of COW	R 16,445,895.46
TOTAL PROJECT COSTS (excl. VAT)	R 126,085,199
PROJECT FEE CATEGORY	F (water and wastewater treatment works)
PROJECT FEE	13.00%
PROFESSIONAL FEE % ADJUSTED FOR COW	10.67%
PROJECT FEE	R13,451,365
TOTAL PROJECT COST (excl. VAT)	R 139,536,563
VAT @15%	R 20,930,484
TOTAL PROJECT COSTS (incl. VAT)	R 147,015,683

8.1.8 Lifecycle Cost Analysis calculation results for PN/A process

Table 8.12: Life-cycle Cost Analysis inputs for PN/A and MLE process.

Life-cycle Cost Analysis Inputs	
Base date (year), A_0	2020
number of years, n	23
Energy escalation, e1	10.2%
Sludge disposal escalation rate, e2	5.20%
Chemical escalation rate, e3	6.19%
CAPEX escalation rate, e4	5.2%
discount rate, d	7.00%
Construction length, years	3

Table 8.13: PN/A Lifecycle Cost Analysis calculation outputs

		CAPEX			**OPEX**									**Net Present Value**	**Cumulative net present valu**
Year	n				**Energy**			**Sludge disposal cost**			**Chemical Cost**				
		Cost						Cost			Cost				
2020	0	R46,512,188	1.00	1.00	R1,765,154.78	1.00	1.00	R424,146.23	1.00	1.00	R0.00	1.00	1.00	R46,512,188	R46,512,188
2021	1	R48,930,822	0.98	1.98	R1,945,200.56	1.03	2.03		0.98	1.98		0.99	1.99	R45,729,740	R92,241,928
2022	2	R51,475,224	0.97	2.95	R2,143,611.02	1.06	3.09		0.97	2.95		0.98	2.98	R44,960,454	R137,202,382
2023	3		0.95	3.90	R2,362,259.34	1.09	4.18		0.95	3.90		0.98	3.95	R2,331,406	R139,533,788
2024	4		0.93	4.83	R2,603,209.80	1.13	5.31		0.93	4.83		0.97	4.92	R2,382,294	R141,916,082
2025	5		0.92	5.75	R2,868,737.20	1.16	6.47		0.92	5.75		0.96	5.89	R2,435,021	R144,351,103
2026	6		0.90	6.66	R3,161,348.39	1.19	7.66		0.90	6.66		0.96	6.84	R2,489,636	R146,840,738
2027	7		0.89	7.54	R3,483,805.93	1.23	8.89		0.89	7.54		0.95	7.79	R2,546,190	R149,386,929
2028	8		0.87	8.42	R3,839,154.13	1.27	10.16		0.87	8.42		0.94	8.73	R2,604,738	R151,991,666
2029	9		0.86	9.28	R4,230,747.85	1.30	11.46		0.86	9.28		0.93	9.67	R2,665,332	R154,656,998
2030	10		0.84	10.12	R4,662,284.13	1.34	12.80		0.84	10.12		0.93	10.59	R2,728,029	R157,385,028
2031	11		0.83	10.95	R5,137,837.12	1.38	14.18		0.83	10.95		0.92	11.51	R2,792,888	R160,177,916
2032	12		0.82	11.77	R5,661,896.50	1.42	15.61		0.82	11.77		0.91	12.43	R2,859,968	R163,037,884
2033	13		0.80	12.57	R6,239,409.94	1.47	17.08		0.80	12.57		0.91	13.33	R2,929,331	R165,967,215
2034	14		0.79	13.36	R6,875,829.76	1.51	18.59		0.79	13.36		0.90	14.23	R3,001,040	R168,968,255
2035	15		0.78	14.13	R7,577,164.39	1.56	20.14		0.78	14.13		0.89	15.12	R3,075,161	R172,043,416
2036	16		0.76	14.89	R8,350,035.16	1.60	21.74		0.76	14.89		0.89	16.01	R3,151,762	R175,195,178
2037	17		0.75	15.64	R9,201,738.75	1.65	23.39		0.75	15.64		0.88	16.89	R3,230,912	R178,426,089
2038	18		0.74	16.38	R10,140,316.10	1.70	25.09		0.74	16.38		0.87	17.76	R3,312,683	R181,738,772
2039	19		0.72	17.10	R11,174,628.34	1.75	26.84		0.72	17.10		0.87	18.62	R3,397,150	R185,135,922
2040	20		0.71	17.82	R12,314,440.43	1.80	28.65		0.71	17.82		0.86	19.48	R3,484,388	R188,620,310
2041	21		0.70	18.52	R13,570,513.36	1.86	30.50		0.70	18.52		0.85	20.34	R3,574,477	R192,194,788
2042	22		0.69	19.21	R14,954,705.72	1.91	32.42		0.69	19.21		0.85	21.18	R3,667,498	R195,862,286
Present Value, PV		**R137,202,382**			**R51,765,110**			**R6,894,794**			**R0**				
Total PV		**R195,862,286**													

8.1.9 PN/A reactor seeding requirements

Table 8.14: Inputs for seed sludge calculation.

Seed sludge inputs		
Seed sludge solids concentration	30	kg TSS/m3
VSS/TSS ratio	80%	
Anammox sludge % in VSS	27%	
Anammox doubling time	11	days

Table 8.15: The amount of seed sludge required versus the TKN load and nitrogen removal rate achievable.

Start-up time									
Percentage of total load	**Nitrogen load in feed (kgN/d)**	**Nitrogen removal**	**TANRR kg N/kg VSS/d**	**kg VSS**	**kg TSS**	**kg pure anammox**	**Pure Anammox concentration (kg/m3)**	**Seed sludge volume (m3)**	**Estimated start up time to full production**
10%	342	226	0.14	1612	2015	435	0.064	67	36.5
13%	427	282	0.14	2015	2518	544	0.080	84	33.0
30%	1025	677	0.14	4835	6044	1305	0.191	201	19.1
40%	1366	903	0.14	6447	8059	1741	0.255	269	14.5
50%	1708	1128	0.14	8059	10073	2176	0.319	336	11.0
60%	2049	1354	0.14	9670	12088	2611	0.382	403	8.1
70%	2391	1579	0.14	11282	14102	3046	0.446	470	5.7
80%	2732	1805	0.14	12894	16117	3481	0.510	537	3.5
90%	3074	2031	0.14	14505	18132	3916	0.573	604	1.7
100%	3415	2256	0.14	16117	20146	4352	0.637	672	0.0

8.1.10 PN/A Incremental effect on mainstream works

Table 8.16: Treated sidestream liquor effluent inputs into the 5-Stage bardenpho model used to calculate the incremental increase on sludge production and aeration at the mainstream works.

WW influent inputs									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**	**TSS**	**ISS**	**VFA**
Raw sewage	3220		499	186			964	331	
0,45um membrane	2283		436	167	136	98			0
WW effluent inputs									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**			
Effluent (raw)	2283		300.0	68.50					
Effluent (settled)	2283		300.0	68.50					

Inputs	Settled
Sludge age, R_s (days)	15
Mixed Liquor cocentration MLSS (kg/m^3)	4
Unaerated mass fraction f_x	0.450
Factor of safety on nitrification, S_f	1.25
S recycle ratio	1.2
A recycle ratio	0.58
Nitrate concentration in Influent, Nn_i (mg/l)	0.00
Dissolved oxygen concentrations in A recycle, O_a (mgO/l)	2
Dissolved oxygen concentrations in S recycle, O_s (mgO/l)	0
DO concentration in Influent, O_i (mg/l)	0
Nitrate concentration in the S recycle, N_{nr} (mg/l)	0
Anaerobic mass fraction	0.09
Number of anaerobic compartments	1
Influent Characteristics	
COD	
Influent unbiodegradable particulate COD concentration, S_{upi} (mg/l)	399.50
Influent unbiodegradable soluable COD concentration, S_{usi} (mg/l)	2283.286
Influent unbiodegradable particulate COD expressed as volatile solids concentration, X_{Ii} (mgVSS/l)	270
Mass flow of COD treated per day FS_{ti} (kg/d)	6569
Mass flow of biodegradable COD FS_{bi} (kg/d)	1096
Mass flow of unbiodegradible soluble COD FS_{usi} (kg/d)	4658
Mass flow of unbiodegradible Particulate COD FS_{upi} (kg/d)	815
Mass flow of biodegradible Soluble COD FS_{bsi} (kg/d)	0
Mass flow of biodegradible Particulate COD FS_{bpi} (kg/d)	1096
TKN	
Mass flow of TKN treated per day FN_{ti} (kg/d)	1019
Mass flow of Free and Saline Ammonia FN_{ai} (kg/d)	278
Mass flow of biodegradable organic soluble TKN FN_{obsi} (kg/d)	0
Mass flow of biodegradible organic Particulate TKN FN_{obpi} (kg/d)	74
Mass flow of unbiodegradible organic soluble TKN FN_{ousi} (kg/d)	612
Mass flow of unbiodegradible Particulate TKN FN_{oupi} (kg/d)	55.0
Phosphorus, UPO and ISS	
Mass flow of total phosphorus FP_{ti} (kg/d)	379
Mass Flow of inorganic suspended solids FX_{IOi} (kgISS/d)	676
Mass of unbiodegradable particulate solids FX_{Ii} (kg/d)	550

Activated Sludge Growth	Tmin
AS system	**Settled**
Active OHO solids Mass, $M(X_{BH})$ (kg Vss)	1681
OHO Endogenous Residue solids Mass, $M(X_{EH})$ (kg Vss)	1143
Inert Volatile solids, $M(X_i)$ (kg Vss)	8254
Total Volatile Suspended Solids Mass, $M(X_v)$ (kg VSS)	11079
Total suspended solids mass $M(X_t)$ (kg TSS)	21466
Total Inorganic suspended solids mass $M(X_{IO})$ (kg ISS)	10387
Active VSS fraction with respect to VSS f_{av}	0.15
Active VSS fraction with respect to TSS f_{at}	0.08
VSS/TSS ratio	0.516
Process Volume $V_{p,}$ (m^3)	5367
OHO Active Concentration, X_{BH} (mgVss/l)	313
OHO Endogenous Residue concentration, X_{EH} (mgVss/l)	213
Inert Volatile solids concentration, X_i (mgVss/l)	1538
Total Volatile Suspended Solids concentration, X_v (mgVss/l)	2064
Inorganic suspended solids concentration X_{IO} (mgISS/l)	1936
Total Solids concentration in aerobic zone X_t (mgTSS/l)	3929
Hydraulic Retention time, R_{hn} (days)	2.63
Mass of Sludge wasted X_tQ_w (kgTSS/d)	1431
Mass of Sludge wasted X_vQ_w (kgVSS/d)	739
Wasted sludge volumetric flow Q_w (Ml/d)	0.358
Effluent Volumetric Flow Q_e (Ml/d)	1.7
OHO Carbonaceous Oxygen Flux FO_c (kg O/d)	817
OHO Carbonaceous Oxygen Utilization Rate (mgO/(l.h))	11
Total Oxygen Flux FO_m (kgO/d)	1690
Total Oxygen Utilization Rate (mgO/(l.h))	23.44

Nitrification	Tmin
	Settled
minimum sludge age, S_{fmin} (days)	6.3
Factor of safety on nitrification, S_f (days)	2.4
Maximum unaerated mass fraction, f_{xm}	0.634
No Nitrification	
Free and Saline ammonia in reactor and effluent, N_{ae} (mg TKN-N/l)	163.2
Soluble unbiodegradable organic nitrogen, N_{ouse} (mg TKN-N/l)	300
Total soluble TKN in the effluent, N_{te} (mg TKN-N/l)	463.2
Nitrogen incorporated in sludge generation, N_s (mg TKN-N/l)	36.2
Concentration of nitrates generated in the reactor and in effluent, N_c (mg TKN-N/l)	0
Mass of nitrifier organisms, $M(X_A)$ kg VSS	0
Conc. of nitrifier organisms, X_A mgVSS/l	0
% N removal	7.2
Nitrification	
Free and Saline ammonia in reactor and effluent, N_{ae} (mg TKN-N/l)	0.90
Soluble unbiodegradable organic nitrogen, N_{ouse} (mg TKN-N/l)	300.00
Total soluble TKN in the effluent, N_{te} (mg TKN-N/l)	300.90
Nitrogen incorporated in sludge generation, N_s (mg TKN-N/l)	36.21
Concentration of nitrates generated in the reactor, N_c (mg TKN-N/l)	162.30
Nitrification Oxygen demand FO_n (kg/d)	1513
Nitrification Oxygen demand FO_n (mg O/(l.h))	61.8
Mass of nitrifier organisms, $M(X_A)$ kg VSS	317.0
Conc. of nitrifier organisms, X_A mgVSS/l	59.1
Flux of N in waste VSS (kg/d)	73.9
% N removal	39.7

Denitrification	Tmin
	Settled
De Nitrification Potential, D_{pp}	
Minimum primary anoxic mass fraction, f_{x1min}	0.00
Nitrate load on primary anoxic zone, mg N-NO3/l influent	34.2
Denitrifcation potential of primary anoxic zone, Dp1	34.2
Denitrifcation potential of secondary anoxic zone, Dp3	14.2
Complete denitrification possible?	**No**
A recycle ratio that loads the anoxic reactor with nitrates to its denitrification potentail, a_{0pt}	
A	n/a
B	n/a
C	n/a
Actual nitrates in the effluent if a recycle > than a_{opt}, N_{nemin} (mgN/l)	52.64
Nitrates in the effluent at chosen a recycle, Nne (mgN/l)	52.64
Total TKN and nitrates in the effluent, Nte + Nne (mg/l)	353.55
Primary anoxic mass fraction at a_{opt}, fx1	
Primary anoxic mass fraction at chosen a recycle, fx1	
Primary anoxic mass fraction at chosen a recycle corrected, fx1,actual	
secondary anoxic mass fraction at a_{opt}, fx3	
secondary anoxic mass fraction at chosen a recycle, fx3	
secondary anoxic mass fraction at chosen a recycle, fx3, corrected	
% N removal	29.2
Total anoxic mass fraction selected	0.3600
Total anoxic mass fraction calculated (fx1+fx3)	0.0000
fx1 existing	0.19
fx3 existing	0.16

Phosphorus	Tmin
No BPR	**Settled**
Phosphorus incorporated in sludge generation, P_s (mg P/l)	10.86
Total soluble Phosphorus in the effluent, P_{te} (mg P/l)	175.10
Flux of P in waste VSS (kg/d)	22
% P removal (without BEPR)	5.8
With BEPR	
Assume the PAO get all RBCOD?	no
RBCOD available for PAO's, S'_{bsi} (mgCOD/l)	0.0
RBCOD concentration leaving the nth compartment in anaerobic zone, S_{bsfn} (mgCOD/l) - **Guess**	0.00
Active OHO Mass, MX_{BH} kgVSS	1681
RBCOD concentration leaving the nth compartment in anaerobic zone, S_{bsfn} (mgCOD/l) - **Calculated**	0.00
RBCOD concentration leaving the nth compartment in anaerobic zone, S_{bsfn} (mgCOD/l) -	0.0
Flux of influent COD available for PAO's, $FS_{b,PAO}$ (kg/d)	0
Flux of remaining influent COD available for OHO's, $FS_{b,OHO}$ (kg/d)	1096
PAO active mass, MX_{BG} (kgVSS)	0
PAO endogenous residue mass, MX_{EG} (kgVSS)	0
Concentration of Phosphorus removed in sludge, P_{rem} (mgP/l influent)	10.9
Total soluble Phosphorus in the effluent, P_{te} (mg P-N/l)	175.10
Fraction of polyphosphates in PAO, f_{XBGPP}	0.270
Active PAO concentration, X_{BG} (mg VSS/l)	0
Endogenous PAO concentration, X_{BE} (mgVSS/l)	0
PAO Carbonacous Oxygen Flux (kgO/h)	0
% P removal (with BEPR)	5.8

Reactor Volume	Tmin
	Settled
Anaerobic Zone	
Sludge mass fraction	0.09
Sludge mass (kg)	1932
Concentration TSS (kg/m3	4
Volume (m3)	483
Anoxic Zone 1	
Sludge mass fraction	0.19
Sludge mass (kg)	4079
Concentration TSS (kg/m3	4.0
Volume (m3)	1020
Aerobic Zone	
Sludge mass fraction, fx,a	0.56
Sludge mass, MXt,aero (kg	12021
Concentration TSS (kg/m3	4
Volume (m3)	3005
Anoxic Zone 2	
Sludge mass fraction	0.1600
Sludge mass, MXt,ano2 (k	3435
Concentration TSS (kg/m3	4.0
Volume (m3)	859
Total Process Volume, Vp (m3)	5366.585

Oxygen Demand	Tmax
	Settled
Total Carbonaceous Oxygen Flux FO_c (kg O/d)	838
Nitrification Oxygen demand FO_n (kg/d)	1522
DeNitrification Oxygen credit FO_n (kg/d)	643
Total Oxygen Flux FO_m (kgO/d)	1716

COD balance	
Mass of COD treated per day FS_{ti} (kg O/d)	6569
Total COD in	6569
Mass of COD leaving in WAS (kg O/d)	1911
Mass of COD leaving in the effluent (kg O/d)	3841
Mass of COD from Oxygen Demand (kg O/d)	817
Total COD out	6569
Balance	0

Nitrogen Balance (without nitrification)	
Mass of N treated per day FN_{ti} (kg N/d)	1019
Total N in	1019
Mass of N leaving in WAS (kg N/d)	239.6
Mass of N leaving in the effluent (kg N/d)	779.2
Mass of N leaving in effleunt as nitrates (kg N/d)	0
Total N out	1019
Balance	0

Nitrogen Balance (with nitrification)	
Mass of TKN treated per day FN_{ti} (kg N/d)	1019
Total N in	1019
Mass of N leaving in WAS (kg N/d)	239.6
Mass of N leaving in the effluent (kg N/d)	506.2
Mass of N leaving in effleunt as nitrates (kg N/d)	273
Total N out	1019
Balance	0

TOD Balance	
Mass of TOD in	11224.7171
Total N in	11225
TOD leaving in WAS	2740
TOD leaving in effluent	6154.3
Total Oxygen demand	1690.4
TOD leaving as Nitrogen gas	640
Total out	11225
Balance	0

Phosphorus Balance	
Mass of P treated per day FP_{ti} (kg P/d)	379
Total P in	379
Mass of P leaving in WAS (kg P/d)	85
Mass of P leaving in the effluent (kg P/d)	294.6
Total P out	379.4
Balance	0

Figure 8.2: Spreadsheet results of the 5-Stage Bardenpho model used to calculate the incremental increase in sludge production and aeration at the mainstream treatment process at Cape Flats WWTP treating sidestream effluent from the PN/A process.

Table 8.17: Calculation of the incremental increase in aeration due to the recycling of treated sidestream liquor back to the mainstream wastewater treatment works at Cape Flats WWTP from the PN/A process.

Fine Bubble Aeration Calculation	
Actual oxygen requirement (kg/h)	71.5
Diffuser submergence (est.)	4.2
OTE per meter	7
Oxygen Transfer Efficiency, OTE	29.40
Water temperature	18
Elevation (m)	0
Atmospheric pressure at elevation, Psite (kPa)	101
Atmospheric pressure at elevation, Psite (mmHg)	761
Water saturated vapour pressure, psite (mmHg)	15.5
DO saturation concentration at atmospheric pressure corrected for temperature, Cs (mg/l)	9.44
DO saturation corrected for water depth and temperature, $C_{s,corrected}$ (mg/l)	11.43
Pressure at aerator depth (mmHG gauge)	309
Assumed operating oxygen concentration (mg/l)	2
Beta factor to account for wastewater characteristics	0.95
Alpha factor to correct for aerator type, basin geometry, degree of mixing and wastewater	0.6
Fouling factor for diffusers	0.9
Combined AOR/SOR correction factor	0.503
Standard oxygen requirement, SOR (kg/h)	142
Adjust SOR for partial dissolution, i.e. incorporate OTE	484
Mass flow of air required, (kg/h)	2084
Volumetric air flow required at 20°C and 0% RH, (m3/h)	1737
Assumed FBDA equipment efficiency (kgO2/kWh)	5
Aeration energy requirement (shaft power)	28.4
Shaft to wire efficiency	70%
Total aeration energy (kW)	40.6

8.2 MLE process

8.2.1 MLE model inputs

Table 8.18: MLE model inputs

Sidestream liquor									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**	**TSS**	**ISS**	**VFA**
Sidestream liquor (unfiltered)	4500		1674	186			511	277	
Sidestream liquor (filtered)	4151		1655	177	1355	109			1868
Methanol									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**	**TSS**	**ISS**	**VFA**
Methanol (unfiltered)	1,185,000		0	0			0	0.0	
Methanol (filtered)	1,185,000		0	0	0	0			0
Sidestream liquor Uniodegradible solubles									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**			
Effluent	2283		300	68.5	0	0			

Total flow to reactor (Ml/d)	2.05
Reactor Temperature (°C)	24
Methanol flow rate (Ml/d)	0.0057
Methanol COD (mg/l)	1,185,000
Blended TKN/COD ratio	0.215

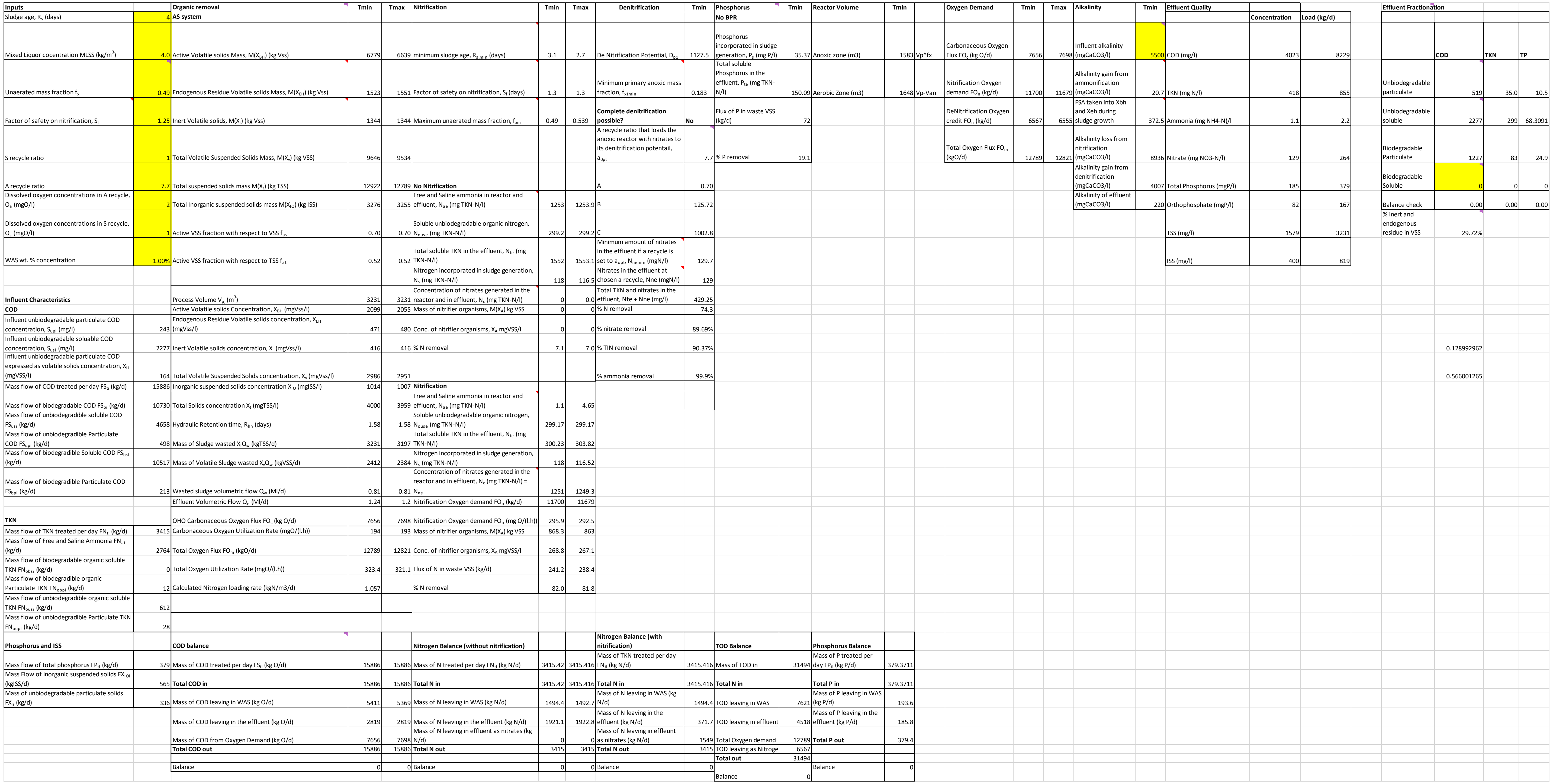
Table 8.19: Calculation results of the blended feed (sidestream + organic dosing) to the MLE process.

Blended Feed to MLE system									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**	**TSS**	**ISS**	**VFA**
Unfiltered	7766		1670	185			510	276	
Filtered (0,45um membrane)	7418		1650	177	1351	108			1863
Blended Unbiodegradable inputs									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**			
Effluent (raw)	2277		299	68	0	0			
Effluent (settled)									

Table 8.20: Fractionated organic components of the blended feed to the MLE process

Blended Feed Fractionation			
	COD	**TKN**	**TP**
Unbiodegradable particulate	243	14	6
Unbiodegradable soluble	2277	299	68
Biodegradable Particulate	104	6	3
Biodegradable Soluble	5141	0	29

8.2.2 MLE model raw outputs



Inputs	
Sludge age, R_s (days)	4
Mixed Liquor cocentration MLSS (kg/m^3)	4.0
Unaerated mass fraction f_x	0.49
Factor of safety on nitrification, S_f	1.25
S recycle ratio	1
A recycle ratio	7.7
Dissolved oxygen concentrations in A recycle, O_a (mgO/l)	2
Dissolved oxygen concentrations in S recycle, O_s (mgO/l)	1
WAS wt. % concentration	1.00%
Influent Characteristics	
COD	
Influent unbiodegradable particulate COD concentration, S_{upi} (mg/l)	243
Influent unbiodegradable soluable COD concentration, S_{usi} (mg/l)	2277
Influent unbiodegradable particulate COD expressed as volatile solids concentration, X_{ii} (mgVSS/l)	164
Mass flow of COD treated per day FS_{ti} (kg/d)	15886
Mass flow of biodegradable COD FS_{bi} (kg/d)	10730
Mass flow of unbiodegradible soluble COD FS_{usi} (kg/d)	4658
Mass flow of unbiodegradible Particulate COD FS_{upi} (kg/d)	498
Mass flow of biodegradible Soluble COD FS_{bsi} (kg/d)	10517
Mass flow of biodegradible Particulate COD FS_{bpi} (kg/d)	213
TKN	
Mass flow of TKN treated per day FN_{ti} (kg/d)	3415
Mass flow of Free and Saline Ammonia FN_{ai} (kg/d)	2764
Mass flow of biodegradable organic soluble TKN FN_{obsi} (kg/d)	0
Mass flow of biodegradible organic Particulate TKN FN_{obpi} (kg/d)	12
Mass flow of unbiodegradible organic soluble TKN FN_{ousi} (kg/d)	612
Mass flow of unbiodegradible Particulate TKN FN_{oupi} (kg/d)	28
Phosphorus and ISS	
Mass flow of total phosphorus FP_{ti} (kg/d)	379
Mass Flow of inorganic suspended solids FX_{IOi} (kgISS/d)	565
Mass of unbiodegradable particulate solids FX_{Ii} (kg/d)	336

Organic removal	Tmin	Tmax
AS system		
Active Volatile solids Mass, M(X_{BH}) (kg Vss)	6779	6639
Endogenous Residue Volatile solids Mass, M(X_{EH}) (kg Vss)	1523	1551
Inert Volatile solids, M(X_i) (kg Vss)	1344	1344
Total Volatile Suspended Solids Mass, M(X_v) (kg VSS)	9646	9534
Total suspended solids mass M(X_t) (kg TSS)	12922	12789
Total Inorganic suspended solids mass M(X_{IO}) (kg ISS)	3276	3255
Active VSS fraction with respect to VSS f_{av}	0.70	0.70
Active VSS fraction with respect to TSS f_{at}	0.52	0.52
Process Volume $V_{p,}$ (m^3)	3231	3231
Active Volatile solids Concentration, X_{BH} (mgVss/l)	2099	2055
Endogenous Residue Volatile solids concentration, X_{EH} (mgVss/l)	471	480
Inert Volatile solids concentration, X_i (mgVss/l)	416	416
Total Volatile Suspended Solids concentration, X_v (mgVss/l)	2986	2951
Inorganic suspended solids concentration X_{IO} (mgISS/l)	1014	1007
Total Solids concentration X_t (mgTSS/l)	4000	3959
Hydraulic Retention time, R_{hn} (days)	1.58	1.58
Mass of Sludge wasted X_tQ_w (kgTSS/d)	3231	3197
Mass of Volatile Sludge wasted X_vQ_w (kgVSS/d)	2412	2384
Wasted sludge volumetric flow Q_w (Ml/d)	0.81	0.81
Effluent Volumetric Flow Q_e (Ml/d)	1.24	1.2
OHO Carbonaceous Oxygen Flux FO_c (kg O/d)	7656	7698
Carbonaceous Oxygen Utilization Rate (mgO/(l.h))	194	193
Total Oxygen Flux FO_m (kgO/d)	12789	12821
Total Oxygen Utilization Rate (mgO/(l.h))	323.4	321.1
Calculated Nitrogen loading rate (kgN/m3/d)	1.057	

Nitrification	Tmin	Tmax
minimum sludge age, $R_{s,min}$ (days)	3.1	2.7
Factor of safety on nitrification, S_f (days)	1.3	1.3
Maximum unaerated mass fraction, f_{xm}	0.49	0.539
No Nitrification		
Free and Saline ammonia in reactor and effluent, N_{ae} (mg TKN-N/l)	1253	1253.9
Soluble unbiodegradable organic nitrogen, N_{ouse} (mg TKN-N/l)	299.2	299.2
Total soluble TKN in the effluent, N_{te} (mg TKN-N/l)	1552	1553.1
Nitrogen incorporated in sludge generation, N_s (mg TKN-N/l)	118	116.5
Concentration of nitrates generated in the reactor and in effluent, N_c (mg TKN-N/l)	0	0.0
Mass of nitrifier organisms, M(X_A) kg VSS	0	0
Conc. of nitrifier organisms, X_A mgVSS/l	0	0
% N removal	7.1	7.0
Nitrification		
Free and Saline ammonia in reactor and effluent, N_{ae} (mg TKN-N/l)	1.1	4.65
Soluble unbiodegradable organic nitrogen, N_{ouse} (mg TKN-N/l)	299.17	299.17
Total soluble TKN in the effluent, N_{te} (mg TKN-N/l)	300.23	303.82
Nitrogen incorporated in sludge generation, N_s (mg TKN-N/l)	118	116.52
Concentration of nitrates generated in the reactor and in effluent, N_c (mg TKN-N/l) = N_{ne}	1251	1249.3
Nitrification Oxygen demand FO_n (kg/d)	11700	11679
Nitrification Oxygen demand FO_n (mg O/(l.h))	295.9	292.5
Mass of nitrifier organisms, M(X_A) kg VSS	868.3	863
Conc. of nitrifier organisms, X_A mgVSS/l	268.8	267.1
Flux of N in waste VSS (kg/d)	241.2	238.4
% N removal	82.0	81.8

Denitrification	Tmin
De Nitrification Potential, D_{p1}	1127.5
Minimum primary anoxic mass fraction, f_{x1min}	0.183
Complete denitrification possible?	**No**
A recycle ratio that loads the anoxic reactor with nitrates to its denitrification potentail, a_{0pt}	7.7
A	0.70
B	125.72
C	1002.8
Minimum amount of nitrates in the effluent if a recycle is set to a_{opt}, N_{nemin} (mgN/l)	129.7
Nitrates in the effluent at chosen a recycle, Nne (mgN/l)	129
Total TKN and nitrates in the effluent, Nte + Nne (mg/l)	429.25
% N removal	74.3
% nitrate removal	89.69%
% TIN removal	90.37%
% ammonia removal	99.9%

Phosphorus	Tmin
No BPR	
Phosphorus incorporated in sludge generation, P_s (mg P/l)	35.37
Total soluble Phosphorus in the effluent, P_{te} (mg TKN-N/l)	150.09
Flux of P in waste VSS (kg/d)	72
% P removal	19.1

Reactor Volume	Tmin	
Anoxic zone (m3)	1583	Vp*fx
Aerobic Zone (m3)	1648	Vp-Van

Oxygen Demand	Tmin	Tmax
Carbonaceous Oxygen Flux FO_c (kg O/d)	7656	7698
Nitrification Oxygen demand FO_n (kg/d)	11700	11679
DeNitrification Oxygen credit FO_n (kg/d)	6567	6555
Total Oxygen Flux FO_m (kgO/d)	12789	12821

Alkalinity	Tmin
Influent alkalinity (mgCaCO3/l)	5500
Alkalinity gain from ammonification (mgCaCO3/l)	20.7
FSA taken into Xbh and Xeh during sludge growth	372.5
Alkalinity loss from nitrification (mgCaCO3/l)	8936
Alkalinity gain from denitrification (mgCaCO3/l)	4007
Alkalinity of effluent (mgCaCO3/l)	220

Effluent Quality	Concentration	Load (kg/d)
COD (mg/l)	4023	8229
TKN (mg N/l)	418	855
Ammonia (mg NH4-N)/l	1.1	2.2
Nitrate (mg NO3-N/l)	129	264
Total Phosphorus (mgP/l)	185	379
Orthophosphate (mgP/l)	82	167
TSS (mg/l)	1579	3231
ISS (mg/l)	400	819

Effluent Fractionation			
	COD	TKN	TP
Unbiodegradable particulate	519	35.0	10.5
Unbiodegradable soluble	2277	299	68.3091
Biodegradable Particulate	1227	83	24.9
Biodegradable Soluble	0	0	0
Balance check	0.00	0.00	0.00
% inert and endogenous residue in VSS	29.72%		
		0.128992962	
		0.566001265	

COD balance		
Mass of COD treated per day FS_{ti} (kg O/d)	15886	15886
Total COD in	15886	15886
Mass of COD leaving in WAS (kg O/d)	5411	5369
Mass of COD leaving in the effluent (kg O/d)	2819	2819
Mass of COD from Oxygen Demand (kg O/d)	7656	7698
Total COD out	15886	15886
Balance	0	0

Nitrogen Balance (without nitrification)		
Mass of N treated per day FN_{ti} (kg N/d)	3415.42	3415.416
Total N in	3415.42	3415.416
Mass of N leaving in WAS (kg N/d)	1494.4	1492.7
Mass of N leaving in the effluent (kg N/d)	1921.1	1922.8
Mass of N leaving in effluent as nitrates (kg N/d)	0	0
Total N out	3415	3415
Balance	0	0

Nitrogen Balance (with nitrification)	
Mass of TKN treated per day FN_{ti} (kg N/d)	3415.416
Total N in	3415.416
Mass of N leaving in WAS (kg N/d)	1494.4
Mass of N leaving in the effluent (kg N/d)	371.7
Mass of N leaving in effleunt as nitrates (kg N/d)	1549
Total N out	3415
Balance	0

TOD Balance	
Mass of TOD in	31494
Total N in	
TOD leaving in WAS	7621
TOD leaving in effluent	4518
Total Oxygen demand	12789
TOD leaving as Nitroge	6567
Total out	31494
Balance	0

Phosphorus Balance	
Mass of P treated per day FP_{ti} (kg P/d)	379.3711
Total P in	379.3711
Mass of P leaving in WAS (kg P/d)	193.6
Mass of P leaving in the effluent (kg P/d)	185.8
Total P out	379.4
Balance	0

Figure 8.3: Spreadsheet results from the MLE model biological processes.

Table 8.21: SST design inputs and results for the MLE process.

SST design		
PWWF factor	1	
PDWF	1	
Sludge DSVI	125	ml/g
Sludge SSVI	83.75	
V0/n	17.78	
n	0.389	
V0	6.91	
Flux rating	80%	
Xf	4	g/l
Design Equations		
$X_{R,crit}$	10.3	kg/m3
$R,_{crit}$	0.516	
$q_{R,crit}$	0.94	m/h
X_{Lmin}	5.14	kg/m3
J_{Lmax}	9.6	kg/m2 h
$q_{AmaxSHCII}$	1.46	m/h
$R_{min,PWWF}$	0.650	
q_{ASHCI}	1.44	m/h
From Diurnal data		
ADWF	85	m3/h
PWWF	85	m3/h
Settling tank aera required, A_{ST}	73	m2
Conventional settling tank sizing		
SST diameter	10	m
number of SSTs	1.0	
Total area	79	m2
$Q_{R,\ crit}$	73	m3/h
qa at PWWF	1.09	m/h
Rmin at PWWF	0.65	
Qr at PWWF	55	m3/h

8.2.3 MLE EQI outputs

Table 8.22: EQI calculation results for the MLE process

	Limit	Predicted	βx	PUx	Pux - EQI
COD sol. (mg/l)	75	2277	1	2277.0	2276.97
COD part. (mg/l)	75	1746	1	1745.9	1745.94
Ammonia (mg NH4-N)/l	10	1	7.5	1.1	1.06
Nitrate (mg NO3-N/l)	10	129	7.5	129.0	129.02
Phosphorus (mgP/l)	5	185	15	185.5	185.45
TSS (mg/l)	25	1579	3	1579.2	1579.21
EQI	**22036**	**kg pollutants/d**			

8.2.4 MLE - Aeration calculation

Table 8.23: Inputs and calculation results of the aeration requirements for the PN/A process.

Fine Bubble Aeration Calculation	
Actual oxygen requirement (kg/h)	534
Diffuser submergence (est.)	5.2
OTE per meter	7
Oxygen Transfer Efficiency, OTE	36.40
Water temperature	24
Elevation (m)	0
Atmospheric pressure at elevation, Psite (kPa)	101
Atmospheric pressure at elevation, Psite (mmHg)	761
Water saturated vapour pressure, psite (mmHg)	22.4
DO saturation concentration at atmospheric pressure corrected for temperature, Cs (mg/l)	8.42
DO saturation corrected for water depth and temperature, $C_{s,corrected}$ (mg/l)	10.54
Pressure at aerator depth (mmHG gauge)	382
Assumed operating oxygen concentration (mg/l)	2
Beta factor to account for wastewater characteristics	0.95
Alpha factor to correct for aerator type, basin geometry, degree of mixing and wastewater	0.6
Fouling factor for diffusers	0.9
Combined AOR/SOR correction factor	0.524
Standard oxygen requirement, SOR (kg/h)	1019
Adjust SOR for partial dissolution, i.e. incorporate OTE	2799
Mass flow of air required, (kg/h)	12065
Volumetric air flow required at 20°C and 0% RH, (m3/h)	10054

Assumed FBDA equipment efficiency (kgO2/kWh)	5
Aeration energy requirement (shaft power)	204
Shaft to wire efficiency	70%
Total aeration energy (kW)	291

8.2.5 MLE – Energy balance

Table 8.24: MLE energy balance inputs

Operating Conditions	**Winter**	**Winter low**	
Sidestream Feed Flow rate	85	85	m3/h
Sidestream Feed Temperature	25	25	°C
Dilution Feed flowrate	0.00	0.00	m3/h
Dilution water temperature	14	25	°C
Organic source flowrate (PS or glycerol)	0.24	0.24	m3/h
Organic source temperature (PS or glycerol)	14	10	°C
Blended temperature in equalisation tank	25	25	
Reactor Operating temperature	24	24	°C
Ambient Temperature	13	8.0	°C
Ground temperature	13	8.0	°C

Table 8.25: MLE major equipment dimensions and heat transfer areas

Balancing tank dimensions		
Design retention time of tank	12	hours
Volume of balancing tank working	1023	m3
Tank feeboard	30%	
Total tank volume	1461	m3
Tank Height	6	m
Tank Diameter	18	m
Tank Wall thickness	0.25	m
Surface area of walls	339	m2
Surface area through floor	254	m2
Surface area of roof	254	m2
Reactor Dimensions		
Number of reactors	1	
Reactor working volume	3231	m3
Reactor Freeboard	10%	
Reactor Total volume	3589	m3
Reactor Height	6	m
Reactor L/W ratio	2	
Reactor Length	35	m
Reactor Width	17	

Reactor Wall thickness	0.25	m
Total Surface area of walls	623	m2
Total Surface area through floor	598	m2
Total Surface area of roof	598	m2
SST dimensions		
number of SST's	1	
Sidewall height	3	
SST diameter	10	
SST Area	79	m2
SST Volume	236	m3
SST wall thickness	0.25	m
Surface area of walls	94	m2
Surface area through floor	79	m2
Surface area of roof	79	m2

Table 8.26: Calculation of overall heat transfer co-efficients for the MLE process

Overall heat transfter Coefficient		
Convective Heat Transfer Inside Wall, h_i	10000	W/m2 K
Concrete Wall thickness	0.25	m
Concrete Plinth thickness	0.5	m
Concrete conductivity	1.21	W/m K
Convective Heat Transfer Outside Wall, h_o	40.8	W/m2 K
Convective Heat Transfer tank head space, h_o	26.2	
$U_{\text{concrete walls, no insulation}}$	4.3	W/m2 K
$U_{\text{roof, closed concrete}}$	3.7	W/m2 K
$U_{\text{roof, open}}$	40.6	W/m2 K
$U_{\text{floor concrete}}$	1.6	W/m2 K

Table 8.27: Calculation of heating requirements of MLE process.

Heat losses to environment	**Winter ave.**	**Winter low**	
Balancing tank			
Walls (concrete)	17.4	24.9	kW
Floor (concrete)	4.9	7.0	kW
Roof (closed)	11.2	16.0	kW
Total	33.4	47.9	kW
Reactor			
Walls (concrete)	29.3	43.1	kW
Floor (concrete)	10.5	15.4	kW

Roof (open)	264.1	388.9	kW
Total	303.9	447.4	kW
SST			
Walls (concrete)	4.4	6.5	kW
Floor (concrete)	1.4	2.0	kW
Roof (open)	34.7	51.1	kW
Total	40.5	59.6	kW
Total heat loss through major equipment	377.8	554.9	kW
Sensible heat loss	-96.3	-95.2	kW
Total heating requirement	281	460	kW

8.2.6 Operational cost index calculations for MLE process

Table 8.28: OCI inputs and calculation results for the MLE process

MLE Aeration Energy	291	kW
Additional aeration energy to process effluent in Cape Flats WWTP	49.8	kW
Pumping Energy		
A recycle		
Volumetric flow rate	656	m3/h
Hydraulic Lift Height	1	m
Hydraulic power	1.8	kW
Hydraulic to wire efficiency	50%	
Total Power	3.6	
S recycle		
Volumetric flow rate	85	m3/h
Hydraulic Lift Height	10	m
Hydraulic power	2.3	kW
Hydraulic to wire efficiency	50%	
Total Power	4.65	kW
Feed Pump Power		
Volumetric flow rate	85	m3/h
Pump differential head	10	m
Hydraulic power	2.3	kW
Hydraulic to wire efficiency	50%	
Total Power	4.65	kW

Total pumping Engergy (kW)	12.87	kW
Mixing Energy		
Anoxic Zone Volume	1583	m3
Mixing Energy requirement	5	W/m3
Total Mixing Energy	7.91	kW
Electrical energy cost		
Total power consumption	361.7	kW
Electricty cost at base date	1.07	R/kWh
Electricity price increase 2019/20	5.80%	
Electricty price increase from 2020/21	6.90%	
Calculated electricty cost for 2020	1.21	R/kWh
Total electricity cost at base date	R3,834,320	
Heating energy		
Total heating demand	281	kW
Sludge Handling Cost		
Waste sludge produced	2005	kg TSS/d
Sludge disposal cost	R812.00	R per t
Sludge disposal cost per day	R1,628.44	
Sludge disposal cost per year	R594,382	
Chemical Costs		
methanol addition	5.7	m3/d
methanol density	0.792	
methanol cost	R10.60	R/kg
methanol cost per day	R47,507	
methanol cost per annum	R17,339,917	
OCI	R21,768,619	

8.2.7 Capital cost breakdown for MLE process

Item No.	Description	Qty	Budget Unit Cost	Section Budget Estimate
100.00	**MLE (Methanol organic source) Capital Cost Estimate**			
101.00	**Reactors**			
101.01	FBDA network in reactor	1	R 3,758,852	R 3,758,852
101.02	Anoxic recycle pumps	2	R 317,017	R 634,033.61
101.03	Sludge recylce pumps	2	R 100,000.00	R 200,000.00
101.04	Sludge piping and fittings	75	R 2,420.00	R 544,500.00
101.05	Valves on sludge piping	4	R 20,000.00	R 80,000.00
101.06	Treated water piping	100	R 2,420.00	R 726,000.00
101.08	Settling tank	1	R 349,670	R 349,669.60
101.09	Mixers	2	R 106,306	R 212,611.07
101.10	Tube in tube heat exchanger	1	R 1,686,190	R 1,686,190.16
101.11	Galvanized steel platforms and walkways	1	R 450,000.00	R 450,000.00
101.12	Installation, commissioning and upholding			R 1,728,371.30
				R 10,370,227.80
102.00	**Aeration system**			
102.01	Blowers	5	R 1,700,000	R 8,500,000.00
102.02	Aeration piping main header	40	R 10,337.12	R 1,033,712.00
102.03	Aeration piping sub headers	20	R 10,337.12	R 620,227.20
102.04	Control valves	3	R 100,000.00	R 300,000.00
102.05	Subheader isolation valves	3	R 50,000.00	R 150,000.00
102.06	Blower Isolation valves	5	R 40,000.00	R 200,000.00
102.07	Blower NRV's	5	R 40,000.00	R 200,000.00
102.08	Galvanized steel pipe bridge	1	R 175,000.00	R 175,000.00
102.09	Installation, commissioning and upholding			R 2,235,787.84
				R 13,414,727.04
103.00	**Utilities**			
103.01	Utility equipment and installation	1	R 1,856,121	R 1,856,121.02
103.02	Installation, commissioning and upholding	1		
				R 1,856,121.02
104.00	**HVAC**			
104.01	HVAC equipment and installation		R 5,714,922	R 5,714,922.10
				R 5,714,922.10

105.00	**Chemical Dosing (Methanol)**			
105.01	Storage tank	2	R 600,000.00	R 1,200,000.00
105.02	Dosing pumps	2	R 112,500.00	R 225,000.00
105.03	SS piping	80	R 450.00	R 108,000.00
105.04	Valves	sum	R 10,000.00	R 10,000.00
105.05	Installation and commissioning			R 308,600.00
				R 1,851,600.00
106.00	**Buffer Tank**			
106.01	Vertical shaft mixers	4	R 150,000.00	R 600,000.00
106.02	Lifting davits/ guide rails	3	R 45,000.00	R 135,000.00
106.03	Galvanized steel platforms and walkways	1	R 300,000.00	R 300,000.00
106.04	Feed pumps	4	R 100,000.00	R 400,000.00
106.05	Feed piping to tank	130	R 3,000.00	R 1,170,000.00
106.06	Discharge piping to reactor	120	R 2,000.00	R 720,000.00
106.07	Overflow/drain piping	30	R 5,200.00	R 312,000.00
106.08	Isolation valves	20	R 45,000.00	R 900,000.00
	Installation, commissioning and upholding			R 907,400.00
				R 5,444,400.00
107.00	**Seed Sludge (Not applicable)**			
				R 0.00
108.00	**Electrical and instrumentation**			
	Total			**R 14,521,420.42**
109.00	**Civil Costs**			
109.01	Site reshaping	sum	R3,090,288	R 3,090,288.46
109.02	Reactor	sum	R4,737,690	R 4,737,689.70
109.03	SST's	sum	R815,385	R 815,384.62
109.04	Anammox reactor buffer tank	sum	R3,907,692	R 3,907,692.31
109.05	Methanol dosing tank ground slab	sum	R112,308	R 112,307.69
109.06	MCC and blower house	sum	R5,735,769	R 5,735,769.23
109.07	Road and offloading bay	sum	R3,626,154	R 3,626,153.85
109.08	Underground drainage	sum	R824,615	R 824,615.38
109.09	Application, approvals and permits	sum	R76,923	R 76,923.08
109.10	Additional civil items (stairs, ladders, piping support, etc)	sum	R384,615	R 384,615.38
				R 23,311,439.70

110.00	**P&G's**	
110.01	Mechanical P&Gs	R 7,730,400
110.02	Electrical P&G's	R 2,904,284
110.03	Civil	R 3,496,716
		R 14,131,399.63
111.00	**Extras**	
111.01	Health and Safety	R 500,000.00
111.02	Environment	R 250,000.00
111.03	Sundries	R 750,000.00
		R 1,500,000.00
	SUB- TOTAL (equipment & contracts)	R 92,116,257.72
	PROJECT CONTINGENCIES 15 % of COW	R 13,817,438.66
	PROJECT COST OF WORKS (excl. VAT)	R 105,933,696
	PROJECT FEE CATEGORY	F (water and wastewater treatment works)
	PROJECT FEE	13.00%
	PROFESSIONAL FEE % ADJUSTED FOR COW	10.84%
	PROJECT FEE	R11,485,778
	TOTAL PROJECT COST (excl. VAT)	R 117,419,474
	VAT @15%	R 15,890,054
	TOTAL PROJECT COSTS (incl. VAT)	R 133,309,529

8.2.8 Lifecycle Cost Analysis calculation results for MLE process

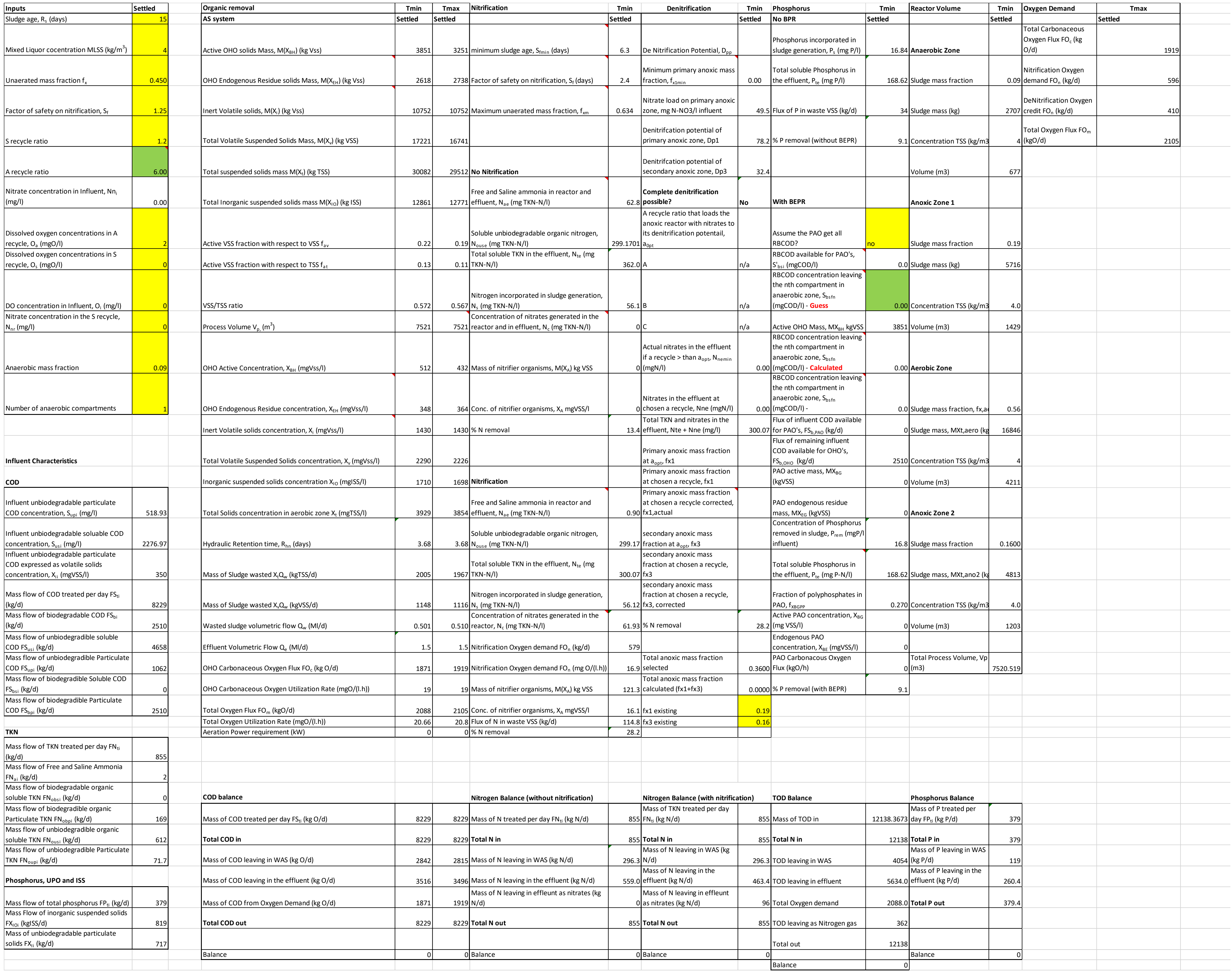
Table 8.29: LCCA calculation results for the MLE process

		CAPEX			OPEX									Net Present Value	Cumulative net present value
Year	n				Energy			Sludge disposal cost			Chemical Cost				
		Cost						Cost			Cost				
2020	0	R39,139,825	1.00	1.00	R3,834,320	1.00	1.00	R594,381.71	1.00	1.00	R17,339,917.21	1.00	1.00	R39,139,825	R39,139,825
2021	1	R41,175,096	0.98	1.98	R4,225,420	1.03	2.03		0.98	1.98		0.99	1.99	R38,481,398	R77,621,223
2022	2	R43,316,201	0.97	2.95	R4,656,413	1.06	3.09		0.97	2.95		0.98	2.98	R37,834,047	R115,455,270
2023	3		0.95	3.90	R5,131,367	1.09	4.18		0.95	3.90		0.98	3.95	R21,702,707	R137,157,977
2024	4		0.93	4.83	R5,654,767	1.13	5.31		0.93	4.83		0.97	4.92	R21,690,169	R158,848,146
2025	5		0.92	5.75	R6,231,553	1.16	6.47		0.92	5.75		0.96	5.89	R21,682,507	R180,530,653
2026	6		0.90	6.66	R6,867,172	1.19	7.66		0.90	6.66		0.96	6.84	R21,679,826	R202,210,479
2027	7		0.89	7.54	R7,567,623	1.23	8.89		0.89	7.54		0.95	7.79	R21,682,229	R223,892,708
2028	8		0.87	8.42	R8,339,521	1.27	10.16		0.87	8.42		0.94	8.73	R21,689,827	R245,582,535
2029	9		0.86	9.28	R9,190,152	1.30	11.46		0.86	9.28		0.93	9.67	R21,702,731	R267,285,265
2030	10		0.84	10.12	R10,127,547	1.34	12.80		0.84	10.12		0.93	10.59	R21,721,058	R289,006,323
2031	11		0.83	10.95	R11,160,557	1.38	14.18		0.83	10.95		0.92	11.51	R21,744,928	R310,751,251
2032	12		0.82	11.77	R12,298,934	1.42	15.61		0.82	11.77		0.91	12.43	R21,774,466	R332,525,717
2033	13		0.80	12.57	R13,553,425	1.47	17.08		0.80	12.57		0.91	13.33	R21,809,800	R354,335,517
2034	14		0.79	13.36	R14,935,874	1.51	18.59		0.79	13.36		0.90	14.23	R21,851,063	R376,186,580
2035	15		0.78	14.13	R16,459,334	1.56	20.14		0.78	14.13		0.89	15.12	R21,898,391	R398,084,970
2036	16		0.76	14.89	R18,138,186	1.60	21.74		0.76	14.89		0.89	16.01	R21,951,925	R420,036,895
2037	17		0.75	15.64	R19,988,281	1.65	23.39		0.75	15.64		0.88	16.89	R22,011,813	R442,048,708
2038	18		0.74	16.38	R22,027,085	1.70	25.09		0.74	16.38		0.87	17.76	R22,078,203	R464,126,912
2039	19		0.72	17.10	R24,273,848	1.75	26.84		0.72	17.10		0.87	18.62	R22,151,253	R486,278,165
2040	20		0.71	17.82	R26,749,780	1.80	28.65		0.71	17.82		0.86	19.48	R22,231,122	R508,509,287
2041	21		0.70	18.52	R29,478,258	1.86	30.50		0.70	18.52		0.85	20.34	R22,317,977	R530,827,264
2042	22		0.69	19.21	R32,485,040	1.91	32.42		0.69	19.21		0.85	21.18	R22,411,987	R553,239,250
Present Value, PV		**R115,455,270**			**R112,445,655**			**R9,662,091**			**R315,676,235**				
Total NPV		**R553,239,250**													

8.2.9 MLE Incremental effect on mainstream works

Table 8.30:Treated sidestream liquor effluent inputs into the 5-Stage bardenpho model used to calculate the incremental increase in sludge production and aeration at the mainstream works.

WW influent inputs									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**	**TSS**	**ISS**	**VFA**
Raw sewage	4023		418	185			1579	400	
0,45um membrane	2277		300	150	1.06	82			0
WW effluent inputs									
	COD	**TOC**	**TKN**	**TP**	**FSA**	**OP**			
Effluent (raw)	2277.0		299.2	68.31					
Effluent (settled)	2277.0		299.2	68.31					



Inputs	**Settled**
Sludge age, R_s (days)	15
Mixed Liquor cocentration MLSS (kg/m^3)	4
Unaerated mass fraction f_x	0.450
Factor of safety on nitrification, S_f	1.25
S recycle ratio	1.2
A recycle ratio	6.00
Nitrate concentration in Influent, Nn_i (mg/l)	0.00
Dissolved oxygen concentrations in A recycle, O_a (mgO/l)	2
Dissolved oxygen concentrations in S recycle, O_s (mgO/l)	0
DO concentration in Influent, O_i (mg/l)	0
Nitrate concentration in the S recycle, N_{nr} (mg/l)	0
Anaerobic mass fraction	0.09
Number of anaerobic compartments	1
Influent Characteristics	
COD	
Influent unbiodegradable particulate COD concentration, S_{upi} (mg/l)	518.93
Influent unbiodegradable soluable COD concentration, S_{usi} (mg/l)	2276.97
Influent unbiodegradable particulate COD expressed as volatile solids concentration, X_{ii} (mgVSS/l)	350
Mass flow of COD treated per day FS_{ti} (kg/d)	8229
Mass flow of biodegradable COD FS_{bi} (kg/d)	2510
Mass flow of unbiodegradible soluble COD FS_{usi} (kg/d)	4658
Mass flow of unbiodegradible Particulate COD FS_{upi} (kg/d)	1062
Mass flow of biodegradible Soluble COD FS_{bsi} (kg/d)	0
Mass flow of biodegradible Particulate COD FS_{bpi} (kg/d)	2510
TKN	
Mass flow of TKN treated per day FN_{ti} (kg/d)	855
Mass flow of Free and Saline Ammonia FN_{ai} (kg/d)	2
Mass flow of biodegradable organic soluble TKN FN_{obsi} (kg/d)	0
Mass flow of biodegradible organic Particulate TKN FN_{obpi} (kg/d)	169
Mass flow of unbiodegradible organic soluble TKN FN_{ousi} (kg/d)	612
Mass flow of unbiodegradible Particulate TKN FN_{oupi} (kg/d)	71.7
Phosphorus, UPO and ISS	
Mass flow of total phosphorus FP_{ti} (kg/d)	379
Mass Flow of inorganic suspended solids FX_{IOi} (kgISS/d)	819
Mass of unbiodegradable particulate solids FX_{Ii} (kg/d)	717

Organic removal	**Tmin**	**Tmax**
AS system	**Settled**	**Settled**
Active OHO solids Mass, $M(X_{BH})$ (kg Vss)	3851	3251
OHO Endogenous Residue solids Mass, $M(X_{EH})$ (kg Vss)	2618	2738
Inert Volatile solids, $M(X_i)$ (kg Vss)	10752	10752
Total Volatile Suspended Solids Mass, $M(X_v)$ (kg VSS)	17221	16741
Total suspended solids mass $M(X_t)$ (kg TSS)	30082	29512
Total Inorganic suspended solids mass $M(X_{IO})$ (kg ISS)	12861	12771
Active VSS fraction with respect to VSS f_{av}	0.22	0.19
Active VSS fraction with respect to TSS f_{at}	0.13	0.11
VSS/TSS ratio	0.572	0.567
Process Volume V_p, (m^3)	7521	7521
OHO Active Concentration, X_{BH} (mgVss/l)	512	432
OHO Endogenous Residue concentration, X_{EH} (mgVss/l)	348	364
Inert Volatile solids concentration, X_i (mgVss/l)	1430	1430
Total Volatile Suspended Solids concentration, X_v (mgVss/l)	2290	2226
Inorganic suspended solids concentration X_{IO} (mgISS/l)	1710	1698
Total Solids concentration in aerobic zone X_t (mgTSS/l)	3929	3854
Hydraulic Retention time, R_{hn} (days)	3.68	3.68
Mass of Sludge wasted X_tQ_w (kgTSS/d)	2005	1967
Mass of Sludge wasted X_vQ_w (kgVSS/d)	1148	1116
Wasted sludge volumetric flow Q_w (Ml/d)	0.501	0.510
Effluent Volumetric Flow Q_e (Ml/d)	1.5	1.5
OHO Carbonaceous Oxygen Flux FO_c (kg O/d)	1871	1919
OHO Carbonaceous Oxygen Utilization Rate (mgO/(l.h))	19	19
Total Oxygen Flux FO_m (kgO/d)	2088	2105
Total Oxygen Utilization Rate (mgO/(l.h))	20.66	20.8
Aeration Power requirement (kW)	0	0

Nitrification	**Tmin**
	Settled
minimum sludge age, S_{fmin} (days)	6.3
Factor of safety on nitrification, S_f (days)	2.4
Maximum unaerated mass fraction, f_{xm}	0.634
No Nitrification	
Free and Saline ammonia in reactor and effluent, N_{ae} (mg TKN-N/l)	62.8
Soluble unbiodegradable organic nitrogen, N_{ouse} (mg TKN-N/l)	299.1701
Total soluble TKN in the effluent, N_{te} (mg TKN-N/l)	362.0
Nitrogen incorporated in sludge generation, N_s (mg TKN-N/l)	56.1
Concentration of nitrates generated in the reactor and in effluent, N_c (mg TKN-N/l)	0
Mass of nitrifier organisms, $M(X_A)$ kg VSS	0
Conc. of nitrifier organisms, X_A mgVSS/l	0
% N removal	13.4
Nitrification	
Free and Saline ammonia in reactor and effluent, N_{ae} (mg TKN-N/l)	0.90
Soluble unbiodegradable organic nitrogen, N_{ouse} (mg TKN-N/l)	299.17
Total soluble TKN in the effluent, N_{te} (mg TKN-N/l)	300.07
Nitrogen incorporated in sludge generation, N_s (mg TKN-N/l)	56.12
Concentration of nitrates generated in the reactor, N_c (mg TKN-N/l)	61.93
Nitrification Oxygen demand FO_n (kg/d)	579
Nitrification Oxygen demand FO_n (mg O/(l.h))	16.9
Mass of nitrifier organisms, $M(X_A)$ kg VSS	121.3
Conc. of nitrifier organisms, X_A mgVSS/l	16.1
Flux of N in waste VSS (kg/d)	114.8
% N removal	28.2

Denitrification	**Tmin**
	Settled
De Nitrification Potential, D_{pp}	
Minimum primary anoxic mass fraction, f_{x1min}	0.00
Nitrate load on primary anoxic zone, mg N-NO3/l influent	49.5
Denitrifcation potential of primary anoxic zone, Dp1	78.2
Denitrifcation potential of secondary anoxic zone, Dp3	32.4
Complete denitrification possible?	**No**
A recycle ratio that loads the anoxic reactor with nitrates to its denitrification potentail, a_{0pt}	
A	n/a
B	n/a
C	n/a
Actual nitrates in the effluent if a recycle > than a_{opt}, N_{nemin} (mgN/l)	0.00
Nitrates in the effluent at chosen a recycle, Nne (mgN/l)	0.00
Total TKN and nitrates in the effluent, Nte + Nne (mg/l)	300.07
Primary anoxic mass fraction at a_{opt}, fx1	
Primary anoxic mass fraction at chosen a recycle, fx1	
Primary anoxic mass fraction at chosen a recycle corrected, fx1,actual	
secondary anoxic mass fraction at a_{opt}, fx3	
secondary anoxic mass fraction at chosen a recycle, fx3	
secondary anoxic mass fraction at chosen a recycle, fx3, corrected	
% N removal	28.2
Total anoxic mass fraction selected	0.3600
Total anoxic mass fraction calculated (fx1+fx3)	0.0000
fx1 existing	0.19
fx3 existing	0.16

Phosphorus	**Tmin**
No BPR	**Settled**
Phosphorus incorporated in sludge generation, P_s (mg P/l)	16.84
Total soluble Phosphorus in the effluent, P_{te} (mg P/l)	168.62
Flux of P in waste VSS (kg/d)	34
% P removal (without BEPR)	9.1
With BEPR	
Assume the PAO get all RBCOD?	no
RBCOD available for PAO's, S'_{bsi} (mgCOD/l)	0.0
RBCOD concentration leaving the nth compartment in anaerobic zone, S_{bsfn} (mgCOD/l) - **Guess**	0.00
Active OHO Mass, MX_{BH} kgVSS	3851
RBCOD concentration leaving the nth compartment in anaerobic zone, S_{bsfn} (mgCOD/l) - **Calculated**	0.00
RBCOD concentration leaving the nth compartment in anaerobic zone, S_{bsfn} (mgCOD/l) -	0.0
Flux of influent COD available for PAO's, $FS_{b,PAO}$ (kg/d)	0
Flux of remaining influent COD available for OHO's, $FS_{b,OHO}$ (kg/d)	2510
PAO active mass, MX_{BG} (kgVSS)	0
PAO endogenous residue mass, MX_{EG} (kgVSS)	0
Concentration of Phosphorus removed in sludge, P_{rem} (mgP/l influent)	16.8
Total soluble Phosphorus in the effluent, P_{te} (mg P-N/l)	168.62
Fraction of polyphosphates in PAO, f_{XBGPP}	0.270
Active PAO concentration, X_{BG} (mg VSS/l)	0
Endogenous PAO concentration, X_{BE} (mgVSS/l)	0
PAO Carbonacous Oxygen Flux (kgO/h)	0
% P removal (with BEPR)	9.1

Reactor Volume	**Tmin**
	Settled
Anaerobic Zone	
Sludge mass fraction	0.09
Sludge mass (kg)	2707
Concentration TSS (kg/m3	4
Volume (m3)	677
Anoxic Zone 1	
Sludge mass fraction	0.19
Sludge mass (kg)	5716
Concentration TSS (kg/m3	4.0
Volume (m3)	1429
Aerobic Zone	
Sludge mass fraction, fx,ae	0.56
Sludge mass, MXt,aero (kg	16846
Concentration TSS (kg/m3	4
Volume (m3)	4211
Anoxic Zone 2	
Sludge mass fraction	0.1600
Sludge mass, MXt,ano2 (kg	4813
Concentration TSS (kg/m3	4.0
Volume (m3)	1203
Total Process Volume, Vp (m3)	7520.519

Oxygen Demand	**Tmax**
	Settled
Total Carbonaceous Oxygen Flux FO_c (kg O/d)	1919
Nitrification Oxygen demand FO_n (kg/d)	596
DeNitrification Oxygen credit FO_n (kg/d)	410
Total Oxygen Flux FO_m (kgO/d)	2105

COD balance		
Mass of COD treated per day FS_{ti} (kg O/d)	8229	8229
Total COD in	8229	8229
Mass of COD leaving in WAS (kg O/d)	2842	2815
Mass of COD leaving in the effluent (kg O/d)	3516	3496
Mass of COD from Oxygen Demand (kg O/d)	1871	1919
Total COD out	8229	8229
Balance	0	0

Nitrogen Balance (without nitrification)	
Mass of N treated per day FN_{ti} (kg N/d)	855
Total N in	855
Mass of N leaving in WAS (kg N/d)	296.3
Mass of N leaving in the effluent (kg N/d)	559.0
Mass of N leaving in effleunt as nitrates (kg N/d)	0
Total N out	855
Balance	0

Nitrogen Balance (with nitrification)	
Mass of TKN treated per day FN_{ti} (kg N/d)	855
Total N in	855
Mass of N leaving in WAS (kg N/d)	296.3
Mass of N leaving in the effluent (kg N/d)	463.4
Mass of N leaving in effleunt as nitrates (kg N/d)	96
Total N out	855
Balance	0

TOD Balance	
Mass of TOD in	12138.3673
Total N in	12138
TOD leaving in WAS	4054
TOD leaving in effluent	5634.0
Total Oxygen demand	2088.0
TOD leaving as Nitrogen gas	362
Total out	12138
Balance	0

Phosphorus Balance	
Mass of P treated per day FP_{ti} (kg P/d)	379
Total P in	379
Mass of P leaving in WAS (kg P/d)	119
Mass of P leaving in the effluent (kg P/d)	260.4
Total P out	379.4
Balance	0

Figure 8.4: Spreadsheet Results of the 5-Stage Bardenpho model used to calculate the incremental increase in sludge production and aeration at the mainstream treatment process at Cape Flats WWTP treating sidestream effluent from the MLE process.

Table 8.31: Calculation of the incremental increase in aeration due to the recycling of treated sidestream liquor back to the mainstream wastewater treatment works at Cape Flats WWTP from the MLE process

Fine Bubble Aeration Calculation	
Actual oxygen requirement (kg/h)	87.7
Diffuser submergence (est.)	4.2
OTE per meter	7
Oxygen Transfer Efficiency, OTE	29.40
Water temperature	18
Elevation (m)	0
Atmospheric pressure at elevation, Psite (kPa)	101
Atmospheric pressure at elevation, Psite (mmHg)	761
Water saturated vapour pressure, psite (mmHg)	15.5
DO saturation concentration at atmospheric pressure corrected for temperature, Cs (mg/l)	9.44
DO saturation corrected for water depth and temperature, $C_{s,corrected}$ (mg/l)	11.43
Pressure at aerator depth (mmHG gauge)	309
Assumed operating oxygen concentration (mg/l)	2
Beta factor to account for wastewater characteristics	0.95
Alpha factor to correct for aerator type, basin geometry, degree of mixing and wastewater	0.6
Fouling factor for diffusers	0.9
Combined AOR/SOR correction factor	0.503
Standard oxygen requirement, SOR (kg/h)	174
Adjust SOR for partial dissolution, i.e. incorporate OTE	593
Mass flow of air required, (kg/h)	2556
Volumetric air flow required at 20°C and 0% RH, (m3/h)	2130
Assumed FBDA equipment efficiency (kgO2/kWh)	5
Aeration energy requirement (shaft power)	34.9
Shaft to wire efficiency	70%
Total aeration energy (kW)	49.8